



PETROCHEMICAL INDUSTRIES
DESIGN & ENGINEERING CO.
PROJ. NO. : 1187



Mitsui Engineering &
Shipbuilding co., Ltd
Proj. NO.: HC3700

KHARG MEG PLANT

COVER SHEET



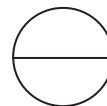
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DEVELOPMENT MANAGEMENT
COMPANY (PIDMCO)
PROJ. NO. : 23-33 KHARG MEG PLANT

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KHARG MEG PLANT OPERATING MANUAL

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1 BASIS OF DESIGN

1.1 Capacity

The plant design capacity is 400 kta EOE and with a water:EO ratio chosen such that 500 kta MEG is produced with balancing DEG and TEG production. The annual capacity is based on 7920 on stream production hours per year. The plant is designed to run on both high activity catalyst such as Shell/CRI S-863 catalyst and high selectivity catalyst such as Shell/CRI S-882 and S-886 catalyst. The design case is based on production of MEG as main product and DEG and TEG as by-products. In addition, the CO₂ formed in the EO section is considered a by-product. Besides being dependent on overall plant through-put, production amounts of CO₂ are largely dependent on type and age of the catalyst.

The plant can run at a turn down of 40%, although the efficiency will be lower than under full load. When operating with high selectivity catalyst and 50% turn down the thermosyphon circulation over the EO reactors may have to be supported by forced circulation using the condensate circulation pump. Reflux flow rates must be kept high enough to avoid separation problems due to insufficient wetting.

In case of a shut-down of either the EO section or the EG section the other section will be able to continue running for maximum 30 minutes under full load if the normal liquid level is present, since no intermediate buffer volume is available. In case of an unscheduled shut-down of the EG section the aqueous EO coming from the EO section will have to be temporarily stored in the EO stripper tops surge drum and the LE column bottom. In case the EG section would not be restarted in a short period of time the aqueous EO will need to be kept cool using the aqueous EO shutdown cooler to prevent the reactive mixture from running away.

Below table provides a summary of the design cases.

Product		Design
MEG	kta	500
DEG	kta	50
TEG	kta	3.4
CO ₂ (dry basis – HA catalyst)	kta	180 - 243
CO ₂ (dry basis – HS catalyst)	kta	109 - 195

The plant is capable to operate at 140% of the normal glycol bleed formation.



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1.2 Products

MEG

Fibre Grade Mono Ethylene Glycol (MEG)		
Appearance		Colourless, clear
Colour (Pt/Co)	max	5
Purity	% wt min	99.8
DEG	% wt max	0.08
Water	% wt max	0.08
Aldehydes (as Formaldehyde)	mg/kg max	8
Acidity (as Acetic Acid)	mg/kg max	10
Iron (as Fe)	mg/kg max	0.1
Inorganic Chlorides (as Cl)	mg/kg max	0.05
Ash	mg/kg max	50
Specific Gravity, 20/20°C		1.1151-1.1156
UV transmittance		
At 220 nm	% min	80
At 275 nm	% min	95
At 350 nm	% min	99
Distillation Range (0.1013 Mpa)		
5 v/v%	°C min	196
95 v/v%	°C max	199

DEG

Di Ethylene Glycol (DEG)		
Appearance		Colourless, clear
Colour (Pt/Co)	max	10
Purity	% wt min	99.8
MEG	% wt max	0.05
TEG	% wt max	0.05
Water	% wt max	0.05
Acidity (as Acetic Acid)	mg/kg max	50
Ash	mg/kg max	50
Specific Gravity, 20/20°C		1.1175 – 1.1195
Distillation Range (0.1013 Mpa)		
5 v/v%	°C min	242
95 v/v%	°C max	250



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TEG

Tri Ethylene Glycol (TEG)

Appearance		Colourless, clear
Colour (Pt/Co)	max	25
Purity	% wt min	99.5
DEG	% wt max	0.35
PEG	% wt max	0.1
Water	% wt max	0.05
Ash	mg/kg max	100
Specific Gravity, 20/20°C		1.124-1.126
Distillation Range (0.1013 Mpa)		
5 v/v%	°C min	280
95 v/v%	°C max	295

CO₂

Carbon Dioxide

Organic chlorides	ppm vol max	10
Sulphur	mg/kg max	0
EO	mg/m ³ max	20
Hydrocarbons (mainly CH ₄ , C ₂ H ₄)	%vol max	0.5
Aldehydes	ppm vol max	300
Glycols	ppm vol max	5
Battery limit condition		
Pressure	barg min	0.1
Temperature	°C max	65



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1.3 Feedstocks

Ethylene

Ethylene		
Ethylene	% mol min	99.95
Methane and Ethane	ppm mole max	500
Propylene and heavier hydrocarbons	ppm mol max	10
Acetylene	ppm mol max	1
Hydrogen	ppm mol max	5
Carbon Monoxide	ppm mol max	0.1
Carbon Dioxide	ppm mol max	0.1
Water	ppm mol max	0.1
Methanol	mg/kg max	0.5
Oxygenated compounds	mg/kg max	0.5
Oxygen	ppm mol max	0.1
Total combined Sulphur	mg/kg max	0.5
Sulphur after guard bed	mg/kg max	0.01
Pressure	bara min	29
Temperature	°C	Ambient
Total combined Nitrogen	ppm mol max	0.2
COS	ppm mol max	0.02
Mercaptans	ppm mol max	0.3

Oxygen

Oxygen		
Oxygen	% mol min.	99.8
Nitrogen + Argon	ppm mole	2000
Trichloroethylene + Trichloroethane	ppm mol max	0.001
Pressure	bara min /Nor	26 / 27
Temperature	°C	Ambient

1.4 Process materials

Catalyst

The design is based on performance predictions for typical S-863 high activity catalyst and typical S-882 high selectivity catalyst.



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Methane

Methane (Notes 1, 2)		
Methane	% mol min	93.0
Hydrogen	% mol max	3.0
CO + CO ₂	% mol max	0.6
Ethane	% mol max	0.05
Nitrogen	% mol max	0.84
Propylene + heavier hydrocarbons	ppm mol max	15
Acetylene	ppm mol max	5
Total Sulphur	mg/kg max	2.0
Sulphur (after guard bed)	mg/kg max	0.01
Methanol and other oxygenated organics	mg/kg max	10
Pressure	bara min	26
Temperature	°C	Ambient

(1) The methane specification design basis for the development of mass balances and flowsheet simulations has been based on 99.11%mol of methane CH₄, 0.05%mol of ethane C₂H₆ and 0.84%mol of nitrogen N₂.

(2) Methane purification package (PK-1001) outlet

Moderator

Ethylchloride		
Appearance		clear and free from foreign material
Purity	% wt min	99.8
Acidity (as HCl)	mg/kg max	10
Water	mg/kg max	100
Non-volatile residue	mg/100 ml	5
	max	
Trichloroethylene + Trichloroethane	mg/kg	50

Caustic

Rayon grade 50wt% caustic solution (0.02 wt% Cl max.). Dilution to 20wt% solution inside battery limits upstream of caustic tank TK-2002.

Potassium carbonate

Technical low-chloride grade (0.02 wt% Cl max.)



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Antifoam

Antifoam is used for the EO absorbent and carbonate systems. BASF polyether glycol, Pluronic L101, is recommended. An equivalent viscous liquid EO or PO polyether glycol with an average MW of 3500 to 4000, free from material under 1000 g/mole can be used. Pour point should be above 0°C. Silicon containing antifoam should never be used, because it has a permanent negative effect on the catalyst performance.

Activated carbon for sulphur guard beds

Refer to the data sheet.

Resin for MEG polishing

Refer to the data sheet.

BFW Treatment Chemicals

After vendor data will be specified.

Chilled Water conditioner

After vendor data will be specified.

1.5

Utilities

For complete specification of all utilities refer to "Ambient, site condition & utility data", ESS-PR-101.

Steam

Import HP steam is required for start-up to heat up the EO and glycol reactor. During normal operation, import of HP steam is required to supplement the generated steam of the EO reactor.

Boiler feed water

Import of demineralised water is required. The demineralised water shall meet all specifications for boiler feed water with the exception of pH and oxygen content, which will be controlled inside the battery limits. The specification of high purity boiler feed water is specified in Process Memorandum 107. The piping materials are specified in Process Memorandum 117.

Cooling water

Inlet temperature 38°C

Outlet temperature 48°C

Note that the provision of cooling water at a temperature of 38°C is critical to the proper performance of the process. Above specifications were provided as the basis for the process specifications.



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Mitsui Engineering &
Shipbuilding co., Ltd
Proj. NO.: HC3700

KHARG MEG PLANT

OPERATING MANUAL

DOC. NO. : MNL-PR-001

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If during the summer the cooling water temperature would exceed 38°C, this would have major consequences for the process. In particular the lean absorbent cooling would be affected and also the cooling of the LE column feed might suffer, as well as the condensation of the dehydrator overhead and the glycol bleed flasher overhead.

Nitrogen

Purity min 99.9 % mol
Hydrocarbons zero

Instrument air, electrical power, firewater, plant air, potable water, service steam and all other utilities have to be provided under applicable codes and internationally acceptable quality standards for quality and reliability. There are no specific other requirements from the process.

1.6

Yields

The plant is designed for operation on both high activity catalyst and high selectivity catalyst. The design base catalyst selectivities as well as plant yields for both catalysts are shown in the table below for operation at 500 kta MEG. Yields are expressed as t/t MEG and t/t EOE, in which EOE is the EO mass equivalent of the mass of the MEG, DEG and TEG produced . It is calculated as:

$$EOE = 0.71 * MEG + 0.83 * DEG + 0.88 * TEG$$

		HA catalyst		HS catalyst	
		SOC	EOC	SOC	EOC
Catalyst selectivity	%	81.60	76.70	87.95	80.40
Ethylene / EOE	t/t	0.80	0.85	0.74	0.81
Oxygen / EOE	t/t	0.86	1.04	0.67	0.90
Ethylene / MEG	t/t	0.64	0.68	0.59	0.65
Oxygen / MEG	t/t	0.69	0.83	0.53	0.72

For HA the expected catalyst life will be 4 years. For HS catalyst the expected catalyst life will be 2 years.



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1.7 Process materials consumption figures

The table below gives estimated consumption figures for the process materials that are consumed continuously.

Estimated consumption figures		HA catalyst		HS catalyst		max.
		SOC	EOC	SOC	EOC	
Methane	t/hr	0.60	0.60	0.64	0.63	1.20 (1)
NaOH (20%w)	ltr/hr	46	178	46	178	(2)
Potassium carbonate	%	2	2	2	2	(3)
Antifoam	ltr/hr	-	-	-	-	(4)
Ethylchloride moderator	ltr/hr	1.1	1.6	1.4	3.7	8.2 (5)

Notes:

- (1) Maximum import for charging the recycle gas loop.
- (2) Depends on acids make in the EO reaction system
- (3) Entrainment loss
- (4) The normal consumption cannot be estimated.
- (5) Moderator injection depends on feedstock purity

Catalyst

For HA the expected catalyst life will be 4 years. For HS catalyst the expected catalyst life will be 2 years. Catalyst volume is approximately 284 m³.

The reactor requires loading of inert balls in the heat transfer section of the catalyst tube and springs in the bottom of the tubes to keep the catalyst in place. Inert balls will be provided under the catalyst supply agreement. The method of installation of inert balls and springs is shown in Process Memorandum 101.

Methane

Methane consumption depends on the purity of the make-up oxygen and make-up ethylene. For the material balance the purities were assumed to be in accordance with the specification provided in the kick-off meeting.

Caustic

Caustic consumption depends on the acids make in the system, which depends on the presence of rust and operating temperatures in the reactor system. The caustic consumption varies widely in various plants.



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Potassium carbonate

Steady state potassium carbonate consumption is zero. Due to mechanical losses a make-up rate of about 2% per year may be expected.

Antifoam

Antifoam consumption depends on the presence of foam promoting conditions in the system. Antifoam consumption varies widely in various plants. If excessive foaming occurs a batch of anti-foam can be injected manually in the suction of the carbonate pump.

Ethylchloride moderator

Ethylchloride moderator consumption depends on the type and age of the catalyst and ethane concentration in the loop gas, which depends on make-up ethylene purity. On average moderator consumption will most likely be between 1.0-1.4 kg/hr for HA catalyst and 1.2-3.1 kg/hr for HS catalyst, but can be as high as 7 kg/hr in case the make-up ethylene quality deteriorates.

Activated carbon

The sulphur guard beds are designed for life of the activated carbon of two years. The exchange interval of the mobile adsorber drums for treatment of the moderator vent is expected to be a minimum of about 40 days per drum.

Resin

The MEG polishing bed is designed for an exchange interval of the resin of one year.

1.8 Utilities consumption figures

For utility consumption figures based on detailed engineering refer to "Utilities consumption table"; LST-PR-004.

1.9 Waste disposal

The source and the destination of waste disposal are specified in Process Memorandum 118.



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2. PROCESS DESCRIPTION

The plant is designed for operation on both HA and HS catalyst. The plant is designed taking into account the effect of catalyst ageing. Thus process conditions may differ between start of catalyst cycle (SOC) and end of catalyst cycle (EOC).

Where necessary, four figures are given: HA SOC (EOC) / HS SOC (EOC). For process conditions for which there is no relevant difference between HA and HS catalyst, only two figures are given: SOC (EOC). In case there is no relevant difference between either SOC (EOC), only one figure is given. Graphs of typical temperature profiles are generated for HA EOC case.

2.1 EO reaction (PFD No. : Unit 100-1/2, 2/2)

Refer to the process flow diagrams for the EO Reaction section.

2.1.1. Recycle gas loop (PFD No. : Unit 200-1/2, 2/2 and Unit 100-1/2,2/2)

After compression in the recycle gas compressor, C-2001, part of the recycle gas is sent through the CO₂ absorber, T-2001, for CO₂ removal. The design CO₂ concentration in the loop gas is 4.5(5.5) / 1.0(2.0) mole% at the EO reactor inlet. A lower CO₂ concentration is achievable at SOC than at EOC due to higher catalyst selectivity at SOC. When operating with HA catalyst approximately 50%, or about 475 t/h, of the recycle gas is fed to the CO₂ absorber to remove CO₂. For HS catalyst the gas flow to the CO₂ removal section is slightly higher at 55% of the total flow or about 500 t/h. The flow is controlled by means of FIC-20101 that gives the set point to PDIC-20104. After CO₂ removal the treated gas is recombined with the recycle gas.

Make-up ethylene is supplied to the recycle gas. Thereafter, the recycle gas is mixed with ethyl chloride that is passed through FV-10302A/B to control the flow of EC and enters the oxygen mixing nozzle, Z-1001, where make-up oxygen feed is supplied to the process. The oxygen mixing nozzle ensures rapid mixing under controlled conditions. The oxygen concentration varies from 100% at the oxygen mixing spargers, to about 8.2(8.0) / 8.2(7.9) %mole in the loop gas after complete mixing. This means that the mixture composition will be in the flammable region in the space directly downstream of the oxygen mixing nozzle. After mixing completely, the mixture will be below the lower flammable limit of the gas. After addition of moderator, the gas is preheated in the feed product exchanger, E-1001, from 48(48) / 46(45)



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°C to 168(174) / 172(179)°C by the EO reactor product gas. The entering flow to the reactor system is controlled by means of FIC-10301.

The preheated reactor feed gas enters the two EO reactors, R-1001A/B, where ethylene and oxygen react over a silver catalyst to produce EO. By-products of the reaction are CO₂ and water as well as some aldehydes and acids. The product gases leave the reactors at a temperature of 222(250) / 242(275) °C. Pressure of reactor system is controlled by PIC-20201 that actuates two control valves, PV-20201A/B while the flow to the reactors is controlled by FIC-10304.

The reactor product gases are cooled in 3 stages. MP steam (14K) is generated in the primary product cooler, E-1002, thereby cooling the product gases to 204(211) / 209(218) °C. In the feed product exchanger, E-1001, the product gases are cooled to 89(91) / 88(91) °C and thereafter cooled to 56°C in the secondary product cooler, E-2005, by fat absorbent from EO absorber, T-2003.

The recycle gas from the oxygen mixing nozzle is below 50°C and as a result local condensation may occur in the hot product side of the feed product exchanger. Frequent draining of the condensate is required to prevent built-up and polymerisation. The cold product gas leaving the feed product exchanger is above the dew point.

Two motor operated block valves together (MOV-10301 and MOV-10302) with a bypass valve, FV-10301 enable the EO reactors to be quickly isolated from the recycle gas loop to protect the catalyst. A bypass around the EO reactors allows normal circulation of the loop gas in case the reactors would be out of operation.

EO is absorbed in the EO absorber, T-2003. The recycle gas compressor, C-2001, recycles the overhead gas from the EO absorber partially to the CO₂ removal section and partially to the oxygen mixing nozzle. A knock out drum protects the compressor.

2.1.2. Ethylene and methane feed treatment (PFD No. : Unit 100-1/2)

A very low sulphur concentration in the ethylene is required to prevent significant catalyst poisoning. A sulphur guard bed with activated carbon, D-1005 ensures that the sulphur level can be reduced to below 0.01 mg/kg. An analyser must monitor the sulphur concentration in the feed.

Although the methane make-up is low compared to the ethylene feed, a sulphur guard bed is also required in view of the high sulphur concentration. Methane is supplied from natural gas and compressed by Methane booster compressor, C-1002A/B and purified in Methane purification unit, PK-1001



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including sulfur absorber. This package has been designed to maximize Methane content of feed natural gas by catalytic reforming. The purified methane is injected into the carbonate flasher, D-2002, to strip hydrocarbons from the fat carbonate. The residual gas compressor, C-3001, compresses the methane with the flashed gas from the carbonate flasher into the recycle gas loop. Alternatively, if the residual gas compressor is out of service, methane may be injected directly into the recycle gas loop upstream the Oxygen mixing nozzle, Z-1001 or into residual gas overhead of carbonate flasher in order to control methane concentration in the CO₂ vent.

The sulphur analyser, AT-10901 alternately analyses ethylene and methane.

Filters downstream of the guard beds, FT-1002A/B and FT-1005A/B protect the gas loop from any particles entering.

In case of oxygen shut-down the ethylene and methane feeds are automatically cut off by means of FV-10102 and FV-10104 respectively.

2.1.3. Oxygen mixing (PFD No. : Unit 100-1/2)

Oxygen from battery limit passes through a filter, FT-1001 before mixing with the hydrocarbon containing recycle gas. An on/off valve HV-10102 at B.L. is considered to be able to cut off the oxygen feed. Oxygen must be supplied free from particles, rust and any foreign matter. The filter acts as a guard filter to protect the oxygen mixing nozzle. In case the oxygen filter would need to be cleaned more than once per four years, the oxygen must be more thoroughly filtered and cleaned at the source of supply.

The oxygen concentration in the plant must be kept below the lower flammable limit. This requirement can not be fulfilled in the oxygen mixing nozzle where oxygen concentrations range from 100 percent to approximately 8.2(8.0) / 8.2(7.9) %mole in the loop gas. The design of the oxygen mixing nozzle minimises the space where oxygen concentrations are in the flammable region. In view of the fact that the mixture is in the flammable region at the oxygen mixing nozzle, a safeguarding system is provided.

Refer to Process Memorandum 103 for the calculation method for the lower flammable limit.

The oxygen feed flow control, FV-10202 which worked based on middle of three flow transmitters, FT-10202A/B/C and shut-down system have automatic block and bleed valves as well as nitrogen purge facilities to prevent inadvertent mixing of oxygen with loop gas and to purge the system in case of potentially unsafe situations. The rate at which the oxygen feed flow set point can be increased by the operator is limited via a ramp function.



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This means that any step change in set point will result in a timed ramp response from the DCS. The rate of the ramp must be less than the trip setting for the rate of change of the flow. Pressure of oxygen feed is controlled by PV-10216.

The oxygen shut-down system activates two quick closing block valves (SDV-10201 and SDV-10203) in less than one second, in addition to the normal oxygen flow control valve (FV-10202). As soon as the quick closing valves are shut, two automatic bleed valves (SDV-10202 and SDV-10204) vent the line sections between the flow control valve and the quick closing block valves. Nitrogen is rapidly injected downstream of this assembly to purge the line towards the mixing nozzle and avoid the formation of a flammable mixture. The initiation of trip system actuates process alarm in DCS, XA-10202A a big bell and XA-10202B air horn to insure immediate operators attention. The sequence of events record will show all events in the system in chronological order for cause analysis.

Prior to start of the oxygen feed to the system, the mixing nozzle must be purged with nitrogen. The flow of nitrogen must be maintained through the oxygen mixing nozzle until the oxygen flow is continuous and stable and high enough to prevent any back mixing of loop gas into the oxygen system. A series of tests should be carried out on the oxygen mixing nozzle to determine the flow of start-up nitrogen required to purge the oxygen line downstream of the last block valve. For more details refer to section 4.4.1.

Trip initiators are provided to shut down the oxygen feed system in case of unsafe conditions. Refer to Process Memorandum 111 (Oxygen Feed Flow Control and Shut-down System) for further details.

Oxygen mixing nozzle

The design of the oxygen mixing nozzle, Z-1001 enhances rapid dilution of oxygen over a very short distance, which, combined with measures taken to avoid ignition sources, ensures a safe mixing operation.

Rapid mixing is obtained and back-flow of hydrocarbon gases in the oxygen feed line is prevented by proper selection of the velocities of the two gas streams. The distribution of oxygen is accomplished by injecting the oxygen into the hydrocarbon stream through three concentric spargers of different diameters mounted like donuts inside the loop gas piping. A minimum of 10 pipe diameters straight pipe length is provided upstream and a minimum of 10 pipe diameters straight pipe length downstream to obtain a uniform velocity and pressure profile and ensuring proper mixing.

The oxygen mixing nozzle is located at the periphery of the plant, away from the main plant equipment and piping. It is mounted horizontally and surrounded by a protective concrete wall.



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2.1.4. Nitrogen purge system (PFD No. : Unit 100-1/2)

Nitrogen for the nitrogen purging of the oxygen system is pressurized by the nitrogen safeguarding compressor C-1001A/B. Nitrogen to the safeguarding compressor is protected against back-flow from other sources to prevent contamination of the safeguarding nitrogen.

The nitrogen purge for start-up and shut-down requires two vessels with different functions.

Upon shut-down the nitrogen buffer vessel D-1001 delivers a flow that purges the volume of the oxygen system within 5 seconds through SDV-10205 and SDV-10206. The purge nitrogen flow rate must never exceed the design oxygen flow rate (DOF), since this nitrogen is displacing oxygen inventory into the recycle gas stream. The system has been specified such that the total volume of nitrogen introduced during the first five seconds is at least the volume of oxygen present in the oxygen line and mixing nozzle sparger rings. After 5 seconds, the nitrogen flow rate decreases to a flow rate of about 0.05 times DOF due to FO-10206. This small purge flow supplied from nitrogen storage vessel, D-1002 should be adequate to keep the oxygen piping between SDV-10203 and the mixing nozzle free from hydrocarbons. If the shutdown takes long duration, the nitrogen purge flow may be decreased to about 0.005 times DOF. This is done by unlocking and opening of HV-10206 and closing of manual valve upstream of PCV-10209 such that the nitrogen flow is set by FO-10207. It is recommended to maintain this very small purge flow of nitrogen as long as recycle gas loop is not depressurized.

The mixing of hydrocarbons with oxygen anywhere else than in the downstream part of the mixing nozzle must be prevented. After a shutdown hydrocarbons may be present in the oxygen feed line downstream of SDV-10203 and will certainly be present in the mixing nozzle sparger rings. It is imperative that these parts are free of hydrocarbons before oxygen is admitted. This is achieved by purging with nitrogen for 3 minutes with flow equivalent to 0.26 DOF through PCV-10220, FO-10204 and HV-10205. FO-10204 must be calibrated before the first start-up at the minimum nitrogen pressure that is the set point of PALL-10208.

The nitrogen storage vessel D-1002 provides start-up nitrogen and provides nitrogen to the nitrogen buffer vessel D-1001. The vessel is designed for one shut-down and one start-up, based on at least 3 minutes purging of the oxygen mixing nozzle during start-up and filling of the nitrogen buffer.

When the pressure in the nitrogen storage has been as low as the pressure of the recycle gas loop, the nitrogen storage must be sampled and analysed



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to check for presence of hydrocarbons due to back-flow of loop gas. In case of contamination, the vessel must be thoroughly purged and recharged.

Process Memorandum 111 provides a detailed description of the function of the nitrogen vessels in relation to start-up and shut down of the oxygen mixing nozzle.

2.1.5. Ethylchloride moderator injection (PFD No. : Unit 100-1/2)

In order to obtain optimum yields of ethylene oxide, it is necessary to provide ethylchloride (EC) as moderator to the catalyst to suppress the reaction to carbon dioxide and water.

The required injection rate depends on the concentration of ethylene and ethane in the loop gas. The expected required injection rate will be between 0-3500 g/hr. The maximum required flow rate may be up to 7000 g/hr in case of pre-soaking of fresh HS catalyst at initial start-up or in case of very high concentrations of ethane or higher alkanes. As there are two valves, FV-10302A/B in parallel the flow rate could be increased temporarily. Normally only one valve should be operated, with the other valve as an installed spare. Moderator dosing is critical for catalyst performance. Fouling and vibrations of the control valve and flow meter will result in performance degradations of the catalyst.

The response of the catalyst to moderator dosing is different for HA and HS catalyst. The required moderator concentration is expressed as I-factor, which is defined in the section 6.1.2.2 paragraph on moderator make-up. The required injection rate is based on a certain target I-factor. For HA catalyst the target is constant over the life of the catalyst. For HS catalyst the target varies with the ageing of the catalyst and needs to be re-optimised frequently.

For HA catalyst the ethylchloride acts as a moderator since the required coolant temperature will increase. For HS catalyst, the moderator actually promotes the activity of the catalyst.

Moderator is filtered in moderator EC filter, FT-1004A/B and then injected into the recycle gas loop upstream of the oxygen mixing nozzle, Z-1001. In shutdown case EC is cut off.



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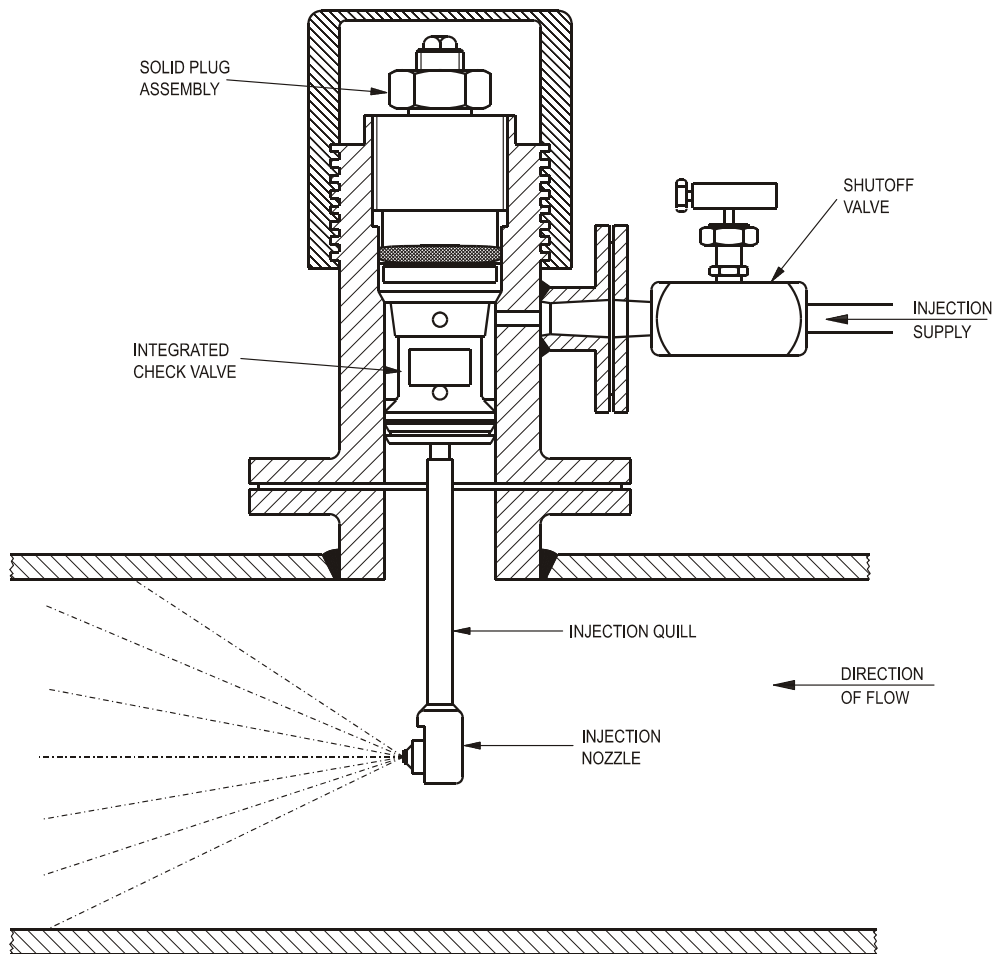


Figure 2.1.5-1: Typical chemical injection quill for injection into gas lines

Moderator is pumped from the transport container into the moderator feed vessel D-1003 by moderator loading pump, P-1003. A level switch, LSHH-10304 protects the vessel from overfilling. The moderator feed vessel is pressurized with ethylene from the ethylene feed line and will be injected under pressure into the gas loop. Accurate flow control in steps of 25 g/hr is required to optimize HS catalyst performance. In case of oxygen shut-down, the moderator injection is stopped.

Moderator feed vessel and EC system are insulated to minimize venting due to the daily temperature cycle.

The pressure of the moderator feed vessel must be kept constant to avoid flow variations of the moderator into the gas loop. Proper adjustment of the split range pressure control by means of split range control valves, PV-10309A/B is critical in this respect. The gas outlet is vented to atmosphere via carbon adsorber bed, Z-1002A/B to absorb ethyl chloride.



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From the source of supply, the delivery time may be several weeks. This requires careful inventory control and planning since the process cannot be operated without moderator. The residence time in the moderator feed vessel may vary between about three weeks up to almost 5 months depending on the required dosing rate. The use of dedicated transport containers is required to protect the cleanliness of the moderator system. Fouling by small particles and rust will make the system very hard to operate. The containers need to be thoroughly cleaned prior to initial use. Due to the typical volume of transport containers a moderator container could be on-site for a period of up to about one year before being returned to the source of supply for a refill. Permanent storage facilities for the transport container, including a shelter and connection to a relief system, are required. Moderator will vaporise and is heavier than air and poses a severe fire hazard. Containers may rupture or explode if exposed to heat. Moderator vapour may ignite at distant ignition sources and flash back.

Due to the very low flows, the moderator injection flow measurement and control system is sensitive to vibrations, which can lead to measurement and control problems. Stiff supporting and installation on a concrete ground floor are aimed at preventing vibrations.

Start-up of high selectivity catalyst and moderator injection at start-up is described in Process Memorandum 102 (Start-up and Shut-down of EO Catalyst).

2.1.6. EO reaction (PFD No. : Unit 100-2/2)

Catalyst selectivity depends primarily on the type of catalyst and catalyst age, although the other factors affecting selectivity should be controlled to optimise selectivity.

The ethylene converted via the side reaction forms CO₂, water and some aldehydes and acids. A total acetaldehyde make of 0.0005 mole ACH/mole EO in the reactor product stream is assumed for the design. A total formaldehyde make of 0.00012 mole FAH/mole EO in the reactor product stream is assumed for the design.

The material balances reflect the specific selectivity for HA and HS catalyst at both SOC and EOC.

In the material balance it is assumed that methane and ethane, as well as obviously argon and nitrogen, are completely inert to the reaction under the



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selected conditions. Ethane, argon and nitrogen are removed from the loop gas by the process vent, a continuous purge from the top of D-2004.

The loading of the catalyst is expressed using the concept of work rate (WR) that represents the kg EO produced/m³catalyst/hr.

During normal operation, water is vaporised along the catalyst containing section of the tubes, driving the thermosyphon coolant circulation. The steam rises in the shell and a portion of it re-condenses in the top section thus heating the feed gas to reaction temperature. The steam leaves the shell of the reactors as a two-phase mixture via four steam risers.

The steam from both reactors is separated from the coolant in the common HP steam drum, D-1007. The generated steam leaves the HP steam drum via two pressure control valves. These control valves, PV-10603A/B have been provided to control pressure of steam. The steam pressure controls the temperature of the reactor. During HS run the biggest one (PV-10603A) should be completely closed to avoid disturbances in the HCS header.

Make-up boiler feed water is provided to the HP steam drum. The coolant flows back to the reactor shell. The coolant is slightly sub-cooled upon entering the reactor due to the mixing with the colder boiler feed water. The heat transfer coefficient in the bottom section of the tubes in the EO reactor will therefore be lower than elsewhere in the reactor.



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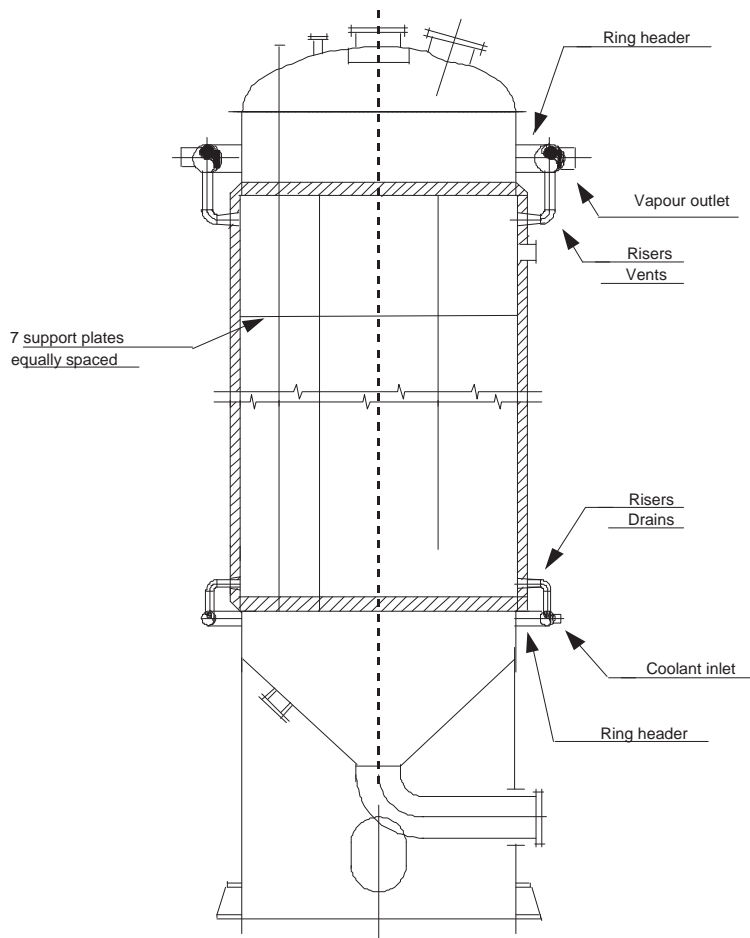


Figure 2.1.6-1 EO reactor

Each reactor is equipped with 4 vapour outlet risers of 16" each. Together they form the symmetrical transport lines for the coolant/steam mixture from the reactor to the HP steamdrum during HA catalyst operation. When the applied catalyst is HS catalyst, only two vapour outlet risers are required for stable operation. In that case, two non-adjacent risers shall be blinded off close to the ring header such that the remaining line-up will be symmetrical.

The reactant gases normally enter the top head of the reactor at 168(174) / 172(179)°C and are heated in the preheat section to within 2°C of the coolant temperature. Upon entering the catalyst bed, the gases react exothermally and heat up to a peak temperature difference (PTD) between catalyst and adjacent coolant of 10-15°C. As the catalyst ages, the point at which the PTD will be reached shifts towards the bottom end of the tubes

with the PTD increasing slightly. Higher oxygen conversions will also give increased PTD's.



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Reaction temperatures are measured by 6 tubes with 5 thermocouples each in the catalyst thermocouple tubes. Coolant temperatures are measured by 3 tubes with 5 thermocouples each inserted in closed and perforated reactor tubes. The local temperature difference between the reaction gases and the coolant depends on the extent of reaction and the rate of heat transfer between reactants and coolant. Adjusting the reactor steam drum pressure, which in turn fixes the coolant temperature, controls the extent of reaction, or conversion. The coolant temperature required for design conversion depends upon several factors such as feed composition, gas feed flow rate, moderator injection, type of catalyst and catalyst age.

Although the gas temperature is considerably higher than the coolant over most of the tube length, the tube wall temperature is close to the temperature of the coolant. This is due to the high heat transfer coefficient from the tube wall to the boiling coolant compared to heat transfer at the gas side. The preheat-section tube wall is cooler than the coolant by not more than a few degrees averaged over the tube preheat length.

The EO reactor is constructed as a large fixed tube sheet heat exchanger containing 11088 tubes, of which 11085 are filled with catalyst and inert balls in the top. The reactor tubes have a diameter of 44.9 OD / 38.9 ID mm and are 11.8 m long, excluding tubesheets. The coolant is used on the shell side to remove excess heat and to heat the entering reaction gas up to reaction temperature. The heat of reaction leaves the reactor as sensible heat with the gas, which rises in temperature over the reactor. The remainder of the heat is transferred to the coolant. Support grids are required to provide intermediate tube support.



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EO reaction conditions

		HA SOC	HA EOC	HS SOC	HS EOC
Ethylene conversion	%	12.8	13.6	11.9	13.0
Oxygen conversion	%	37.5	46.3	28.9	40.5
Selectivity	%	81.6	76.7	88.0	80.4
Oxygen inlet	% mol	8.20	7.96	8.23	7.94
Flammable limit inlet	% mol	9.00	8.76	9.03	8.74
Margin at inlet	% mol	0.80	0.80	0.80	0.80
Oxygen outlet	% mol	5.19	4.33	5.93	4.78
Flammable limit outlet	% mol	7.43	6.48	6.95	5.83
Margin at outlet	% mol	2.24	2.15	1.02	1.05
GHSV	Nm ³ /m ³ /hr	3500	3500	3500	3500
Work rate	kg/m ³ /hr	180	180	180	180
Pressure inlet	bara	17.5	17.5	17.5	17.5
Temperature outlet	°C	222.0	250.0	242.0	275.0

Refer to the paragraph on flammable limits and Process Memorandum 103 (Reactor Loop Gas Flammable Limit Calculations) for further details.

2.1.7. Reaction control (PFD No. : Unit 100-2/2)

The major variables to be controlled for the EO reaction are conversion and selectivity while observing the flammable limits and maintaining proper feed gas composition. For a discussion of the factors affecting the performance of the EO reaction refer to paragraph on operating variables of the EO reaction section.

The reactor control system depends primarily on the control of the coolant temperature. The coolant temperature is determined by the pressure control of the HP steam drum. Adjusting the pressure in HP steam drum controls the boiling temperature of the coolant of both reactors. The coolant temperature controls conversion, which, together with the gas feed flow and selectivity, determines the heat flow to the coolant.

The coolant temperature is well below the runaway reaction temperature. In a runaway reaction all oxygen is burnt to CO₂ via the non-selective side reaction, which has a significantly higher heat of reaction and which accelerates itself.



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The temperature profile of the EO reactor is of the type normally encountered with exothermic reactions. In the reactor, the gas temperature increases very quickly in the preheating section of the tube. Sensible heat is supplied to the gas by condensing coolant vapour outside the tubes. In the active zone of the tube, the initial rate of temperature rise is dependent primarily on catalyst activity. The subsequent temperature profile is determined by decreasing oxygen concentration (which slows the reaction rate) and the heat removal rate by boiling the coolant.

The peak temperature difference (PTD) is the maximum difference between catalyst temperature and coolant temperature measured at the same tube level and increases with increasing coolant temperature. For aged catalyst, the coolant temperature required to maintain conversion must be increased and the location of the PTD will shift towards the end of the tubes.

The margin from runaway (MFR) is the difference between the coolant temperature at runaway conditions and the actual coolant temperature. When increasing the heat generation of the catalyst, such as by increased conversion, the effective MFR will reduce. At design conditions, the MFR should remain relatively constant with catalyst age.

When the temperature of any of the catalyst thermocouples rises about 20°C above the highest normal temperature, it would be an indication that a runaway reaction may be developing or has developed. The reactor temperature profile during a runaway is completely upset due to the local total conversion.

When a runaway reaction is detected, the oxygen feed will be stopped by automatic intervention of the oxygen shut-down system. A runaway may result in an unsafe situation in the plant and the yield during the runaway period is zero and further operation is wasteful. A narrow zone of the catalyst may be deactivated by runaways of short duration. Prolonged runaways will lead to permanent damage of the catalyst. A sustained runaway is never purposely permitted, although a few poorly packed tubes may operate continuously under runaway conditions. Proper catalyst loading, careful pressure drop measurement and correction of tubes with a too high or too low pressure drop is critical in this respect.



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Typical reactor temperature profile for fresh S863 catalyst

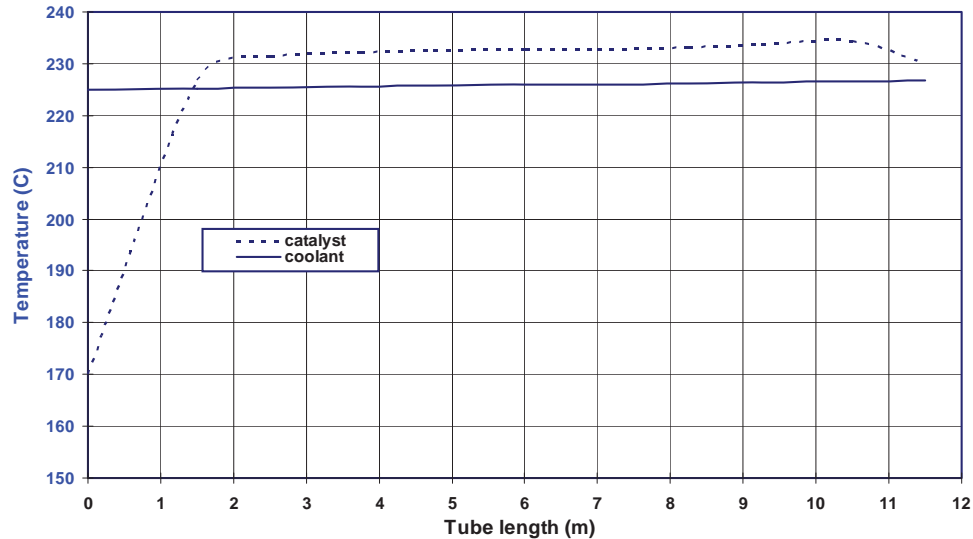


Figure 2.1.7-1: Typical temperature profile for S-863

Typical reactor temperature profile for fresh S882 catalyst

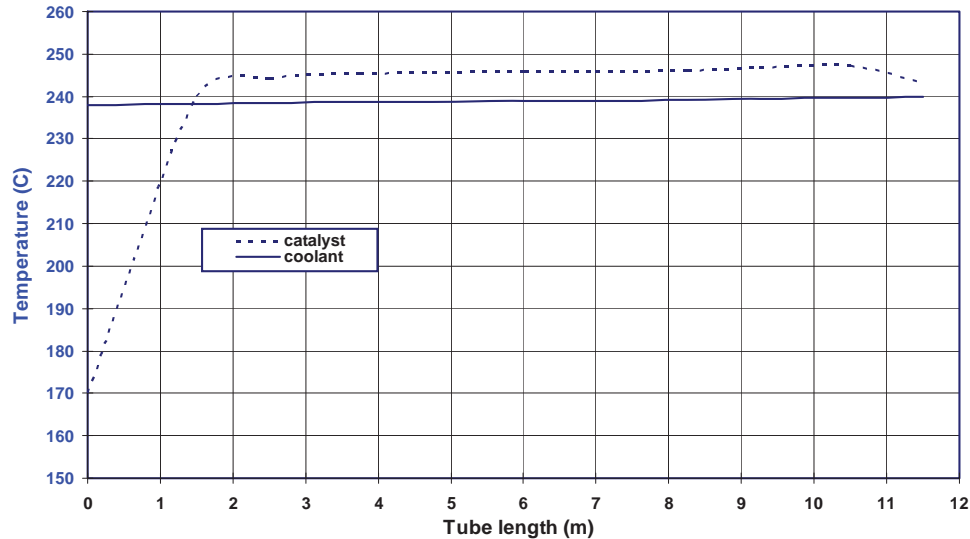


Figure 2.1.7-2: Typical temperature profile for S-882



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Typical Peak Temperature Difference verses coolant temperature for fresh S863 catalyst

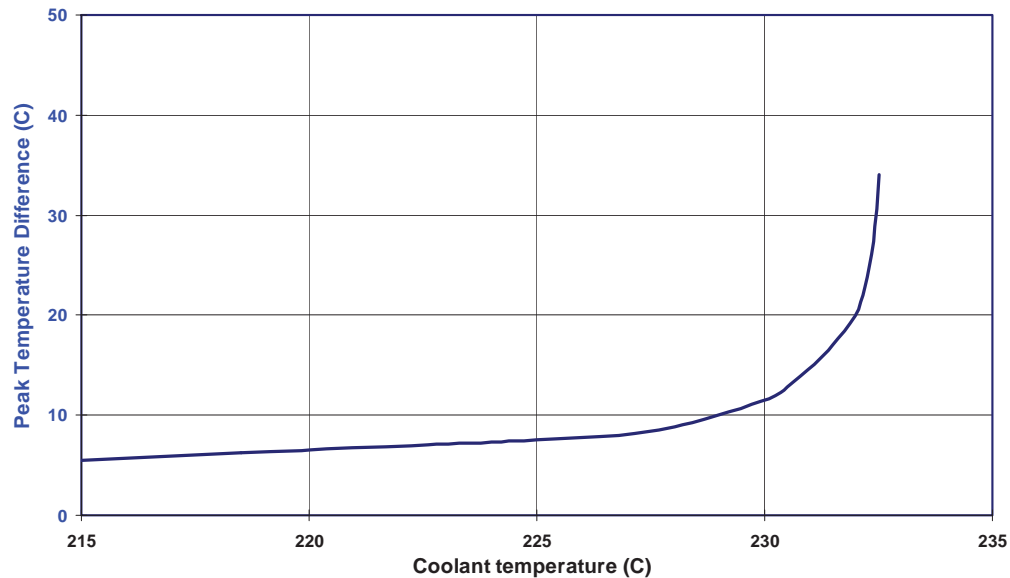


Figure 2.1.7-3: Typical temperature profile for S-863

Typical Peak Temperature Difference verses coolant temperature for fresh S882 catalyst

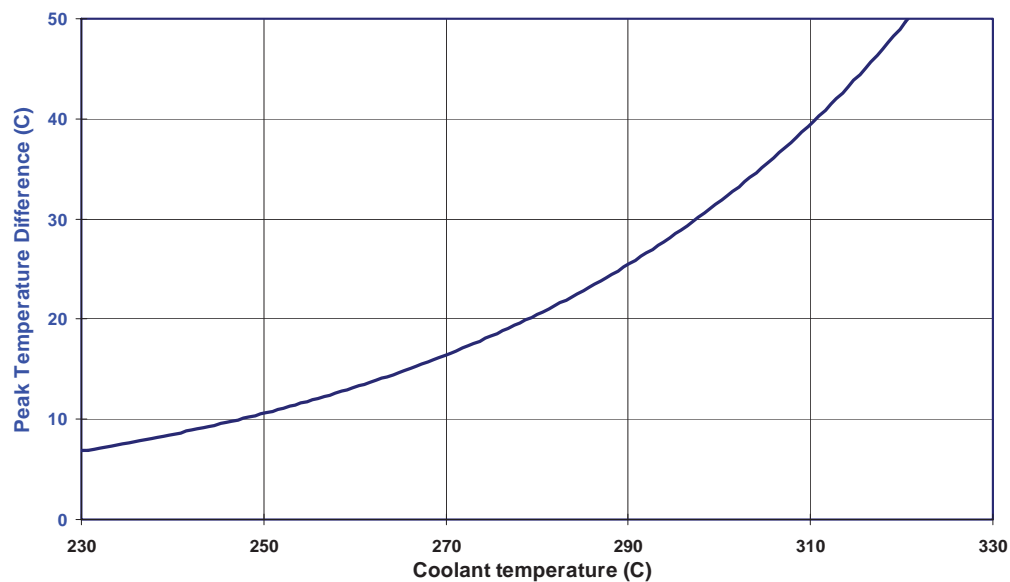


Figure 2.1.7-4: Typical peak temperature difference for S-882



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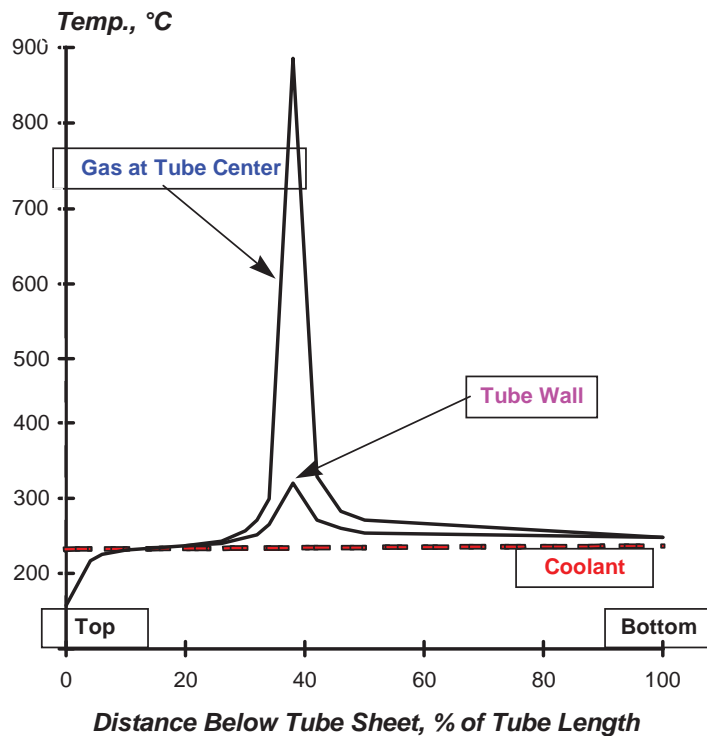


Figure 2.1.7-5: Expected temperature profile during runaway



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2.1.8. Catalyst performance decline

The activity of both HA and HS catalyst gradually declines over time and as a function of cumulative production. The decline is probably due to both feed impurities and agglomeration of silver on the catalyst surface. The catalyst decline rate has varied significantly amongst licensees due to variation in feedstock impurities and severity of operation and possibly other factors.

It has been found profitable to remove as much as possible trace impurities from the feed gases to prolong catalyst life. Refer to Process Memorandum 107 for details of various catalyst poisons.

Frequent review of catalyst performance both by the plant in cooperation with catalyst vendor is warranted to ensure optimum operation of the catalyst.

As activity declines, the yield decreases, requiring increasing coolant temperatures and pressures to increase conversion.

2.1.9. Steam generation (PFD No. : Unit 100-2/2)

Generated steam from the EO reactor (CS 20K)

The heat of reaction is removed by the loop gas and by vaporising water in the shell side of the reactor. The reactor coolant system operates on the thermosyphon principle. The reactors R-1001A/B have one common HP steam drum. Coolant from the HP steam drum, D-1007, enters the bottom of the reactor shell via a ring header and distribution branches. The steam and coolant leave the reactor shell via a branched outlet ring system and flows back into the HP steam drum, where steam is separated from the coolant and discharged into the CS 20K header. The coolant is mixed with make-up boiler feed water.

To ensure stable thermosyphon operation the steam drum is located such that the minimum liquid level is 16 m above the liquid coolant inlet nozzle of the EO reactors.

The pressure of the coolant system determines the temperature, conversion and selectivity of the EO reaction.

Total HP steam generation rates depend on the type of catalyst, the performance of the catalyst and vary between SOC and EOC.

The liquid volume of the HP steam drum, D-1007, is based on a hold-up time of 1 minute for the circulation flow plus a hold-up time of 3 minutes for the



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sum of the product streams (the blown down stream plus the HP steam). At low level of the HP steam drum the oxygen shut down system is activated. The minimum inventory of water is required to be able to convert all oxygen in the loop in a non-selective run-away reaction and remove this heat of the reaction in case of a boiler feed water trip.

In order to maintain the reactor coolant quality and prevent build-up of impurities, a continuous blow-down into the MP steam drum D-1006 is maintained. The pH of this stream should be controlled. A dump valve, HV-10601 allows to rapidly discharge coolant in case of high level.

Make-up boiler feed water for the HP steam generation is provided by the BFW pump, P-6001A/B from the deaerator, D-6001. It is preheated in the boiler feed water preheater, E-1004A/B to 163/156/177/161 °C using CS 14K condensate from MP steam drum, D-1006 and sent to HP steam drum, D-1007. BFW flow is controlled by FIC-10605 that takes the set point from HCS flow rate, FI-10601 and HP steam drum level LIC-10601. To secure the supply of boiler feed water, the spare BFW pump will start automatically in case of too low water supply pressure.

The CS (20K) with flow control, FIC-10604 heats the HP steam drum in start-up case, while HV-10602 depressurise the system in shutdown case. The system is completed by condensate circulation pump, P-1001 reactor shutdown cooler, E-1003 and temperature control loop TIC-10602 (working during pre start-up and shut down cases).

A boiler feed water specification applies. Any residual hardness will precipitate as a hard scale which is difficult to remove and will hamper heat transfer. The boiler feed water quality is critical to the performance of the process because boiler feed water enters, via the steam system, into the process. The pH of the BFW blow down from the HP steam drum should be maintained above 9.5 at 25°C.

Generated steam from the primary product cooler (CS 14K)

The reactor product is cooled from 222(250) / 242(275)°C to 204(211) / 209(218)°C by generating steam at 14K (198)°C in the primary product cooler E-1002. E-1002 is located as close as possible to the reactor exit to minimize product residence time. Condensate from the MP steam drum D-1006 flows to the shell side of the primary product cooler, E-1002. The steam and condensate leaving E-1002 return to vessel D-1006 by thermosyphon circulation. To ensure stable thermosyphone operation, MP steam drum is located such that the minimum bottom T.L. elevation of D-1006 is at least 12m above bottom tubesheet of E-1002.

Make-up to D-1006 is provided by CS 20K condensate from the glycol reactor preheater-3, E-4003, and the concentrator-1 reboiler, E-4004, blowdown from HP steam drum and boiler feed water. These condensate



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streams will flash to some extent. The generated steam from D-1006 supplies the CS 14K steam header. Condensate from the MP steam drum is used to preheat the boiler feed water (E-1004A/B) and is combined with CS 14K condensate from the reboilers of unit 400/500 and is discharged to the steam condensate flash drum, D-4001. Split range controllers on condensate outlet, LV-10801B and BFW inlet, LV-10801A control the level of the drum.

Total MP steam generation rate depends on the type of catalyst and varies between SOC and EOC. The liquid volume of the MP steam drum is based on an hold-up time of 1 minute for the circulation flow plus an hold-up time of 3 minutes for the sum of the two product streams (the blowdown stream plus the MP steam). This leads to a hold-up time of more than 20 minutes inventory in terms of the MP steam produced.

At low level in the MP steam drum the oxygen shut down system is activated. The minimum inventory of water is required to be able to remove the heat that would result from conversion of the oxygen in the loop during a post-ignition in case of a boiler water feed water pump trip, part of which would be removed via the primary product cooler.

2.1.10. Analysers

Two analyzer houses have been considered in process unit. Analyzer house 1 is located at the ground floor of structure A and including all analysers relevant to EO production units. Analyzer house 2 is located near glycol slop tank and consisting of MEG/ DEG and TEG product analyzer.

Continuous and fast analysis of reactor feed and product gases is essential for safety and optimum operation.

The analyser system comprises of a primary fast loop, which provides a sample of the gas at reduced pressure to the analyser house, where a fraction of the sample is diverted to a secondary loop leading to the analyser instruments. The gas from the primary fast loop is recovered via the residual compressor suction k.o. drum, D-3002. Low flow in the primary fast loop will trip the oxygen supply.

The primary fast loop ensures that the sample gas will arrive within a few seconds at the analyser house. The total maximum allowable response time between a change in oxygen feed flow to the response of the oxygen analyser must be well below 20 seconds.

Around the EO reactor the following analysers are installed (at analyzer house 1):



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- Two oxygen analysers, AI-10903A/B in the common reactor feed gas stream, one oxygen analyser, AI-10906 in the common reactor product gas stream with a common spare, AI-10911 for the reactor feed gas and reactor product stream. The oxygen analysers are connected to the oxygen shut-down system.
- A common oxygen analyser, AI-10918, switching between the reactor outlets, connected to each product outlet for performance monitoring of each individual reactor. This analyzer will not be connected to the oxygen safeguarding system.
- An on-line process gas chromatograph, AI-10904 to analyse CH₄, C₂H₄, C₂H₆, CO₂, N₂ and O₂ + Ar in reactor feed and product
- A separate on-line process gas chromatograph, AI-10908 to determine EO in the common reactor product line for reactor selectivity calculations
- A separate ethylene analyzer, AI-10902 in the reactor inlet gas for ethylene feed rate control.
- Carbon monoxide analyser, AI-10909 on the common reactor feed to detect pre-ignition. It is infrared absorption type.
- On-line process gas chromatograph, AI-10912 to measure the EO concentration at the recycle gas outlet from EO absorber for the performance of the EO absorber and at the overhead gas outlet from the residual absorber for the performance of the residual absorber. The analyzer alternatively switches between the EO absorber overhead and the residual absorber.
- As an alternative to gas chromatographs, a mass spectrometer can be used to analyse CH₄, C₂H₄, C₂H₆, CO₂, N₂, O₂, Ar, and EO. Irrespective of the selected analyser type, the reactor inlet gas and product gas streams are analysed alternately. The results are used for flammable limit calculations, reaction selectivity calculations, control of performance of the CO₂ removal section and to maintain optimum concentrations in the loop gas.
- On-line continuous sulphur analyser that alternately analyses the ethylene and methane feeds for total sulfur to safeguard the catalyst.
- Chloride analyser, AI-10905 in the reactor inlet stream in order to determine the chlorinated hydrocarbons for calculation of the I-factor and catalyst performance optimization.



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MEG, DEG and TEG product analysers (at analyzer house 2) are installed as following:

- Three on-line gas chromatographs at MEG, DEG and TEG product streams to storage area to determine composition of MEG, DEG and TEG alternatively.
- An UV analyzer of MEG, AI-10916 at MEG product stream from E-5004.

Equations for calculating actual selectivity and conversion from the inlet and outlet concentrations of the analysers around the EO reactors are given in Process Memorandum 104.

For more information on the sampling systems and analysers refer to Process Memorandum 109 (Sampling Systems & Sample Points) and Process Memorandum 108 (Process Analysers).

2.2 CO₂ removal and EO recovery (PFD No. : Unit 200-1/2,2/2)

Refer to the process flow diagrams for EO recovery and CO₂ removal. The CO₂ removal section has been designed based on the Shell hot potassium carbonate process and is suitable for HA catalyst operation. Performance of the CO₂ section will improve when special activators are applied. In particular, it is recommended to apply activators when very low CO₂ concentrations are warranted in the recycle gasloop, such as is preferred when the plant is operating on HS catalyst. Activators can be applied without making significant modifications to the current design.

2.2.1 CO₂ absorption (PFD No. : Unit 200-1/2)

Part of the recycle gas is heated in the CO₂ absorber feed preheater E-2001 by the lean potassium carbonate solution and enters the CO₂ absorber, T-2001, at the bottom. The flow is controlled by FIC-20101 that gives the set point to PDV-20104. Lean carbonate solution passes through the CO₂ absorber feed preheater, E-2001 and enters the CO₂ absorber at the top. As gas flows upwards in counter current to the lean carbonate, CO₂ is absorbed and converted to KHCO₃ in a reaction with K₂CO₃ and water.

The CO₂ absorber overhead gas leaves the column at 95-100°C and is saturated with water. Water is condensed by cooling the overhead gas in the CO₂ absorber overhead gas cooler E-2002A/B, where the gas is cooled to below 45°C. Water must be removed because it affects catalyst performance. After condensate separation, gas exits from the top of the



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water knock out drum D-2001 and joins the bypassed recycle gas.

Condensate from the water knock out drum is discharged on level control, LV-20107 to the top of the carbonate flasher D-2002 to wash the flashed gas. The flashed gas is washed and directed to the residual gas compressor.

In case of high level in the water knock out drum eg. as a result of liquid carry-over from the CO₂ absorber, it can be discharged directly via the dump line through LV-20105 to the carbonate flasher bottom, D-2002. In case of low level the bottom valve of the water knock out drum LV-20108 will shut to prevent breakthrough of loop gas. The conductivity of the condensate analysed by AI-20101 is an indicator for the presence of carbonate in the knock out water.

With the absorbed CO₂ the fat carbonate leaves the bottom of the CO₂ absorber on level control, LIC-20101 to the EO converter, Z-2001 for conversion of the absorbed EO to MEG.

The absorption of CO₂ by the carbonate solution is followed by chemical reaction. At low temperatures, the absorption equilibrium is favorable, but the slow rate of reaction is controlling. As the temperature increases, the equilibrium shifts and the reaction rate increases.

The design CO₂ concentration in the loop gas is 4.5(5.5) / 1.0(2.0)%mole at the inlet of the reactor. Note the significantly lower CO₂ concentrations when operating the plant with HS catalyst. For such low CO₂ concentrations, the application of activators in the carbonate solution is required to enhance the reaction rate.

The CO₂ absorber is a column of 5900 mm internal diameter with 3 packed beds of 8820 mm each of SS Mellapak 250Y. A demister is provided at the top of the column to minimise absorbent carry-over. In case of salt deposits on the demister leading to increased pressure drop over the column, a boiler feed water wash is installed above the demister. This water wash is normally not in use. During demister washing with BFW, gas load should be reduced slightly, since demister washing might introduce disturbs in the flow.

Various contaminants can enter the system and lead to foaming of the solution. Foaming can cause flooding and lead to severe performance degradation and carry-over. Foaming must be checked frequently by a foam test.

A rising pressure drop over the CO₂ absorber detected by PDI-20101A is an indicator of foaming problems. Foaming problems may develop slowly but



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steadily, sometimes over a 24 hour period, and may therefore be very difficult to notice. Foaming may also be initiated very rapidly due to a sudden load increase.

In case of foaming, the gas load must be reduced and if necessary stopped completely to resolve the problem. When the gas feed to the CO₂ absorber will be shut-down, the CO₂ concentration in the loop gas starts to rise rapidly and within minutes will require adjustment of the coolant temperature to maintain conversion to keep oxygen concentration under control. If the gas feed is not restored within minutes, oxygen shut-down is inevitable.

Injection of antifoam in the fat carbonate is required to suppress foaming. Antifoam is continuously fed into the system upstream E-2010A/B/C/D by means of P-2007A/B/C/D head I. In case of sudden increase of the foaming it has been foreseen the possibility to charge batch loads of antifoam directly into D-2006 by means of the batch antifoam filling system, pressurized after the charge with nitrogen from nitrogen network.

Proper operation of knock out drum D-2001 is required to prevent water carry over into the reactor catalyst tubes. Carry over of carbonate to the EO reactor permanently deactivates the catalyst and would require exchange of the catalyst.

To prevent a carry over from occurring the CO₂ removal unit is protected by the CO₂ removal unit shut-down system. Refer to Process Memorandum 112 (CO₂ Removal Shut-down System).

2.2.2 Carbonate flashing and stripping (PFD No. : Unit 200-2/2)

The fat carbonate passes through the EO converter Z-2001 where any absorbed EO from the CO₂ absorber is converted to MEG in order to prevent emission of EO with the CO₂ stripper overhead stream. Thus, glycols will build up to a steady-state level in the carbonate solution, with small amounts of glycol leaving with the CO₂ stripper overhead system.

After the EO converter, the fat carbonate is let down in pressure to 2.5 bara and flashes in the carbonate flasher D-2002, which consists of a water wash section in the top and a stripping section below the feed stage where make-up methane is injected to aid stripping of dissolved light ends from the fat carbonate. The methane injection rate is governed by the required methane concentration in the gas loop, not by the stripping requirement of the fat carbonate. Methane is injected in the carbonate flasher to reduce the ethylene content of the CO₂ vent. In case the supply pressure of the methane drops, it can also be injected directly into the inlet stream to the residual gas compressor suction knockout drum D-3002.



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Water for the water wash is provided from the water knock out drum, D-2001. Direct cooling with water minimizes any carbonate carry-over and reduces the flashed gas temperature for compression in the residual gas compressor.

The pressure of carbonate flasher is controlled by PIC-20310 on gas outlet line to the residual gas compressor, C-3001.

The fat carbonate leaves the bottom of the carbonate flasher on level control, LV-20306 and enters the top of the CO₂ stripper, T-2002. The solution will flash upon entering the CO₂ stripper and vapour and liquid are separated in the inlet device of the CO₂ stripper. CO₂ is stripped out of the fat carbonate by steam. This steam is partly generated in CO₂ stripper reboiler E-2004 and partly injected as open steam through two steam jet ejectors EJ-2001A/B. The open steam injection is required to make up for water lost in the CO₂ stripper overheads and is also used to drive CO₂ stripper jet EJ-2001A/B. The steam jets reduce the operating pressure of the vacuum flasher D-2006.

CO₂ stripper reboiler, E-2004 supplies heat required for reboiling of bottom product by use of PS(5K). the heat is controlled by FIC-20307.

Stripped CO₂ and steam together with traces of impurities leave the top of the CO₂ stripper. The stripper overhead is cooled in the CO₂ stripper overhead condenser AE-2003 and passes through the CO₂ vent knock out drum D-2005. The CO₂ vent knock out drum D-2005 has a special demister design to accomodate large variations in CO₂ production whilst maintaining adequate demisting. Water condensate is separated and partly recycled to the CO₂ stripper that is controlled by FIC-20314 to maintain the water balance and the remainder is discharged to the waste water system by condensate pump, P-2003A/B. The CO₂ gas is routed offsite as product. Any excess CO₂ production may alternatively be vented to a safe location. Pressure of CO₂ vent KO drum is controlled by FIC-20403 that takes the set point from PT-20402.

The lean carbonate solution from the bottom of the CO₂ stripper is flashed in vacuum flasher D-2006 that operates at lower pressure than the CO₂ stripper. Flashed CO₂ and steam are pressurised to the CO₂ stripper via CO₂ stripper jet EJ-2001 A/B. This flashing step enhances CO₂ removal from the lean solution and reduces the temperature. The flashed lean carbonate solution is subsequently pumped by carbonate pump, P-2001A/B via CO₂ absorber feed preheater E-2001 on flow control FIC-20304 to the CO₂ absorber, completing the cycle. A small slipstream of lean carbonate is continuously filtered in FT-2001 and recycled to vacuum flasher in order to remove suspended material.

The CO₂ stripper T-2002 has an internal diameter of 5000 mm and two packed sections of 7980 mm each of SS Mellapak 250Y.



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2.2.3 Carbonate storage and Carbonate drain header (PFD No. Unit 200-1/2)

Carbonate is supplied in bags and the solution has to be made up in the carbonate dissolving drum, D-2010. The strength of the carbonate solution should be about 20%. Prepared carbonate is pumped by carbonate make up pump, P-2002 to carbonate storage tank, TK-2001. Carbonate tank is equipped with internal heating coil for freezing protection. Make up carbonate is sent to vacuum flasher by carbonate make up pump, P-2002 on manual control with regards to FI-20306.

All carbonate drains are sent to carbonate drain drum, D-2011 via carbonate drain header to collect and recover carbonate drains. Collected carbonate drains is sent by carbonate drain pump, P-2011 to carbonate storage tank via make up carbonate filter, FT-2002 to remove suspended materials.

2.2.4 Control of CO₂ removal system (PFD No. : Unit 200-1/2)

The CO₂ concentration of the reactor feed gas must be controlled to maintain the design concentrations of CO₂ in the loop gas. The reactor feed is continuously analysed for CO₂ by the analyser in the reactor feed. For HA

catalyst operation, the optimum CO₂ concentration is 4.5(5.5)%mole. HS catalyst benefits from low CO₂ concentrations optimized at 1.0(2.0)%mole.

The CO₂ concentration can be decreased in a number of ways, amongst others by increasing the flow of recycle gas to the CO₂ absorber, the carbonate circulation rate to the CO₂ absorber and increasing the heat input to the CO₂ stripper. During the course of the catalyst lifecycle, the concentration of the carbonate solution is increased to increase the CO₂ absorption capacity. Operating conditions of CO₂ removal unit will need to be adjusted based on actual operation.

In general the flow of recycle gas to the absorber and the carbonate circulation are kept close to the design, while the heat input to the stripper is held to the minimum value required to maintain the desired CO₂ concentration in the reactor feed. Carbonate circulation adjustments should be made in small steps, of around 5-10 t/h. At least one hour should be allowed between changes for the system to reach a steady state.

The carbonate concentration is controlled by the amount of make-up water added as open steam to the system. The water recycle from D-2005 can also be used to control the carbonate solution concentration. However, if the



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build-up of glycols would be excessive, the water recycle has to be reduced and more open steam and less reboiler steam are required instead.

The vacuum in the vacuum flasher can be adjusted by the open steam injection rate through EJ-2001A/B by FV-20308 and FV-20309 or by the butterfly valve in the outlet of the vacuum compartment. The open steam flow cannot be set in isolation since it influences the carbonate concentration. The vacuum influences the temperature of the lean carbonate solution going to the CO₂ absorber.

The temperature in the CO₂ absorber affects the absorption in two ways: (1) a low temperature favors the equilibrium for CO₂ absorption and (2) high temperatures increase the absorption rate by enhancing the reaction rate. Exchanging heat between the lean carbonate solution and the feed gas to the CO₂ absorber in the feed preheater E-2001 establishes a more uniform temperature profile across the CO₂ absorber. In addition, a lower top temperature of the CO₂ absorber will reduce humidification of the overhead gas stream. A bypass valve on the lean carbonate side of the CO₂ absorber feed preheater, E-2001 can be used to control the absorption temperature. Antifoam must be injected continuously through a metering pump, P-2007 II A/B to the liquid outlet from vacuum flasher. The injection rate depends on the performance of the carbonate solution in the foam test.

A batch antifoam filling line with a hopper is also provided to add antifoam directly to vacuum flasher, in case of sudden increase of the foaming.

The extent of CO₂ removal is a function of the following process parameters:

- gas feed
- carbonate circulation rate
- total alkalinity
- solution conversion
- solution temperature

Gas feed

The gas feed puts a maximum to the CO₂ that could be removed. When insufficient gas is fed to the unit it is impossible to reach the required CO₂ concentration in the reactor feed. It is recommended to keep the gas load near the design since it will provide sufficient driving force, as the gas stream will not have to be depleted completely. Because of the lower partial pressure of CO₂ and thus a smaller driving force, there is a slightly higher gas flow rate to the CO₂ absorber in HS catalyst operation. It should be noted that increasing the gas flow rate to the CO₂ absorber will also increase the water concentration at the reactor inlet, which has an adverse effect on



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the catalyst. The gas feed to the CO₂ absorber should not exceed the maximum design flow rate for risk of flooding the column.

Carbonate circulation rate

Over-circulation will lead to poor stripping in the CO₂ stripper. This will result in reduced conversion of KHCO₃ back to K₂CO₃ and this will increase the CO₂ slip due to the absorption pinch at the top of the CO₂ absorber.

Under-circulation will decrease the CO₂ removal capacity in the bottom of the CO₂ absorber and force the fat carbonate to approach equilibrium due to absorption pinch in the bottom of the CO₂ absorber.

Always increase the circulation rate prior to an increase of the gas feed.

Alkalinity

Total alkalinity is a measure of solution strength and can be expressed as equivalent K₂CO₃ wt%. Under the same circulation, a higher alkalinity will increase the solution capacity. However, a higher alkalinity can increase the solution equilibrium CO₂ pressure and affect clean-up. Therefore, when the plant is operated at HS catalyst and the recycle gas loop already operates at low CO₂ partial pressure, the solution alkalinity should be reduced. High alkalinity will lead to crystallization if the fat carbonate is allowed to cool down. Excessive concentrations tend to make the solution more erosive and corrosive. Build-up of glycols in the solution will suppress the bicarbonate solubility.

The lean carbonate specific gravity is a crude measure of solution strength.

Alkalinity affects the CO₂ removal capacity in the same way as carbonate circulation. Slightly low alkalinity can be made up for by increased carbonate circulation.

In general, heat-stable salts do not interfere with CO₂ removal. Build-up of heat-stable salts increases the solution precipitation temperature and thus limits the CO₂ removal capacity of the solution. Generally, as heat-stable salts accumulate in the solution, the solution specific gravity will increase.

Solution conversion

Solution conversion is a measure of the CO₂ loading of the solution. It is defined as the percentage of K₂CO₃ being converted to KHCO₃. Thus, pure K₂CO₃ and KHCO₃ solutions will have conversions of 0% and 100%, respectively.

Solution conversion is an indication of the effectiveness of stripping in the CO₂ stripper. High solution conversion, usually due to insufficient stripping or too low circulation will lead to the equilibrium conversion being reached in the



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CO₂ absorber. Low solution conversion indicates that excessive steam is being used in the CO₂ stripper.

The vacuum flasher is instrumental in reducing the conversion of the lean carbonate solution.

Solution temperature

As the temperature is raised, the rate of absorption of CO₂ into the solution is accelerated. At the same time, the equilibrium vapour pressure of CO₂ over the carbonate increases, which decreases the driving force for absorption.

Increasing carbonate temperature to the CO₂ absorber will increase the residual CO₂ concentration in the treated gas (CO₂ slip) if absorption at that temperature would be controlled by equilibrium and decrease the CO₂ concentration in the treated gas if the absorption at that temperature would be controlled by the reaction.

It should be noted that the solution temperature can not be controlled independently as it is the result of gas feed, circulation rate and pressure in the vacuum flasher.

2.2.5 EO absorption (PFD No. : Unit 200-2/2)

The reactor product gas cools down in the primary product cooler, E-1002 and feed product exchanger, E-1001 to about 90°C. Residual gas from C-3001 is added and then cooled further to about 56°C in the secondary product cooler E-2005 against the fat absorbent from the EO absorber T-2003. The fat absorbent warms up from 40°C to about 50°C.

The cooled reactor product gas enters the quench section of the EO absorber T-2003. The quench circulation absorbs formaldehydes, light organic acids and decomposition products of the moderator. The pH of the quench liquid is controlled by caustic injection to upstream of quench pump, P-2004A/B by caustic metering pump, P-2006 I A/B.

The quench liquid leaves the column bottom at 49°C. A bleed stream is discharged to quench bleed stripper, T-2005 on flow control, FIC-20601, to remove the water condensed during cooling of the reactor product gas. FIC-20601 takes its set point from quench section level LIC-20601. The quench liquid circulates over quench cooler E-2006 which cools it to 40°C before it re-enters the quench section of the EO absorber. The temperature of the quench water is controlled by TIC-20701 acts on cooling water outlet. The quench liquid circulation rate is 830 t/hr.



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The boiler feed water cooled by jacketed pipe is used as make up water for quench section and is added to quench liquid downstream of P-2004A/B on flow control, FIC-20703.

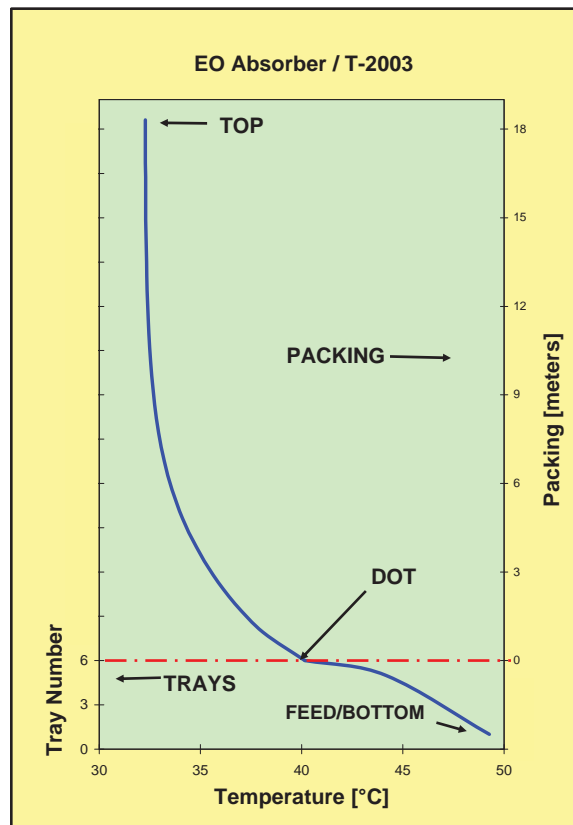


Figure 2.2.5-1: Temperature profile of EO absorber T-2003

The gas leaving the quench section is scrubbed with lean absorbent at 32°C in the EO absorber to recover EO. A small amount of caustic is continuously injected into the lean absorbent (to P-2005A/B suction) to control the pH at about 7.3-7.5 by caustic metering pump P-2006II A/B.

The absorbent rate is approximately 1600 t/hr and is based on 10 ppm EO in the absorber overhead gas. The EO content in the absorber overhead gas is monitored with an EO analyser which switches between this stream and the overhead gas of the residual absorber T-3002. The system is expected to be moderately foaming and therefore antifoam must be added to the lean absorbent upstream of E-2010A/B/C/D from the metering pump system, P-2007A/B that also provides antifoam for the CO₂ removal unit.

The fat absorbent stream leaves the column from the total draw off tray at about 40°C on flow control FIC-20603 that takes its set point from draw-off tray level, LIC-20604. A low level trip closes the fat absorbent outlet, SDV-



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20601 to prevent break through of loop gas to the EO stripper and from there via the relief valve to atmosphere. Also to prevent break through of loop gas to the EO stripper, the lean absorbent inlet is closed by a low differential pressure trip, PDT-20702 in case of back-flow via SDV-20701. Moreover, Low flow of the lean absorbent trips, with a delay of 30 seconds, the oxygen safeguarding system and the CO₂ removal unit.

The pressure of the recycle gas loop is controlled by purging a small fraction of the loop gas to remove inert from the system. This purge, the process vent, is removed from the loop gas at the recycle compressor suction knock out drum D-2004. Based on the design specifications for impurities in the oxygen and ethylene, the calculated process vent is approximately 1.4 t/hr. When the feed stock purities would be well within the specifications then the process vent rate will be much lower. In addition to the process vent, small purge losses may occur at the recycle compressor seals, sample connections and from analyser sample flows. This could reduce the required normal purge rate.

2.2.6 EO stripping (PFD No. : Unit 200-2/2)

The EO recovered in the EO absorber, T-2003 is stripped off the fat absorbent in the EO stripper T-2004.

The fat absorbent leaving the EO absorber is heated to 101°C in E-2007A~F before entering the top of the EO stripper, T-2004. The EO stripper overhead is condensed in AE-3001 and E-3002 and sent to the light ends column, T-3001, via the EO stripper tops surge drum, D-3001. The lean absorbent leaves the bottom of the EO stripper at 120°C and is recycled to the EO absorber after being cooled in E-2007A~F, E-3009, E-2010A~D and E-2011A/B to 32°C. The heat of the lean absorbent is recovered to heat up the fat absorbent in the EO stripper feed/bottoms exchanger E-2007A~F, and to deliver duty in the light ends column reboiler, E-3009.

Hydrolysis of EO occurs in the fat absorbent and EO stripper. The hydrolysis of EO to glycols is minimized by application of plate type heat exchangers to reduce the residence time in those parts of the fat absorbent system where temperatures are relatively high. For the same reason packing is applied in the top section of the EO stripper.



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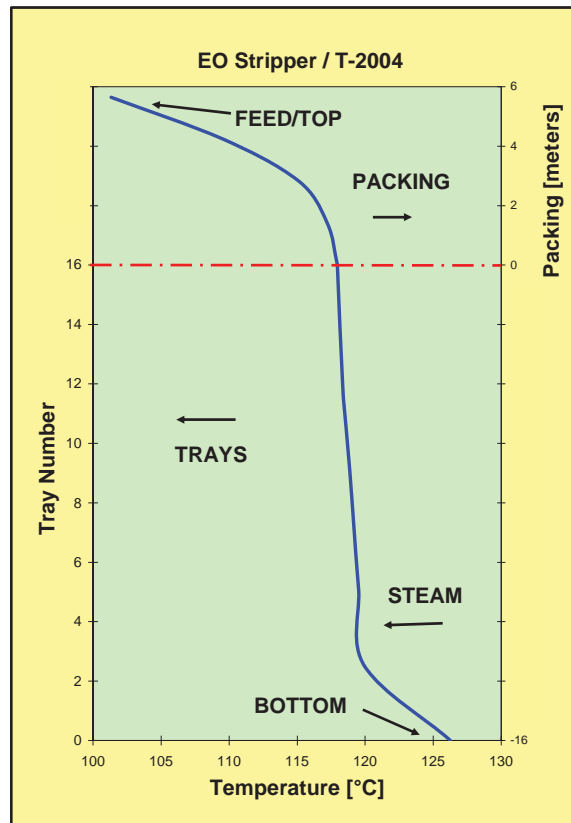


Figure 2.2.6-1: Temperature profile of EO stripper T-2004

Hydrolysis of EO to glycols may increase if a too low boil-up rate is applied and the concentrations of EO in the lower parts of the column are allowed to increase, or if the column pressure and consequently the column temperature profile is allowed to increase.

The overhead vapour contains about 60 wt % EO. Small amounts of hydrocarbon gases and CO₂, absorbed in the EO absorber, also pass overhead to the LE column.

The column has 12 HIFI calming section sieve trays and 5640 mm IMTP 50. The column diameter is 5450 mm.

The pressure controller, PIC-30401 on the suction knockout drum of the residual gas compressor, D-3002 controls the pressure in the EO stripper. If the pressure controller does not function properly, a vacuum may be created in the EO stripper overhead system by the compressor. Therefore nitrogen is introduced through a self-actuated pressure control valve, PCV-30107 on the stripper top surge drum, D-3001, to prevent vacuum and potential ingress of air into the system. For safety and integrity reasons, all equipment in the EO stripper overhead system shall be designed for full vacuum. Nitrogen inlet is closed by a differential pressure trip in case of back flow. Differential



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pressure trip activates two quick closing valves, SDV-30103 and SDV-30101. As soon as the quick closing valves are shut, an automatic bleed valve, SDV-30102, vent the line section between the quick closing valves.

The vapour flow needed to strip out the EO comes partly from the knock back section due to stripper reboiler, E-2008 that works by PS(5K) and partly from CS 2K stripping steam. The boil-up is essentially constant, with the reboiler steam flow controlled by the column top temperature, TIC-20908 to obtain the desired EO recovery by FV-20903. The liquid level in the bottom of the EO stripper controls the open steam flow. The total heat input is affected by the open steam injection rate. Therefore the reboiler steam flow is corrected via a calculator to keep the total heat duty constant.

2.2.7 Glycol bleed (PFD No. : Unit 200-2/2)

In order to remove sodium salts and glycols from the absorbent circulation, a relatively small bleed stream from tray 5 is sent to the EO stripper bottoms knock back section via P-2005A/B on level control, LIC-20901. The knock back section is located in the shell of the EO stripper itself. A concentrated glycol bleed leaves the knock back section and is sent to the glycol bleed flasher T-4005, via glycol bleed pump, P-2008A/B on flow control, FIC-20901.

The EO stripper bottoms knock back section has 4 calming section valve trays. Vapour from the top of the knock back section returns to the EO stripper below tray 5. A thermosyphon reboiler, E-2008, heated by PS 5K process steam, is used for boil-up.

The glycol bleed flow from the bottom of the knock back section is maintained reasonably constant to maintain the glycols concentration in the lean absorbent constant. The pH of the glycol bleed residue from the glycol bleed flasher is controlled by caustic injection in the feed stream to the knock back section.

The concentration of glycols in the circulating absorbent can be reduced by increasing the glycol bleed rate causing an import of clean condensate controlled by FIC-20906 from the steam condensate flash drum D-4001 to the lean absorbent. An increased glycol bleed rate will also result in a decrease of the glycol concentration in the bleed stream.

In case the circulation of lean absorbent is stopped, the absorbent liquid should be kept on the EO absorber draw off tray and in the EO stripper bottom. The level on EO absorber draw off tray should not be lowered to avoid overfilling of EO stripper bottom.



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2.2.8 Quench bleed processing (PFD No. : Unit 200-2/2)

In the quench section of the EO absorber, the reactor product gas is cooled and acidic components and formaldehyde are absorbed in the quench liquid. A bleed from the quench liquid is required to remove impurities and the water condensed from the reactor product gas. The EO dissolved in the quench bleed stream is recovered in quench bleed stripper T-2005 using CS 2K open steam as a stripping medium that controlled by FIC-21002.

The quench bleed stripper, T-2005 is a packed column of 700 mm internal diameter packed with two 4000 mm beds of stainless steel IMTP 25 random packing. The overhead is returned to the EO stripper below the packed section.

The stripped liquid leaving the column bottom contains water, glycols, organic sodium salts, formaldehyde and NaHCO_3 . This quench bleed residue is sent to the quench bleed flasher, T-2006 on flow control FIC-21001 that takes its set point from LIC-21001, from which the quench bleed residue (bottom product) is sent to disposal on level control LIC-21004 via heavy glycols cooler E-5014 and heavy glycols tank TK-5002. Overheads are condensed and cooled to 65°C by quench bleed flasher condenser, AE-2013 and discharged to waste water treatment and condensate vessel D-5003.

A reboiler, E-2012 heated by PS(5K) is used for boil-up. Input heat is controlled by steam flow that takes its set point from bottom temperature of the column, TIC-21002. The column pressure is controlled by PIC-21008 on overhead product stream.

Quench bleed flasher has 6 calming section sieve trays. The column diameter is 1800mm.

Caustic is continuously added to the feed of the quench bleed flasher to control the downstream pH at about 7.5 – 8.0.



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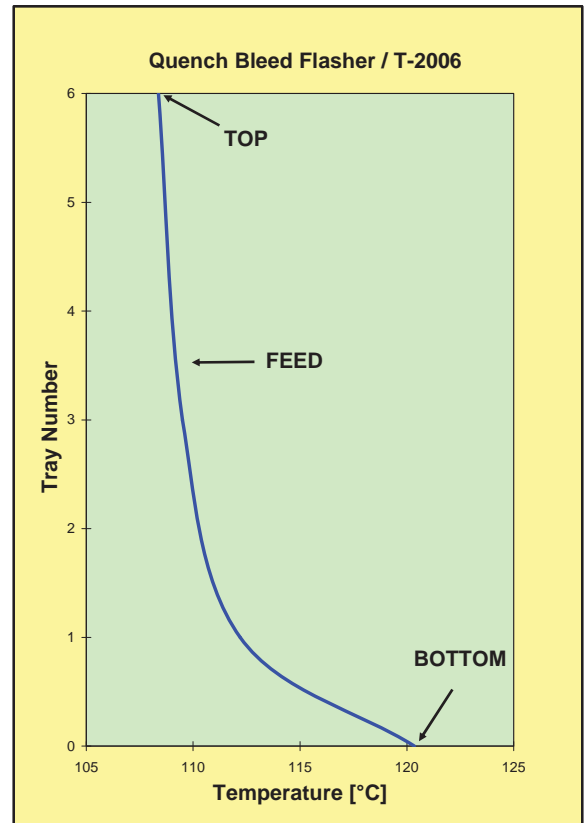
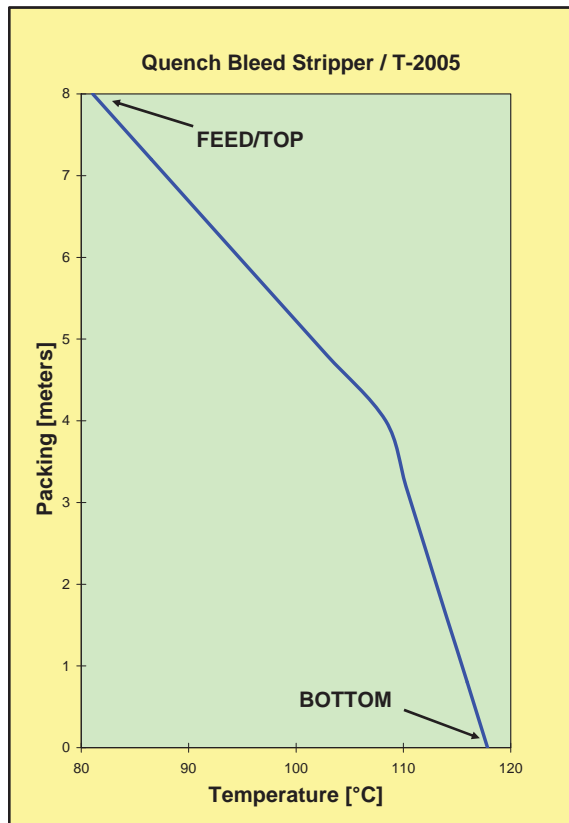


Figure 2.2.8-1: Temperature profile of quench bleed stripper T-2005

Figure 2.2.8-2: Temperature profile of quench bleed flasher T-2006

2.2.9 Lean absorbent loop (PFD No. : Unit 200-2/2 and Unit 300-1/1)

In the lean absorbent loop, the fat absorbent from the EO absorber is heated in the secondary product cooler E-2005 from 40°C to about 50°C by the reactor product gas. Thereafter it is heated in the EO stripper feed/bottoms exchangers E-2007A~F to 101°C before entering the EO stripper T-2004. In order to minimise the hydrolysis of the EO to glycols a plate type heat exchanger is applied for low residence time.

The lean absorbent from the EO stripper bottom is cooled in the EO stripper feed/bottoms exchanger E-2007, from 120°C to about 68°C. Thereafter, the lean absorbent is used to provide heat to the light ends column reboiler, E-3009. Finally, the stream is further cooled to 37°C in the lean absorbent cooler, E-2010, and is then chilled to 32°C in the lean absorbent chiller, E-2011. Temperature of the lean absorbent is controlled by TIC-20705 on chilling water outlet. This cold lean absorbent is then used as absorbent in the EO absorber, T-2003 and the residual absorber column, T-3002. Lean absorbent enters from the top of EO absorber on flow control, FIC-20702.



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The chilling of the lean absorbent is achieved by chilled water from the package unit chilled water refrigeration PK-6050A~F.

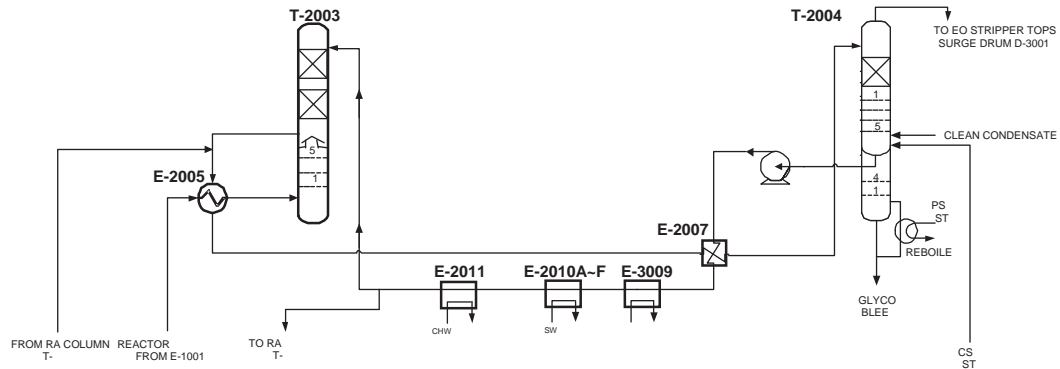


Figure 2.2.9-1: Lean absorbent loop

2.2.10 Recycle gas compression (PFD No. : Unit 200-1/2)

The recycle gas compressor, C-2001 compresses the recycle loop gas from the EO absorber overheads.

Upstream of the recycle gas compressor the gas passes through the suction knock out drum D-2004. In case of high liquid level in this vessel, LT-20202 activates the oxygen feed shutdown and after a 5 second delay the recycle compressor will stop.

A small slipstream is purged from the outlet of D-2004 in order to bleed inerts via PV-20201A. This process vent gas can be used as fuel gas. An emergency hand valve, HV-20201 is available to depressurise the recycle gas loop quickly by discharge to the MEG plant flare.

A fraction of the recycle gas from the compressor discharge is sent to the CO₂ removal section for CO₂ removal. The treated gas leaving the CO₂ removal unit recombines with the recycle gas and enters the EO reaction section. Low differential pressure across the compressor as well as suction high pressure will activate oxygen shutdown system and will trip the compressor. The compressor vendor will specify the required trip initiators for the compressor safeguarding system.

The centrifugal compressor may surge when the flow rate is reduced. In order to avoid surging a minimum flow controller, FIC-10301A in the recycle gas feed to the oxygen mixing nozzle can open a bypass valve, FV-10301 in the reactor feed gas



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line between the oxygen mixing nozzle and the reactor feed product exchanger to obtain a bypass around the EO reactor.

This also ensures full recycle gas flow through the oxygen mixing nozzle to provide safe mixing conditions while allowing control of the gas flow through the reactor.

A low flow trip, FALL-10301B on the gas flow to the oxygen mixing nozzle shuts off the oxygen supply. Only in the case of a compressor trip by high axial displacement or loss of sealing will the compressor be tripped immediately without a 5 second delay. In all other cases the oxygen will be shut down immediately, and after a 5 second delay the compressor will be tripped.

Two motor operated valves MOV-20201 and MOV-20202 are considered for isolating of the compressor.

2.2.11 Chemical Feed

Caustic is used to control the alkalinity of lean absorbent and quench liquid. Caustic 50%wt is supplied from battery limit and diluted with demineralized water in a static mixer, AZ-2111 to provide caustic 20%. Demineralized water exchanger, E-2015 has been considered to maintain temperature of dilution water between 27°C to 38°C. DMW flow is controlled by FIC-21101 that is set by a ratio calculator based on caustic flow by FIC-21102. After static mixer AZ-2111, a jacketed pipe has been considered to remove heat of dilution and prevent heat stress cracking. Caustic 20% is stored in caustic tank, TK-2002. Caustic is sent to process unit by caustic metering pumps, P-2006A~D (A/C in service and B/D stand-by).

Antifoam is used to control the foaming of CO₂ absorber and EO absorber. Antifoam is supplied by container and discharged to antifoam vessel, D-2003 via antifoam loading pump, P-2012. Antifoam is sent to process unit by antifoam pump P-2007A/B that includes two heads, head I for EO absorber and head II for CO₂ absorber.

2.3 LE removal section (PFD No. : Unit 300-1/1)

Refer to the process flow diagrams for LE removal. Refer to section 8.8 describing specific requirements for equipment in EO service.

2.3.1 Light ends removal (PFD No. : Unit 300-1/1)

The LE removal system removes CO₂ and light ends that are absorbed in the EO absorber and stripped out in the EO stripper. Light ends must be



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removed to ensure normal operation of the glycol reaction and to minimise corrosion due to CO₂.

The overhead of the EO stripper, which contain about 60 wt % EO is cooled to 40°C and partly condensed in the EO stripper tops cooler, AE-3001 and the EO stripper tops condenser, E-3002.

The vapour fraction consists mainly of CO₂, ethylene and EO. By chilling to 13°C, after combining with the light ends column overhead, in the stripper tops vent chiller E-3005, most of the EO is recovered and returned as aqueous EO to the EO stripper tops surge drum D-3001. Temperature of vapor outlet from E-3005 is controlled by chilling water outlet rate, TIC-30107. The aqueous EO stream must be kept above the freezing point curve to avoid the formation of EO hydrates.

The remaining vapour contains about >80wt% EO. It is therefore fed to the residual absorber T-3002 for recovery of EO by absorption with lean absorbent.

The condensate from the EO stripper tops condenser, E-3002 enters the stripper tops surge drum, D-3001 at 40°C through a perforated dip pipe. The light ends

column is fed from the stripper tops surge drum. The minimum flow recycle of the light ends column feed pump, P-3001A/B is returned to the EO stripper tops surge drum through an eductor to prevent stratification and keep the liquid inventory well-mixed.



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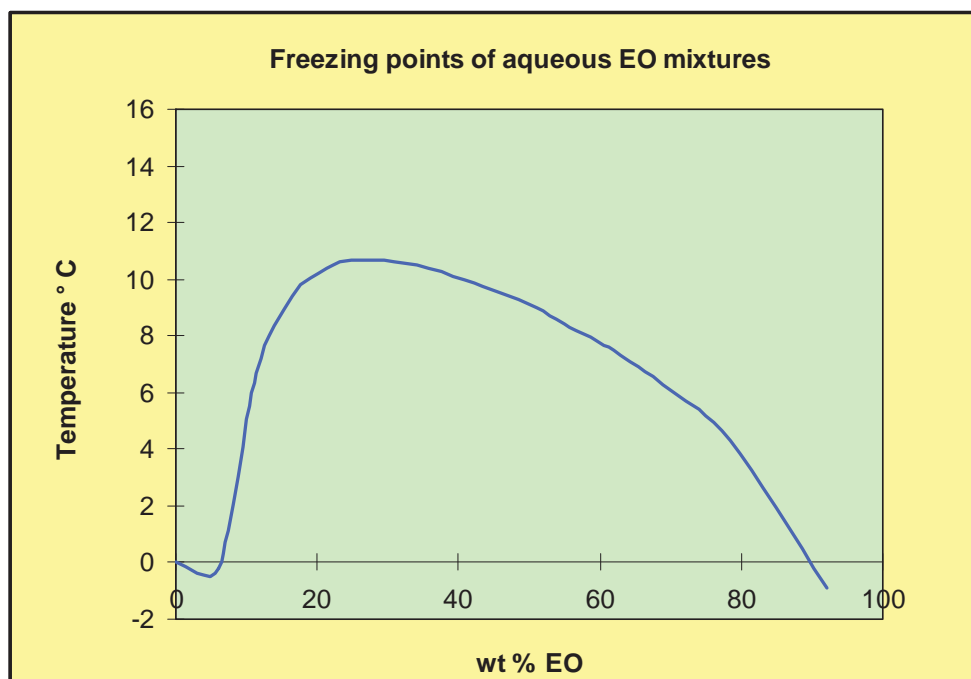


Figure 2.3.1-1: Freezing points of aqueous EO mixtures

The light ends column bottom temperature is 24°C at about 1.5 bara. This allows the use of hot lean absorbent as heating medium for the light ends column reboiler E-3009. During start-up low pressure steam could be injected into the column bottom in case the thermosyphon reboiler does not operate properly.

Lean absorbent flow rate and open steam flow rate are controlled by FIC-30204. The surplus of lean absorbent that not entered to E-3009 is returned via bypass valve, PDV-30204. In case of light end column shutdown, hot lean carbonate to E-3009 and open steam to LE column are shut off by FV-30204B and FV-30204A respectively and the recycle flow via E-3004 is completely opened by FV-30202. For more details refer to "cause and effect chart for interlock and ESD system", ESS-IN-009, LE column shutdown interlock, I-305.

Too low temperatures will lead to inadequate removal of CO₂ resulting in severe corrosion problems in the downstream sections.

The aqueous EO, containing about 55-60wt% EO, proceeds from the bottom of the light ends column to the glycol reaction section.



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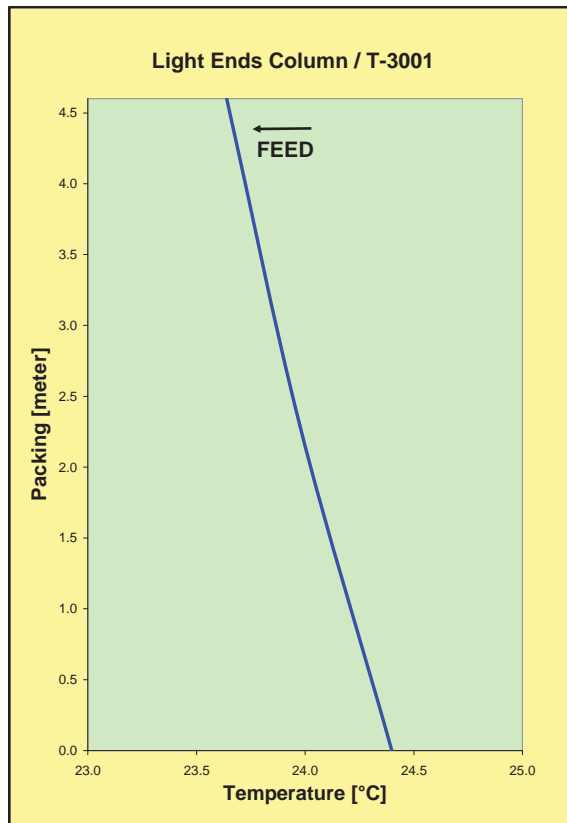


Figure 2.3.1-2: Temperature profile of light ends column T-3001

In case of an unscheduled shut-down of the glycol section leading to a situation where aqueous EO remains in the EO stripper tops surge drum, D-3001 and light ends column system, the inventory must be kept in circulation over the aqueous EO shutdown cooler, E-3004 to cool the mixture and to prevent it from running away. Circulation flow is provided by the light ends column feed pump P-3001A/B and light ends column bottoms pump P-3002A/B. Both pumps will run the A and B pumps in parallel in this case and are equipped with emergency power supply in case of a trip due to a power failure. In time the aqueous EO will convert to glycols. The EO stripper tops surge drum and light ends column should not be used as a storage system for aqueous EO. It should only be used in this manner in case of unscheduled shut-down of the glycol section. In case of cooling water failure, fire water must be connected as cooling medium for the aqueous EO shutdown cooler E-3004.

Proper control of and attention to the level in the bottom of the LE column is critical to the smooth operation of the plant since an excursion to either side of the level will almost certainly lead to a total shutdown of the plant.



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For more information refer to the glycol reactor shutdown system, memo113 and 114.

2.3.2 Residual EO absorber (PFD No. : Unit 300-1/1)

The vapour that is not condensed in the stripper tops vent chiller E-3005 is contacted with lean absorbent in the residual absorber T-3002 to recover EO. The lean absorbent rate that comes from E-2011 is controlled by FIC-30301A. The EO rich absorbent bottoms are pumped by P-3003A/B and mixed with fat absorbent from the EO absorber and pumped back to the EO stripper, T-2004. The minimum flow recycle of P-3003A/B is returned to T-3002 on FIC-30302. The net bottom product from T-3002 to E-2005 is controlled by the level of column bottom LIC-30301.

The overhead gas of the residual absorber column is recycled via the residual gas compressor, C-3001, to the EO absorber, T-2003. The bottom temperature of the residual absorber is 41°C at a pressure of 1.4 bara.

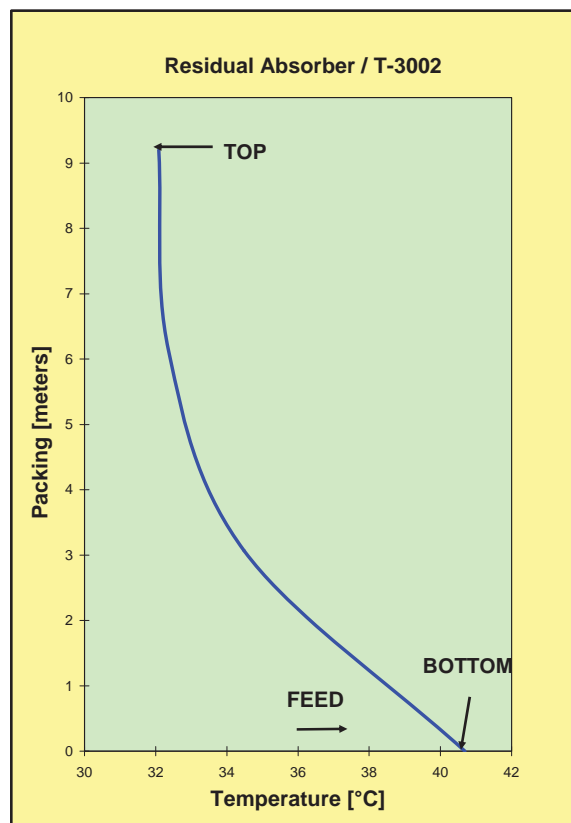


Figure 2.3.2-1: Temperature profile of Residual absorber T-3002

The design of the residual absorber, T-3002 is based on the recovery of almost all EO with <0.1% EO in the overhead stream of the column to



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minimise EO recycle in the residual gas. The residual absorber, with a diameter of 1900 mm, is provided with 2 packed beds, each of 4600 mm IMTP 40.

2.3.3 Residual gas compressor (PFD No. : Unit 300-1/1)

The residual absorber overhead T-3002, the carbonate flasher overhead D-2002 and the gas from the fast loop are compressed by the three stage reciprocating residual gas compressor C-3001 and returned to the recycle loop to reduce losses of ethylene and methane.

The carbonate flasher overhead and the gas from the fast loop are cooled in the residual compressor recycle cooler E-3006, prior to the addition of the residual absorber overhead. Any liquid condensed is knocked out in the residual compressor suction knock-out drum D-3002 and sent to disposal. The liquid level on D-3002 is controlled by LIC-30401. An on/off valve LV-30402 prevents the back through of residual gas to atmosphere during low low liquid level in D-3002.

The liquid condensed in the residual compressor interstage cooler-1, EC-3001-1, and Cooler-2, EC-3001-2, are separated in the residual compressor interstage knock-out drum 1, DC-3001-1, and drum 2, DC-3001-2, and sent to the residual absorber sump. For more detail refer to C-3001 instruction documents.

The residual gas compressor shut-down system is activated under the following conditions:

- low absorbent flow to the residual EO absorber
- high vapour flow at the outlet of the vent chiller, E-3005
- low suction pressure
- high liquid level in the suction knock-out drum
- high liquid level in one of the interstage knock-out drums
- abnormal conditions specified by compressor vendor
- manual intervention by the operator

The shut-down system also closes the shut down valve in the suction line. Refer to Process Memorandum 114 Sec.2 (Residual Gas Compressor Safe Guarding System) and ESS-IN-009, cause & effect chart for Interlock and ESD system.

2.4 Glycol reaction and recovery (PFD No. : Unit 400-1/2,2/2)

Refer to the process flow diagram for glycol reaction and recovery.



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Refer to the section 8.8 describing specific requirements for equipment in EO service.

2.4.1 Glycol reaction (PFD No. : Unit 400-1/2)

The glycol reaction takes place in the liquid phase. Process condensate from D-4002 is added to the aqueous EO stream (55%wt) from the light ends column bottom and sent to glycol reactor via E-4001, E-4002 and E-4003. The rate of condensate is controlled by FIC-40106A that takes the set point from FIC-40102A and HIC-40103. To set the proper ratio water/EO at glycol reactor inlet, the controller HIC-40103 is provided.

The reaction already starts off at this mixing point and the subsequent reactor preheaters, though it is still very slow at lower temperatures. A plug flow reactor provides the required residence time to complete the hydrolysis of EO. In the glycol reactor R-4001 the reaction proceeds adiabatically. At a reactor inlet temperature of 190°C and a residence time of more than 3 minutes more than 99.999% of the EO will be hydrolysed. The reactor inlet temperature of 190°C is attained by using preheaters E-4001, E-4002 and E-4003. the heating rate is controlled by FIC-40115, LIC-40305 and FIC-40109 (that takes the set point from TIC-40108).

The glycol forming reactions are exothermic. Thus the product containing excess water and glycols exits the reactor at about 230°C.



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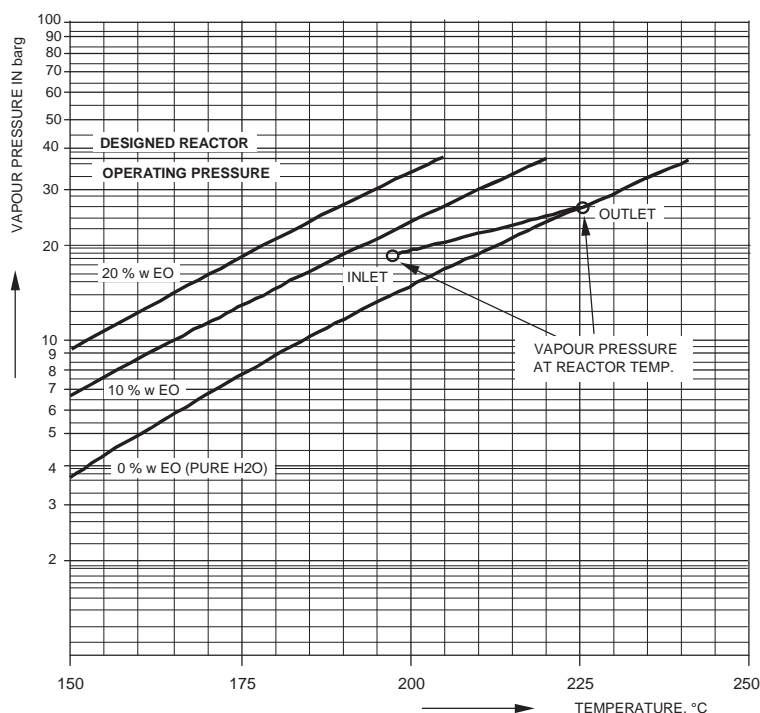


Figure 2.4.1-1: Minimum operating pressure for glycol reactor

The pressure, necessary to keep the reactor contents in the liquid phase, is shown as a function of temperature and EO concentration. With a residence time in the reactor of three minutes, the conversion of EO is essentially one hundred percent. Formation of a vapour-phase in the reactor is to be avoided, since EO in the vapour would pass through the reactor essentially unreacted.

The basic requirement for the EO reaction is a complete conversion of EO into glycols. This requirement is realized by an excess water flow which also provides heat sink for the exothermic reaction and by maintaining the inlet temperature of the reactor at a certain value. A too low temperature will slow down the reaction, whilst at high temperatures part of the EO will vaporize and pass through the reactor unreacted. To ensure that the reaction takes place in the liquid phase, the reactor pressure is safeguarded.

An automatic shutdown system protects the glycol reactor system. Refer to Process Memorandum 113 (Glycol Reactor Shut Down System).

The reactor consists of a 20" pipe with a total tube length of 150 m. Reaction to glycols already starts in the preheat train.



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A fully developed turbulent plug flow will prevail throughout the reactor avoiding back mixing and thus the undesired reaction to higher glycols.

The reactor product distribution is a function of the concentration of EO, MEG and higher glycols in the reactor feed.

If the EO concentration in the feed is reduced, the MEG fraction in the glycol product increases, but the quantity of water to be flashed off also increases. Recycling glycols would then increase the proportion of higher glycols produced and reduce the net MEG production.

The water/EO ratio is 9 (on a mass basis). This is the concentration just after the mixing point.

2.4.2 Glycol concentration (PFD No. : Unit 400-1/2)

The product stream leaving the glycol reactor that contains about 86 wt % water is sent to T-4001 on PIC-40203 to avoid vapor phase formation on reactor. The water is removed in glycol concentrator-1, T-4001, glycol concentrator-2, T-4002, glycol concentrator-3, T-4003, and glycol dehydrator T-4004.

The concentrator columns operate as a multiple effect evaporation unit. CS 20K steam heats glycol concentrator-1 reboiler E-4004A/B and the overhead vapour drives the reboiler of glycol concentrator-2, E-4005A/B. The overhead vapour of concentrator-2 drives the reboiler of the glycol concentrator-3, E-4006A/B. Process steam from the concentrator-3 is used in various reboilers in the plant.

Glycol concentrator-I reboilers levels are controlled by LIC-40205A/B that take the set point from FIC-40202A/B on CS20K steam inlet. Levels of glycol concentrator-II are controlled by LIC-40304A/B that take the set point from PDIC-40306. level of glycol concentrator-III are controlled by LIC-40403A/B that take the set point from PDIC-40405.

The net product from bottom of T-4001 to T-4002 is controlled by the level of the column bottom, LIC-40201. The net product from bottom of T-4002 to T-4003 is controlled by the level of the column bottom, LIC-40301 and the net product from bottom of T-4003 to T-4004 is filtered by concentrator-III bottoms filter, FT-4001A/B and controlled by FIC-40403 which takes the set point from the level of column bottom, LIC-40401.

The glycol concentrator-1 and 2 are equipped with 16 calming section trays each, whilst the last concentrator-3 has 18 calming section trays. Each of these columns is operated with a relatively small reflux in order to reduce the glycols concentration in the overhead streams. Clean condensate is used as



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the reflux and is supplied from the steam condensate flash drum D-4001 under normal operation. Provided that there is no negative impact on the product quality and plant performance, it is possible to recycle some process condensate from the water surge drum D-4002 via P-4003A/B, thereby reducing the amount of waste water produced in the overall process. Particular attention should be paid to monitoring product quality when the glycol dehydrator overhead condensate is also lined up to the water surge drum D-4002. The refluxes of towers are controlled by FIC-40203, FIC-40301 and FIC-40401.

To prevent foaming, chemical antifoam can be pumped to T-4001 by P-2007II A/B.



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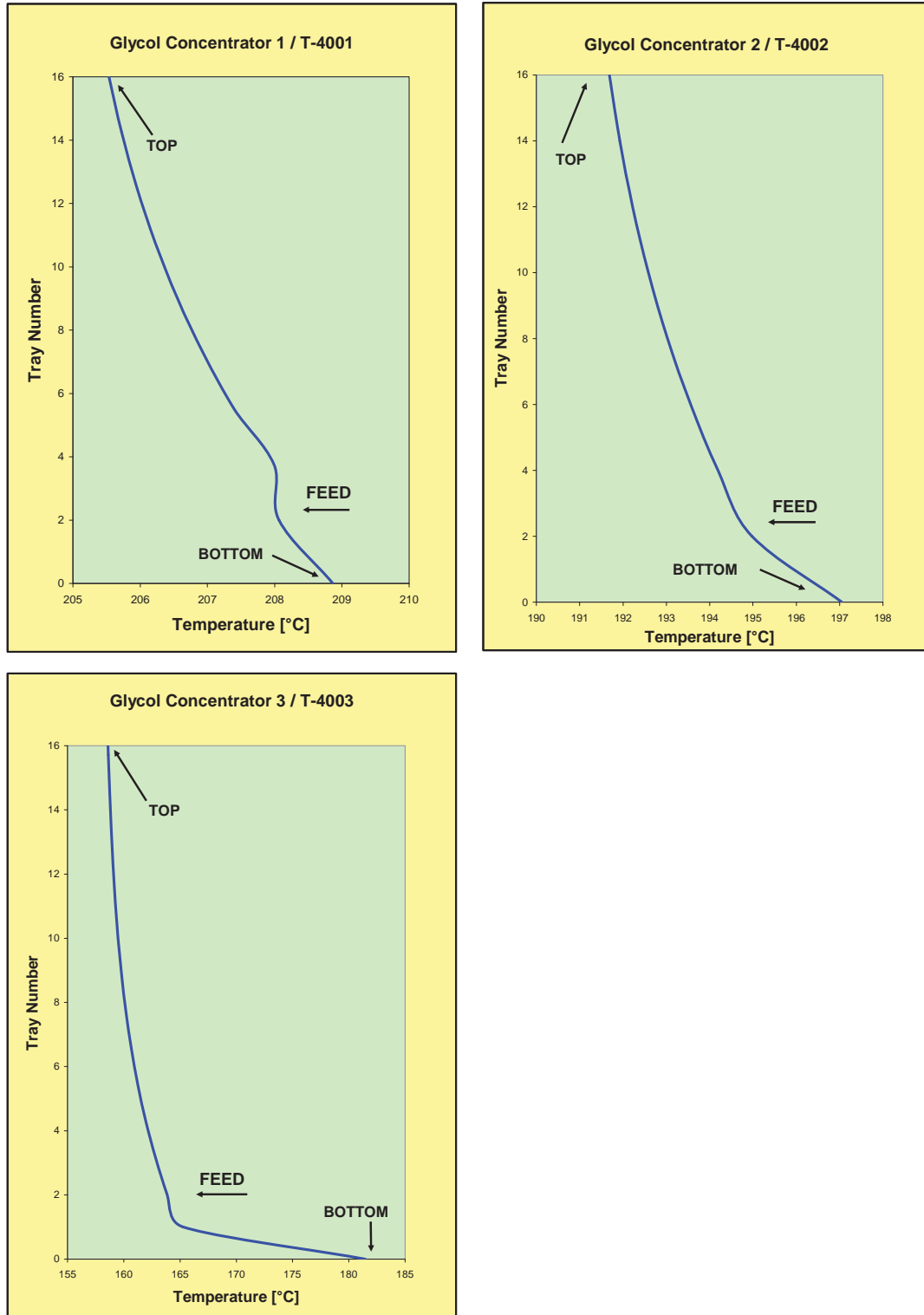


Figure 2.4.2-1: Temperature profiles of glycol concentrators-1/2/3



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2.4.3 Glycol dehydration (PFD No. : Unit 400-2/2)

Refer to the process flow diagram for glycol dehydration.

The bottom of the glycol concentrator-3, T-4003 contains about 32 wt% water, which is removed in glycol dehydrator T-4004. The glycol dehydrator operates under vacuum to remove the remaining water. During normal operation the crude glycols from the bottom of the dehydrator are pumped by glycol dehydrator bottoms pump, P-4009A/B and to the glycol purification section. The net bottom products are filtered by glycol dehydrator filters, FT-4002A/B and sent to MEG column T-5001 on FIC-40501 that takes the set point from LIC-40501. Besides, it is possible to send the bottom of the dehydrator as glycol slops through the slops cooler E-5015 to the glycol slops tank T-5001. The slops cooler E-5015 is designed to cool the glycol dehydrator bottom stream with a flow of 50% of the normal flow to 50°C.

The glycol dehydrator, T-4004 is a swaged column with a diameter of 5000mm in the top and 4000mm in the bottom with two packed sections of 4578 mm Mellapak Plus 252Y in the top and 3150 mm Mellapak 250Y in the bottom. Sufficient reflux is applied to ensure proper wetting of the packing in the top of the column via P-4006A/B from glycol dehydrator reflux drum, D-4004.

Apart from the glycol concentrator-3 bottoms, the MEG recycle from the MEG column T-5001 via P-5005A/B and P-5016A/B and MEG recycle column T-5002 via P-5007A/B and the liquid top product of the glycol bleed flasher T-4005 via P-4008A/B are fed to the dehydrator to remove light ends. These feeds are mixed in line upstream of the dehydrator column feed inlet device.

In the glycol dehydrator the recovery of glycols is completed and the water concentration in the crude glycol stream reduced to less than 250 mg/kg.

The top pressure of the dehydrator is 130 mbara with a temperature of 51°C. The overhead vapors are condensed by glycol dehydrator condenser E-4008A/B that uses cooling water as cooling media. The condensate from E-4008A/B are collected in glycol dehydrator reflux drum, D-4004 and pumped by glycol dehydrator reflux pump, P-4006A/B to wastewater collection basin, Z-7070 via wastewater trim cooler, E-4011 on level of reflux drum LIC-40601. Under normal operation the overhead is discharged to waste water. However, part of the overhead stream may be pumped into the water surge drum D-4002 as make-up as long as product quality is not affected, to reduce the amount of waste water. Particular attention should be paid to product quality monitoring and plant performance when in this case part of the process condensate from the water surge drum is also used as reflux for the glycol concentrators.



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The required vacuum pressure is supplied by vacuum system, PK-5001 and the column pressure is controlled by PV-40504 on nitrogen line.

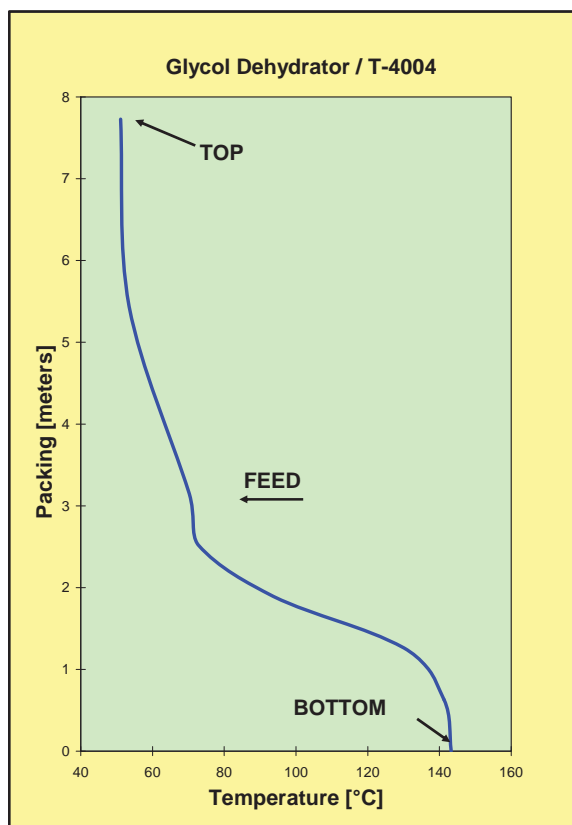


Figure 2.4.3-1: Temperature profile of glycol dehydrator T-4004

2.4.4 Glycol bleed recovery (PFD No. : Unit 400-2/2)

Refer to the process flow diagram for glycol dehydration.

The glycol bleed from the EO stripper bottoms, containing water, glycols and salts, is distilled under vacuum in glycol bleed flasher T-4005 to recover water and glycols, while disposing of the salts and a small amount of organic material.

The glycol bleed flasher is provided with 6 calming section valve trays in the top and 3 in the bottom. The liquid in the bottom section contains salts. A forced circulation, suppressed vaporization type reboiler, E-4009 is selected to avoid precipitation of salts. Heat is supplied to glycol bleed flasher column by MCS with flow control on the steam inlet, FIC-40703 that takes the set point from bottom temperature, TIC-40701.



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The operating pressure of the glycol bleed flasher top is 115 mbara with a top temperature of 101°C. Tray 10 acts as a condenser drum. The temperature on tray 10 is 50°C.

The overhead vapors are condensed by internal condenser, E-4010 that uses cooling water as cooling media. The internal condenser E-4010 is provided to minimize pressure drop.

The liquid overhead products are drawn off tray 10 and pumped by glycol bleed flasher reflux pump, P-4008A/B. The net product which is controlled by FIC-40704 that takes the set point from the level of draw-off tray, LIC-40703 sent to glycol dehydrator, T-4004 while the reflux flow is controlled by FIC-40706.

The required vacuum pressure is supplied by vacuum system, PK-5001. A nitrogen line is provided to compensate leak air in case of excess performance of vacuum system via PV-40704.

The unrecovered glycols and salts from the glycol bleed flasher form the glycol bleed residue. The glycol bleed residue is pumped by P-4007A/B and is combined with the TEG column bottom from T-5004 and quench bleed flasher bottom stream and from T-2006 cooled in the heavy glycols cooler, E-5014, before storage in heavy glycols tank TK-5002.

The glycol bleed flasher is designed for 40% additional capacity for slops processing.



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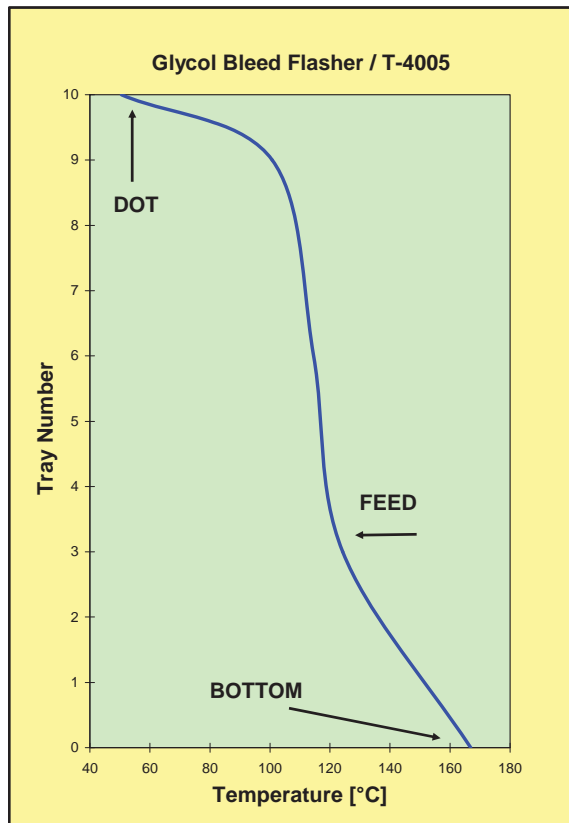


Figure 2.4.4-1: Temperature profile of glycol bleed flasher T-4005

2.5 Glycol purification (PFD No. : Unit 500-1/2,2/2)

Refer to the process flow diagrams for MEG purification and DEG and TEG purification.

In order to operate at temperatures low enough to prevent product degradation, the columns in the glycol purification section are operated under vacuum.

Since oxygen from air leakage may cause rust formation and henceforward could lead to formation of aldehydes the columns and their accessories must meet stringent air leakage criteria. Proper attention must be paid to the prevention of flange leakages.

Because of limitations on allowable pressure drop through the columns low pressure drop packing is applied, such as Sulzer BX and Mellapak 250 Y. The liquid hold up for these packing types is relatively low. Fluctuations in the gas load of the column will therefore lead to fluctuations in the product composition with a short response time.



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Falling film type reboilers are selected to minimize thermal degradation of the glycol product. This type of reboiler has following characteristics:

- low pressure drop
- low skin temperatures
- low residence time.

Filters are installed in the bottom pumps discharge for protection of the distributor systems in the falling film reboilers and to avoid fouling of the downstream column packing.

2.5.1 MEG purification (PFD Mo. : Unit 500-1/2)

Refer to the process flow diagram for MEG purification.

The purification of the MEG product is achieved in two steps: the MEG column, T-5001, and the MEG recycle column, T-5002.

Refer to the process flow diagram for MEG purification.

MEG column

The crude glycols from the glycol dehydrator, T-4004 enter the MEG column. The MEG product is drawn off as a side stream.

The column has a diameter of 4700 mm and contains three sections of Mellapak 250 Y. The total packed height is 7560 mm.

The overhead vapours at 151°C and 226 mbara are condensed in a steam generator E-5001 which operates under thermosyphon. Utility grade steam of CS 2K is generated using clean condensate from the steam condensate flash drum, D-4001.

To minimise losses of MEG in the vacuum system, a vent condenser, E-5002, using cooling water is installed in the vapour line from the MEG column reflux drum D-5002.

D-5002 has two parts which are separated by a dividing wall. Condensate from E-5001 enters one side and is pumped by MEG column overhead pump, P-5005A/B. the reflux which is controlled by FIC-50302 that takes the set point from LIC-50304 is returned to T-5001, while the net overhead product to T-4004 is controlled by FIC-50301. The condensate from E-5002 enters the other side of D-5002 and is pumped by P-5016 to T-4004. The flow rate is controlled by FIC-50304 that takes the set point from LIC-50306 on reflux drum.



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OPERATING MANUAL

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Traces of water and other light impurities may leave via the vacuum system.

The purified MEG is withdrawn underneath the pasteurising section by MEG column rundown pumps, P-5003A/B. and cooled to 50°C by MEG product coolers AE-5013 and MEG product cooler, E-5004. Then polished in MEG polishing vessel D-5001A/B and sent to MEG sales tanks storage.

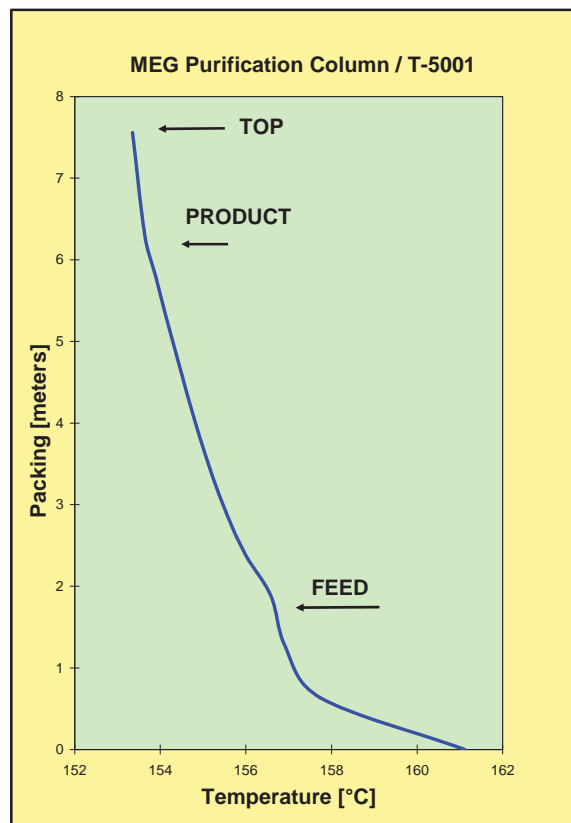


Figure 2.5.1-1: Temperature profile of MEG column T-5001

A MEG polishing bed in D-5001A/B removes traces of iron before the MEG product is sent to MEG product sales tanks (OBL). Under normal operation, both vessels D-5001A/B operate in parallel. Because there is no MEG rundown tank inside the battery limit, strict product quality monitoring is required. A gas chromatograph, AI-10913 and an UV analyser, AI-10916 are provided to determine MEG composition. If the MEG product is off-spec, it must be routed immediately to the glycol slops tank TK-5001.

The sump of the column is at 161°C at a pressure of 253 mbara. The bottom products are pumped by MEG column bottoms pump, P-5002A/B and filtered by MEG column bottom filter FT-5001A/B.



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The net bottom products which are controlled by FIC-50102 that takes the set point from the level of the column bottom, LIC-50101 are sent to MEG recycle column T-5002.

Heat to T-5001 is supplied through a falling film type reboiler, E-5003 by MCS with flow control on the steam inlet, FIC-50103 that takes the set point from TIC-50104.

The required vacuum is supplied by vacuum system, PK-5001 on E-5002. A nitrogen line is provided to compensate leak air in case of excess performance of vacuum system via PV-50103.

MEG recycle column

The bottoms from the MEG column contains 70 wt % MEG. This MEG is recovered in the MEG recycle column. The column has a diameter of 3400 mm and contains two sections of packing, one with Sulzer BX and one with Mellapak 250Y. The total packed height is 5270 mm.

The overhead vapours at 122°C and 57 mbara are condensed in internal condenser E-5005 with cooling water as cooling media.

The liquid overhead products are drawn off and pumped by MEG recycle column overhead pump P-5007A/B. The net MEG product controlled by FIC-50405 that takes the set point from the level of drawn off tray LIC-50403 is sent to dehydrator, T-4004 to allow the removal of any light decomposition products, while the reflux flow is controlled by FIC-50406.

The required vacuum is supplied by vacuum system, PK-5001. A nitrogen line is provided to compensate leak air in case of excess performance of vacuum system via PV-50402.

The column bottoms contains about 0.02 wt % MEG at 168°C and at a pressure of 76 mbara. The bottom products are pumped by MEG recycle column bottom pumps, P-5006A/B. The net bottom products which are controlled by FIC-40502 that takes the set point from the level of the column bottom, LIC-50401 are filtered by MEG recycle column bottoms filter, FT-5002A/B and sent to DEG column.

Heat to T-5002 is supplied through a falling film type reboiler, E-5006 by MCS with flow control on the steam inlet, FIC-50403 that takes the set point from TIC-50402.



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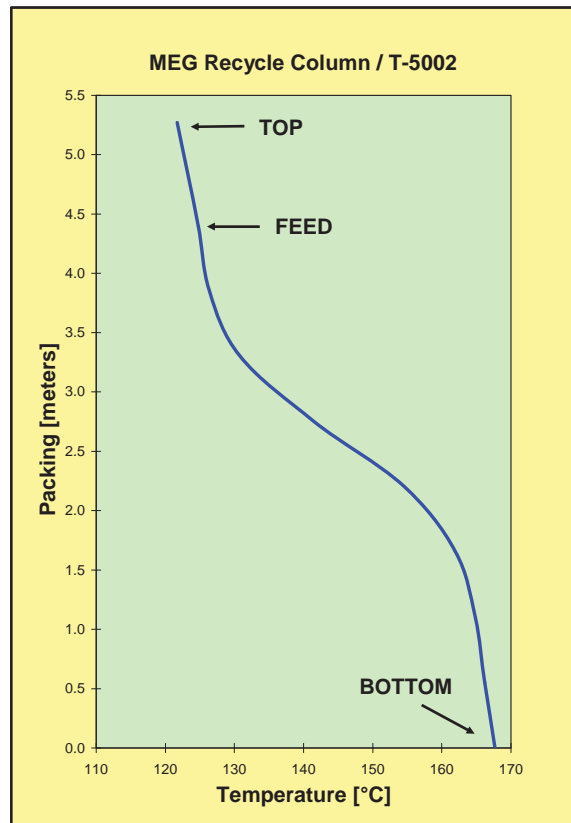


Figure 2.5.1-2: Temperature profile of MEG recycle column T-5002

2.5.2 DEG purification (PFD No. : Unit 500-2/2)

Refer to the process flow diagram for DEG and TEG purification.

The bottoms of the MEG recycle column enter the DEG column T-5003. The column has a diameter of 2800 mm and contains 2 sections of Sulzer BX packing of 3784 mm in total. On top of each BX bed a bed of 420 mm Mellapak 250Y is provided to enhance liquid distribution. The total packed height is thus 4624 mm.

The overhead vapours at 122°C and 10 mbara are condensed in internal condenser E-5007 using cooling water as cooling media. The DEG product is drawn off and pumped by DEG rundown bottoms pump P-5009A/B, cooled to 50°C in E-5010 and sent to a product sales tank outside the battery limit. While the reflux flow is controlled by FIC-50505 and sent to T-5003 after filtering in FT-5006A/B. The flow rate of net DEG product is controlled by the level of draw off tray LIC-50503.

The DEG column bottoms are at 179°C at a pressure of 30 mbara. The bottom products are pumped by DEG column bottom pump, P-5008A/B and



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filtered by DEG column bottoms filter, FT-5005A/B. The net bottom products that are controlled by FIC-50502 are sent to TEG column, T-5004.

The required vacuum is supplied by vacuum system, PK-5001. a nitrogen line is provided to compensate leak air in case of excess performance of vacuum system via PV-50502.

Heat to T-5003 is supplied through the falling film type reboiler E-5008 by MCS with flow control FV-50503.

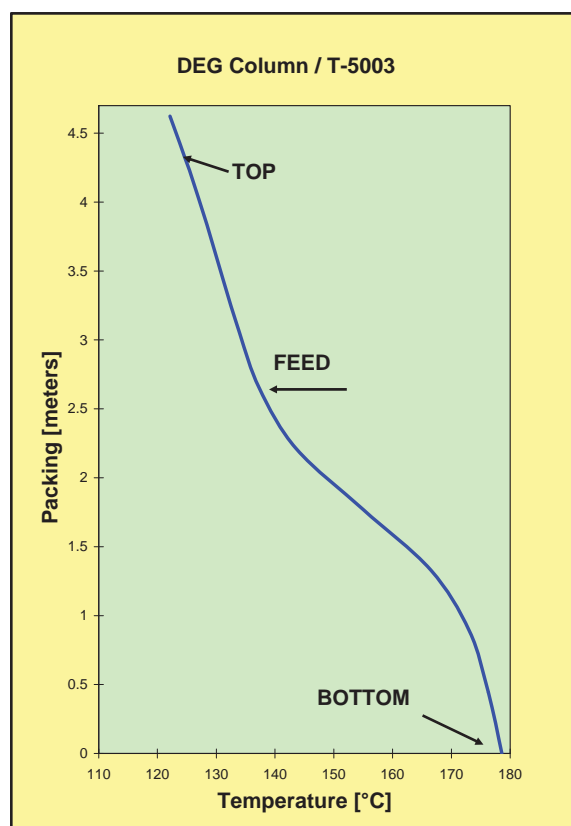


Figure 2.5.2-1: Temperature profile of DEG column T-5003

2.5.3 TEG purification (PFD No. : Unit 500-2/2)

Refer to the process flow diagram for DEG and TEG purification.

The TEG column, T-5004, has a diameter of 650 mm and contains 2 sections of Sulzer type BX packing with a total height of 2408 mm. On top of each BX bed a bed of 420 mm Mellapak 250Y is provided to help liquid distribution. The total packed height is 3248 mm.



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The internal condenser E-5012 using cooling water condenses the overhead vapours at 139°C and 8 mbara. The liquid overhead products are drawn off and pumped by TEG rundown bottoms pump, P-5012A/B and is cooled to 50°C in E-5011 and sent to a product sales tank outside the battery limit While the reflux flow is controlled by FIC-50605.

The column bottoms are at 186°C and 26 mbara and contains TEG plus higher glycols and heavies.

The column bottoms are pumped by TEG column bottoms pump, P-5011A/B and filtered by TEG column bottoms filter FT-5004A/B. The net bottom products which are controlled by FV-50601 are sent to heavy glycols cooler, E-5014 and then to heavy glycols tank, TK-5002.

The required vacuum is supplied by vacuum system, PK-5001 and the column pressure is controlled by PV-50602.

Heat to T-5004 through a falling film type reboiler, E-5009 by MCS with flow control on the steam inlet, FV-50603.

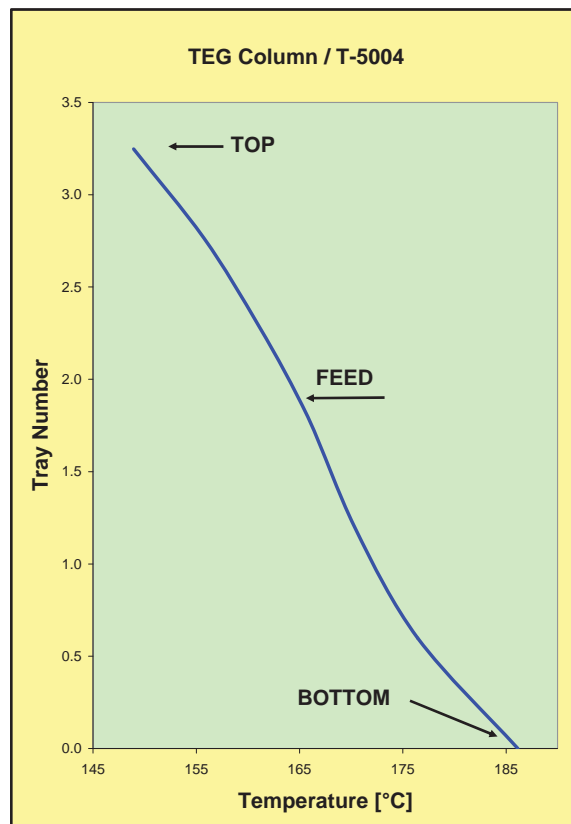


Figure 2.5.3-1: Temperature profile of TEG column T-5004



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2.6 Chilled water circulation system (PFD No. : Unit 600-1/2)

Refer to the process flow diagram for chilled water circulation systems.

Because the temperature of the cooling water system is too high for various applications in the process, a chilled water system is provided, comprising a chilled water tank TK-6050, chilled water circulation pumps P-6050A/B, and package chilling unit PK-6050A~F.

The cooling of the chilled water system is accomplished with an absorption refrigeration system. The driving medium for the absorption refrigeration units is a combination of PS 3K process steam and CS 2K clean steam to cool down the chilled water inventory from 15°C to 10°C.

An automatic pump start system for the circulation pumps P-6050A/B is provided to ensure an uninterrupted chilled water supply.

The chilled water tank will absorb fluctuations in chilled water temperature to prevent the absorption refrigeration unit from falling subject to crystallisation phenomena.

A bypass that is controlled by PDIC-65101 recycles the chilled water to the buffer tank.

2.6.1 Absorption refrigeration unit (PFD No. : Unit 600-1/2)

Six absorption refrigeration units (PK-6050A~F) have been provided to produce chilled water. Absorption refrigeration operates on the principle of evaporating water at low pressure and absorption in lithium bromide solution, which is a strong absorbent for water.

The machine consists of two shells, installed at different elevations. The lower shell is divided into an absorber and an evaporator section, while the upper shell consists of a generator and a condenser section.

The evaporator contains the refrigerant water. The refrigerant temperature is maintained at the desired low level via evaporation. The heat of evaporation is extracted from the water to be chilled and which flows through a coil. Circulating the refrigerant and spraying it on the top of the liquid enhances the evaporation.



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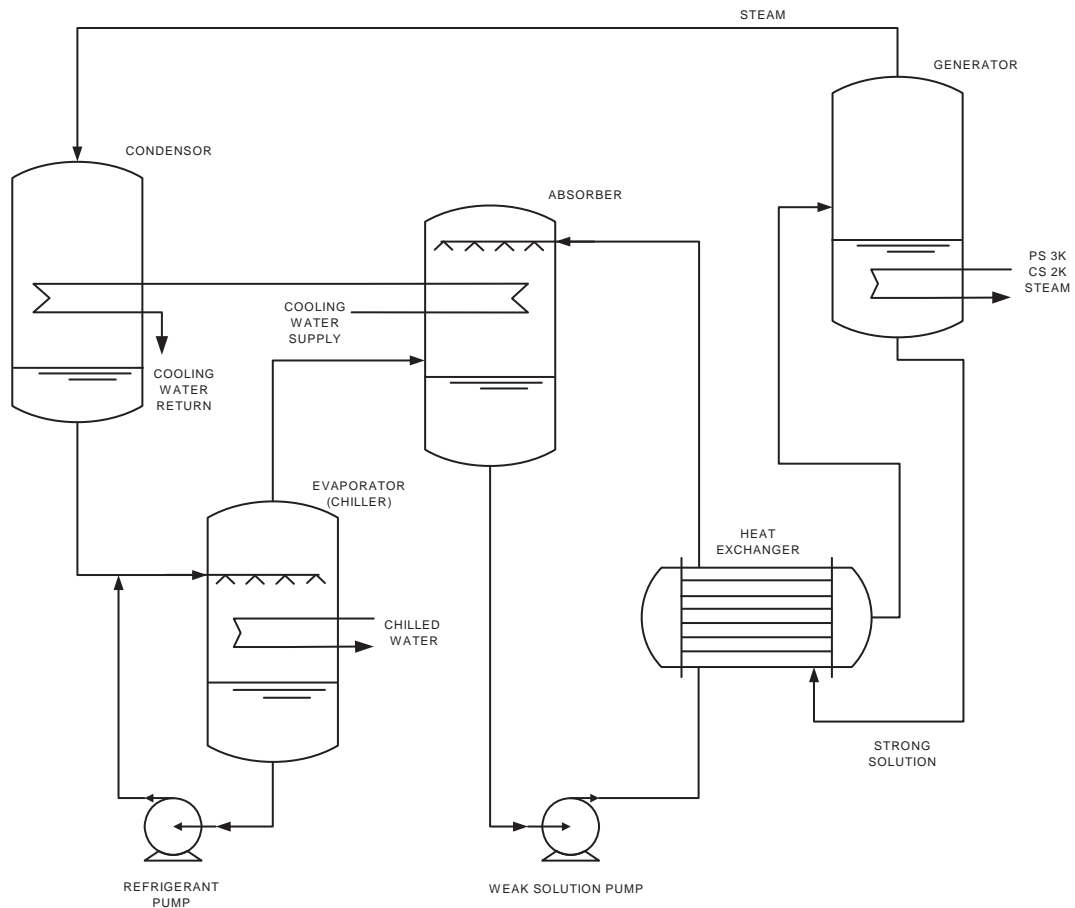


Figure 2.6.1-1: Principle of absorption refrigeration

The low vapour pressure in the evaporator and the pressure gradient to the absorber are created by absorption making use of the affinity of lithium bromide to water. The dilute lithium bromide solution, obtained after water absorption, is pumped via a heat exchanger, in which it takes up heat from the concentrated lithium bromide solution, to the generator. Here the solution is concentrated again by evaporating the water. PS 3K and CS 2K steam supply the heat of evaporation. The water vapour is condensed in the condenser by cooling water. The condensate is recycled to the evaporator. The concentrated lithium bromide solution from the generator is returned to the absorber after being cooled by the weak solution coming from the absorber.

For general information refer to chilling unit vendor documents.



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2.7 Steam and condensate system (PFD No. : Unit 600-2/2)

Refer to the process flow diagram for the steam and condensate balance and boiler feed water, condensate recovery and demineralised water systems and refer to the steam balances for the line up of the steam and condensate systems.

2.7.1 Steam systems (PFD No. : Unit 600-2/2)

Name	Source	Press bara	Temp °C	Condition	Quality
IMPORT HP	IMPORT	39.0	400	Superheated	Utility grade
Reactor steam	HA SOC	22.3	218	Saturated	Clean
Reactor steam	HA EOC	35.9	244	Saturated	Clean
Reactor steam	HS SOC	32.3	238	Saturated	Clean
Reactor steam	HS EOC	54.2	269	Saturated	Clean
CS 20K	HEADER	21.5	216	Saturated	Clean
PS 16K	T-4001	17.1	204.7	Saturated	Process
CS 14K	HEADER	14.9	198	Saturated	Clean
PS 12K	T-4002	12.5	189.8	Saturated	Process
PS 5K	T-4003 / D-4003	5.9	158.4	Saturated	Process
PS 3K	D-4002	3.9	142.8	Saturated	Process
CS 2K	D-4001 / E-5001	2.9	132.9	Saturated	Clean

The plant is designed to use all steam generated within the plant. The process grade steam contains some MEG and traces of aldehydes.

Steam generated in the EO reactors is controlled at the pressure that is required to control the reaction process. Thus this pressure depends on the type and age of the catalyst. The generated steam is let down to the CS 20K header.

HP import steam is desuperheated in EJ-6011 and then supplied to the CS 20K header. Pressure of CS 20K header is controlled by PIC-90103. Temperature is controlled by BFW inlet to desuperheater, TIC-90103. HP import steam can also be supplied directly into the HP steam drum D-1007 to heat up the EO reactors (for start up).

Process grade steam is generated and used in the multiple effect evaporation unit for dewatering of the glycol reactor product. PS 5K process



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grade steam from glycol concentrator-3 T-4003 and the process condensate flash drum D-4003 is used in various reboilers in the plant and steam to package unit vacuum system, PK-5001.

2.7.2 Condensate systems (PFD No. : Unit 600-2/2)

Name	Source	Press bara	Temp °C	Condition
CS 20K	CS 20 K USERS	21.5	216	Clean
PS 16K	E-4005	17.1	204.7	Process
CS 14K	CS 14K USERS	14.9	198	Clean
PS 12K	E-4006	12.5	189.8	Process
PS 5K	PS 5K USERS, D-4003	5.9	158.4	Process
PS 3K	PK-6050, D-4002	3.9	142.8	Process
CS 2K	PK-6050, D-4001	2.9	132.9	Clean
BFW	D-6001		130	Clean
DMW	OBL		27	Clean

In general, hot condensate streams will be flashed where possible to generate low pressure steam. CS 20K condensate, MPC is flashed in the MP steam drum to generate CS 14K steam. CS 14K condensate is flashed in the steam condensate flash drum, D-4001 to generate CS 2K steam. PS 16K and PS 12K condensate are flashed in the process condensate flash drum, D-4003 to generate PS 5K steam. PS 5K condensate is flashed in the water surge drum, D-4002 to generate PS 3K steam.

Clean condensate streams can be re-used in the process. The main consumers of clean condensate in the process are HP steam drum D-1007 and import steam desuperheater EJ-6011. For energy conservation reasons the clean condensate is preheated in boiler feed water preheater E-1004 before it is sent to the HP steam drum.

Before clean condensate is recycled back to the process it is cooled down in deaerator preheater E-6001, filtered in FT-6001A/B and treated in condensate polishing package PK-6001, which includes a trim cooler and polishing bed. The condensate is stored in the demineralised water tank TK-6001 which can also be supplied with import demineralised water. DMW is sent directly from the storage tank TK-6001 to the quench bleed flasher T-2006 as reflux by the demin. water supply pump P-6003A/B. Before sending the demineralised water to the other systems, any dissolved light gases are removed in deaerator D-6001. After leaving the deaerator the clean condensate is conditioned to control pH and an oxygen scavenger is added. Corrosion inhibitor is injected on D-1007 inlet.

PS 3K and PS 5K condensate streams are combined with glycol dehydrator overhead condensate and CO₂ stripper overhead condensate and quench bleed flasher overhead condensate and vacuum condensate from



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vacuum package unit PK-5001 and discharged to waste water after being cooled in waste water trim cooler E-4011.

It is possible to send water from the steam condensate flash drum D-4001 to the water surge drum D-4002. In particular circumstances, such as shutdown of the EO producing part of the process while still producing glycols, this is required. It is to be avoided, however, since it means degradation of the clean condensate.

Process steam generated in the glycol concentrator-1 and 2 gives rise to process condensate, which is collected in process condensate flash drum D-4003. Flashed off process steam is routed to the PS 5K header. The process condensate is used to heat the glycol reactor feed in the glycol reactor preheater-2, E-4002, before being collected in the water surge drum, D-4002, with the process condensate from other reboilers to generate PS 3K steam. These condensate streams contain trace quantities of glycols and aldehydes.

The process condensate from the water surge drum, D-4002, is recycled by process condensate pump, P-4004A/B to provide excess water to the feed of the glycol reactor. As long as there is no negative impact on the MEG quality, the process condensate from D-4002 may also be used as reflux to glycol concentrators-1, 2 and 3.

The condensate from D-4004, the glycol dehydrator reflux drum, is sent to waste water treatment by glycol dehydrator reflux pump, P-4006A/B. However, provided that product quality does not deteriorate, some of the process condensate from D-4004 may be returned to the water surge drum, D-4002, thereby reducing the amount of waste water sent to waste water treatment. In the basic operating case, however, there is no recycle of dehydrator overhead condensate and therefore this special operating case is not reflected in the steam balance.

2.8 Glycol product storage and transfer (PFD No. : Unit 500-2/2)

Each of products MEG, DEG and TEG is sent to its corresponding storage tanks, MEG storage tanks TK-8010A/B, DEG storage tanks TK-8020A/B and TEG storage tanks TK-8030A/B respectively.

The glycol slops tank, TK-5001 is provided to receive off-specification product and/or slops streams from the plant. The slops is recycled by glycol rerun pump P-5001 to glycol dehydrator T-4004 and glycol bleed flasher T-4005 via glycol return filter FT-5009 and it is controlled by FIC-50801.

In the transfer lines to the main storage tanks accurate coriolis flowmeters



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FIQ-50251, FIQ-50551 and FIQ-50651 are installed which can measure the production transfer flow from the plant.

2.9 BFW treatment system (PFD No. : Unit 600-2/2)

BFW to steam generation system, D-1006, D-1007 is produced at Deaerator, D-6001.

Recycle condensate from condensate flash drum, D-4001 and chilling unit are treated to remove dissolved oxygen in deaerator, D-6001 by using stripping steam CS 2K and to remove ionic impurities in condensate polishing package PK-6001.

Clean steam condensate from condensate flash drum D-4001 via P-4003A/B on level control of the drum, LIC-40801 is mixed with condensate from clean condensate suction vessel, D-6051 via P-6051A/B on level control of the mentioned drum, LIC-65202. Total condensates are cooled in deaerator preheater, E-6001 and returned clean condensate cooler, E-6002 to 50°C and sent to condensate polishing package PK-6001.

Demineralized water of high purity outlet from PK-6001 is stored in demineralized water tank T-6001. Make up for compensating the vented steam and condensate to waste water is added on the tank level, LIC-60201 to demineralized water tank TK-6001. TK-6001 has a diameter of 12000mm and height of 11100mm and is made of carbon steel coated with epoxy.

Demineralized water is preheated to 98°C and sent to deaerator D-6001 via demineralized water supply pump P-6003A/B. Flow is controlled on deaerator level, LIC-60101. Dissolved oxygen is boiled out from the water at elevated temperature by injection of clean steam (LLS). Chemicals from chemical system, PK-6011 are added to BFW outlet line and DMW feed line for upgrading the water quality. Boiler chemicals are for PH control, reduction of corrosion and removal of dissolved oxygen.

Deaerator pressure shall be maintained at the set pressure to keep the BFW quality. Pressure of deaerator is controlled by PIC-60101 by adjusting the inlet steam by FIC-70105 and PV-70101 on steam flow outlet.

The high purity boiler feed water (HBFW) is pumped by boiler feed water pump P-6001A/B to the boiler feed water preheater mainly in EO reaction unit. HBFW can also be supplied to HP steam desuperheater EJ-6011. P-6001A/B are equipped with an auto start system to supply the water without interruption.



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2.10 Vacuum system (PFD No. : Unit 500-2/2)

The vacuum system PK-5001 consists of a number of ejectors using low pressure steam (LPS), surface condensers which condense ejector outlets, a condensate collection vessel and condensate pumps.

The flow of LPS can be monitored with FI-50701. Before the steam is sent to the ejectors it passes a strainer and a condensate K.O. pot. The condensate passes a steam trap and is collected in condensate vessel D-5003. Collected condensate is sent to waste water collection basin Z-7070 via E-4011 by vacuum condensate pump, P-5015A/B on LIC-50701.

For more details refer to vacuum package vendor drawings and documents.

2.11 MEG, DEG and TEG main storage and handling (PFD No. : Unit 800-1/1)

Two storage tanks, TK-8010A/B are provided and located in main centralized storage area to receive MEG product from MEG polishing vessel, D-5001A/B. MEG product is sent to marine loading arms (in jetty) by MEG product transfer pump, P-8010A/B.

Each MEG storage tank is designed for 20 days production and product is stored under N₂ blanketing. The minimum flows of transfer pumps are recycled to tanks via mixing eductors Z-8005A/B in order to mix the product before sending to loading arm. An on-off valve XV-80101 is provided on minimum flow line of transfer pumps. It automatically opens when the flow rate delivered by the pumps fall below the threshold value for pump stability. Each storage tank is equipped with breather valve and venting manhole for tank protection.

Two storage tanks TK-8020A/B are provided and located in main centralized storage area to receive DEG product from DEG column T-5003. DEG product is sent to marine loading arms (in jetty) by P-8020A/B.

Each DEG storage tank is designed for 20 days production and product is stored under N₂ blanketing. The minimum flow of transfer pumps are recycled to tanks via mixing eductors Z-8006A/B in order to mix the product before sending to loading arm. An on-off valve XV-80201 is provided on minimum flow line of transfer pumps. It automatically opens when the flow rate delivered by the pumps fall below the threshold value for pump stability. Each storage tank is equipped with breather valve and venting manhole for tank protection. Steam coils are provided for each tank.

Two storage tanks, TK-8030A/B are provided and located in main centralized storage area to receive TEG product from TEG column T-5004. TEG product is sent to drum filling unit PK-8010 by P-8030A/B. At any time one of the pumps is stand by and the other one in service. Each TEG storage tank has



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been designed for 20 days production and the product is stored under N₂ blanketing. Steam coils are provided for each tank.

In order to mix the product before sending to drum filling unit, the minimum flow lines of transfer pumps are recycled to tanks via mixing eductors. Each storage tank is equipped with breather valve and venting manhole for tank protection.

2.12 TEG Drum Filling Unit (PFD No. : Unit 800-1/1)

TEG is fed by TEG product transfer pumps P-8030A/B to the 200 liters drum filling unit.

The filling unit has been designed for 6 days/week, one shift/day working period and is constituted by an automatic train that will fill, seal and mark 200 liters drum. For details and working principles refer to manufacturer documents.

A fan, F-8001 with a capacity 1000Nm³/hr has been provided to take the vapor gases from drum filling package and discharge them to atmosphere.

2.13 EG Drain system (PFD No. : Unit 700-1/1)

EG process drain from the equipment with high EG concentration is collected through a drip funnel and conveyed to the underground drain drum for EG/water D-7060 through an unburied complete closed system. Collected drain is pumped by drain pump for EG/water P-7060 and sent to glycol hydrator T-4004 via hot slope header and glycol slop tanks TK-5001 in case of high concentration EG and otherwise sent to waste water collection, Z-7070.

Drain drum for EG/water D-7060 is designed to collect two times of T-4004 drain volume. Before draining in maintenance and operational drainage, all equipment should be pressurized and the liquid level should be reduced to low liquid level.

2.14 EO Drain System (PFD No. : Unit 700-1/1)

EO process drain is directly collected and conveyed to the underground drain drum for EO/water D-7065 through an unburied complete closed system.

Drain drum for EO/water is designed to collect EO drains in the maintenance. Before draining in maintenance and operational drainage, all EO in the equipment in EO service which is in operation should be converted to EG and sent to EG slops tank. The EO drain drum may be in operation if no