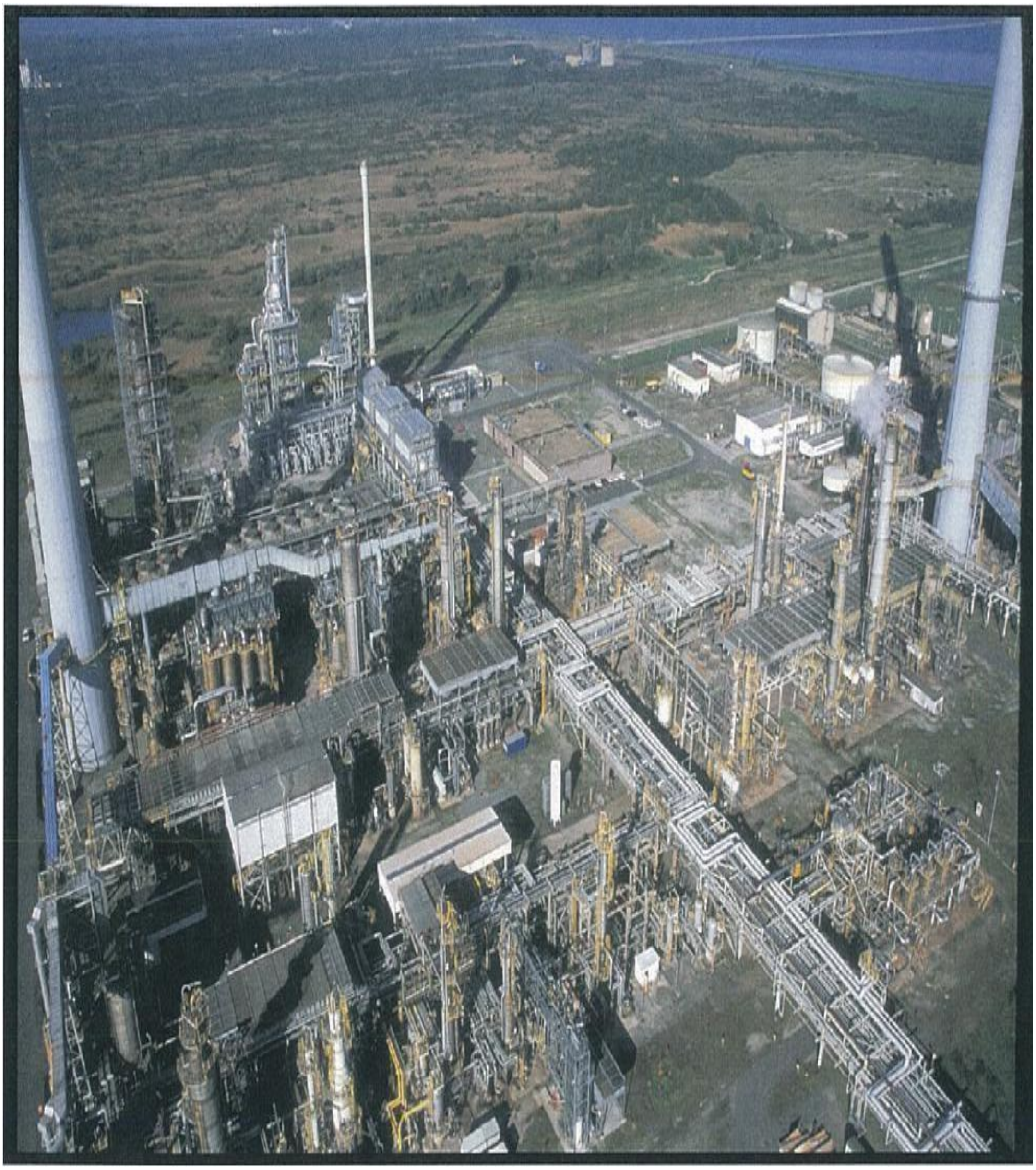
Sale of Refinery Assets



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EXECUTIVE SUMMARY

- Site is currently operated by Hestya Energy operating the storage facilities and terminal with the refinery shutdown.
- The refinery was originally built by Mobil in the period between 1972 and 1975. Mobil operated the refinery for approximately 10 years, making several investments to expand the refinery and improve performance during that period. In 1985 Mobil shutdown and mothballed the refinery, through to 1990 when Mobil sold the refinery to Beta Raffinerie. The facility was restarted in 1991. In 1997 the international trading company, Louis Dreyfus, purchased the refinery. Over the next 8 years the crude processing capacity was expanded and other downstream units were upgraded. In two cases, units were converted to alternative services; the former gas oil dearomatiser unit was converted to a light naphtha isomerisation unit and the distillate dewaxer to a distillate desulphurisation unit. In addition a major investment was made in installing a new high capability vacuum distillation unit. In 2006 the refinery was sold to Conoco

  Phillips Corporation. Conoco Phillips shutdown the refinery in September 2009 and after a fire on the crude atmospheric distillation plant in 2010 during a plant start up the refinery was never restarted. In late 2011 the refinery was sold to Hestya Energy who continue to operate the storage and terminal facilities and are looking at options an what to do with the refinery units.
- The refinery crude oil throughput capacity is 275,000 barrels/day.
- The refinery is a large capacity hydroskimming refinery producing a range of petroleum products including;
- Propane/butane
- petrochemicals
- Gasoline
- Kerosene
- Low Sulphur Diesel
- Sulphur
- Heating oils
- Fuel gas used within the refinery
- VG0
- Heavy fuel oil

2009 Products	
	Thousend Barrels
Propan	417
Butane	1111
Normal Pentane	245
ISO Pentane	66
Naphtha	3371
Gasolin	
Inland Eurograde	7497

Inland Super Plus	54
US Grade 87 Oktane	560
US Grade 93 Octane	2347
Reformate	1581
Kerosine	249
Destillates	
Inland Diesel	5253
Export Diesel	7442
Inland Heating Oil	2054
Export Diesel (50 ppm)	2764
VGO	5007
Fuel Oil	4220
Total	**44238**

- The refinery has 10 main production units; ○ Atmospheric Distillation ○ Vacuum distillation ○ Gasoil Hydrodesulphurisation ○ Naphtha pretreater

○ Catalytic semi regenerative reformer ○ Isomerisation ○ Distillate desulphuriser ○ Kerosene treatment plant

○ Gas plant

○ Claus Sulphur recovery plants

	Demonstrated sustained capacity KBD	Unit No.	Original build date	Last revamp	Design/licence/ catalyst
Crude distillation	275	1100	1974	2005	Mobil/Koch
Vacuum distillation	110	1200	2004	2005	Fluor/Uhde Edeleanu
Naphtha pretreater	88	1400	1974	2005	Mobil/Haldor Topsoe
Reformer	45	1500	1974	2009	Mobil/Criterion
Isomerisation	12	1600	1974	2000	Mobil/UOP
Gas oil Hydrodesulphurisation	78	1900	1974	2009	Mobil/Shell
Distillate desulphuriser	15	2000	1980	2009	Mobil/Albemarle
Gas oil dryer	23	1900	2002		
Kerosene treatment	15	2200	2009		
Gas plant	41	1700	1974	1997	
Sulphur recovery units	2 x 50Mt/day	2300	1974		Axens
Sour water unit	73	2400	1974		

- The refinery has a large tank farm with 63 tanks and a total storage capacity of 8.2 million barrels. This comprises tanks to store crude oil, intermediate product, and finished product and LPG spheres. The refinery also has extensive road, sea and rail terminal facilities.
- The site owns and operates its own marine facilities with deep water port capabilities. It consists of an inner jetty (offshore) connected to the shore by an access and pipe bridge and an off shore jetty connected to the refinery by sub-sea pipelines. Each of the jetties has 2 berths. The outer jetty can receive very large crude carrier of up to 250,000MT deadweight.
- In total the refinery with tank farm and terminal operations and day staff employed 437 people. In operation the refinery employed 146 staff in operations; Most were shift operations personnel organized on a rotating 5 shift cycle system, as detailed below;

○ 23 day personnel ranging from supervision and management to day based routine checking.

○ 112 shift personnel on 5 shifts. Each shift has 2 supervisor positions, 3 panel positions and the rest are outside operators. ○ 14 Appendices in training.

- The refinery is a simple hydroskimming plant which can only refine about 60% of the crude to higher value lighter products, with the rest as heavier fuel oils. With the excess refining capacity in Europe and falling margins it was more challenging to operate the refinery at adequate margins. In addition the refinery was limited to types of crude that could be processed through the plant the plant was installed with a corrosion chemical addition system to allow the processing of higher acidity crudes, but was still limited. An upgrade program was developed by Conoco Phillips to upgrade the refinery from its current hydroskimming configuration to a full conversion refinery capable of processing an entirely sour crude slate. The proposed upgrade project comprised of a high conversion VGO hydrocracker, a delayed cooker, a cooker distillate hydro treater, hydrogen plants, sulphur recovery units and other supporting units. The total cost of this upgrade was **$3b.**

Completion of the upgrade program would allow the refinery to operate on a significantly lower cost Urals crude oil slate versus the historic sweet, light crude slate.

- Operating costs (full refinery including tank farm and terminal operations) ○ Estimated 2011 full year operating cost is $125m. In operating years that exclude major turnarounds, salaries and other personnel costs represent 28% of the total operating costs. Routine maintenance materials and contract costs represent about 27% of the total operating costs. Electrical power (all imported) accounts for a further 18 to 20%. Catalyst costs are typically about 12 % and other costs account for about 6%.

○ Turnaround costs for the current refinery configuration were estimated at $38m in 2014.

- In recent years the Wilhelmshaven refinery processed a sweet light crude oil slate predominately from the North Sea, with lower volumes of North and West African crude oil.
- The refinery's middle distillate hydro processing units have the capability to desulphurise all gas oil product to the European diesel specification or less than 10 ppm sulphur. The VGO produced was exported and sold as a feedstock to other refineries with FCC or hydrocracker based facilities.
- The refinery remains operable and could be re-commissioned without significant cost or delay.
- There are a number of technology licenses to operate the plants. It is likely that only the ISOM plant and UOP will require additional payment.
- Site operates under a Seveso permit (HSE)

- The refinery plant ceased operations in September 2009. A fire on the crude atmospheric distillation plant occurred on restarting the plant in 2010. The cause of the fire was incorrectly installed insulation and corrosion on pipe work. The plant was repaired and many other pipelines checked and repaired where required. Since then the majority of the plant and equipment has been decommissioned. Selected equipment items have been removed to dry storage such as larger motors and the reformer compressor

- The plant was generally in good condition. Detailed records are available on site of maintenance history and routine pressure vessel inspections. The equipment dates from 1972 through to 2004, although considerable sums of money have been spent on various units to upgrade and maintain them.

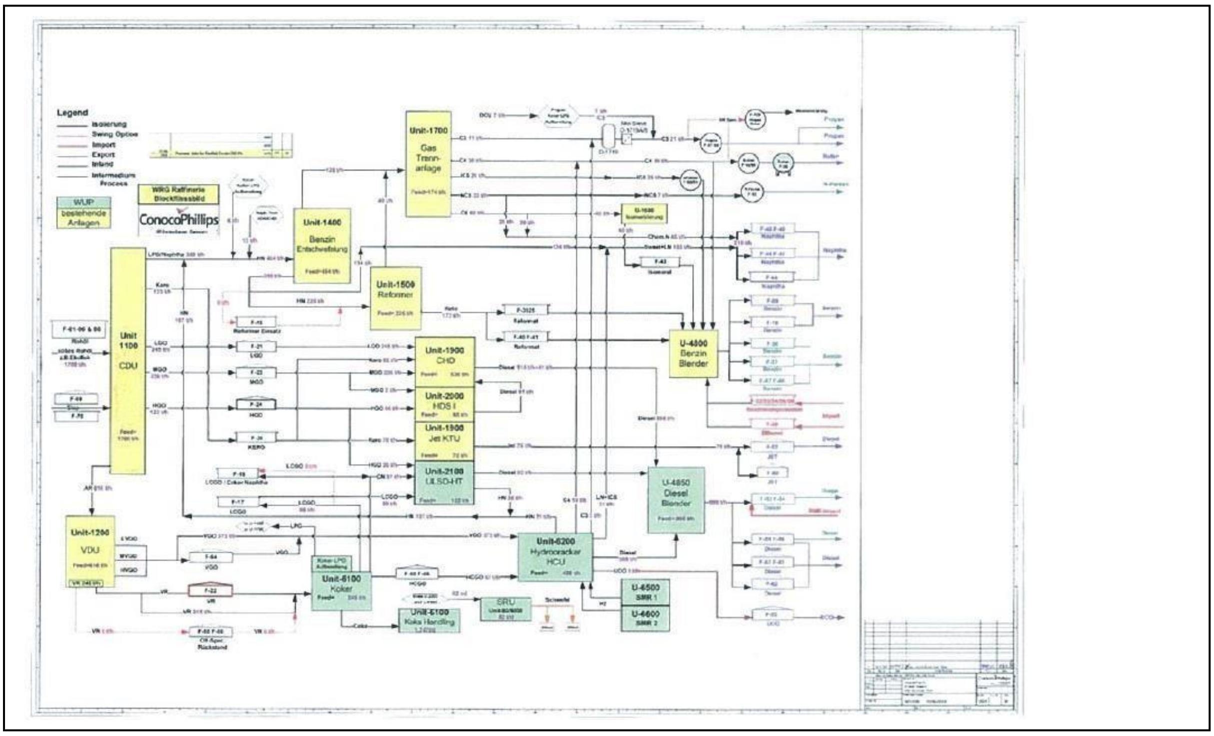
- Documentation is available for equipment, maintenance and operations. Scheduled key plant inspections are carried out approximately every 5 years.

- There is minimal asbestos present although records were not seen. The majority of asbestos was removed from the plant over the years.

- Utility plants consists of;
- large steam boilers with associated ancillary plant
- Cooling water systems
- Instrument air plant
- Electrical distribution (all electricity is imported)

- The refinery plant is fully automated with a modern process control system. This consists of Honeywell DCS including PHD, RMPCT (with Oracle) for process automation and plant information.
- Honeywell DOC3000 DCS documentation tool

Wilhelmshaven refinery process flow diagram (green units not present) Key



Site Information (Lactation map)

The entire site is 513 acres with the refinery process units occupying a small part of this. The refinery was original built in 1972 through to 1975. The plant was shut down in September 2009. The site is made up of 4 main areas;

Refinery process plants

Tank farm
Terminal operations
Offices

There are 63 large storage tanks surrounding the refinery process plants.
Hestya Energy continue to operate the site as a storage facility and terminal operations by road, sea and rail.

Site Plan

Environmental Considerations

The plant appears to have been fully decommissioned and cleaned of chemicals.
Services and utilities are still connected and connections to the tank farms still in place.
There is minimal asbestos, although no asbestos reports were seen.
There are 3 chimney stacks on site, plus the flares. Analysis equipment is fitted to all stacks to monitor SO2 and NOX emissions.

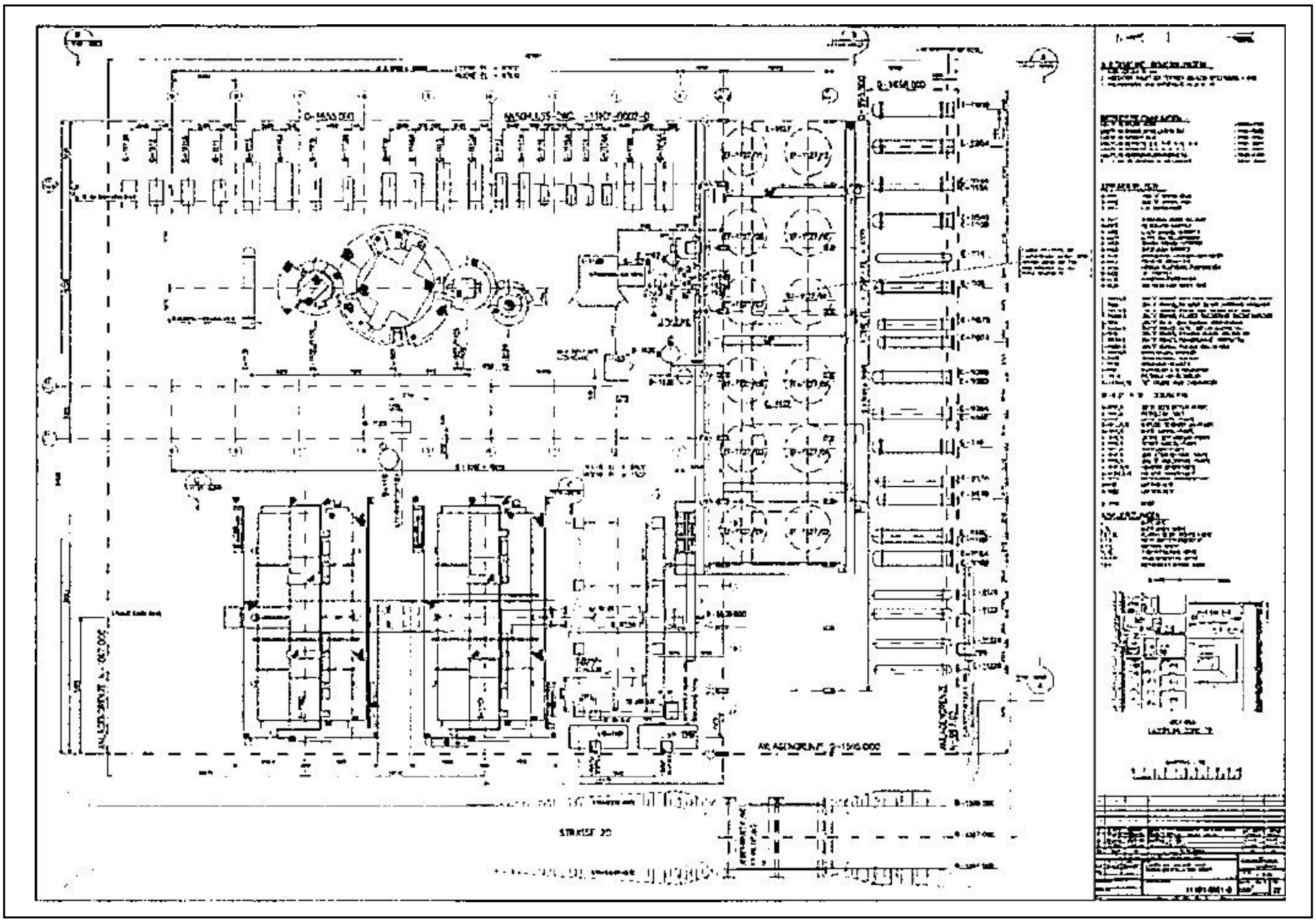
Safety and Environmental performance

All requirements of the environmental and safety authorities are at least fulfilled.

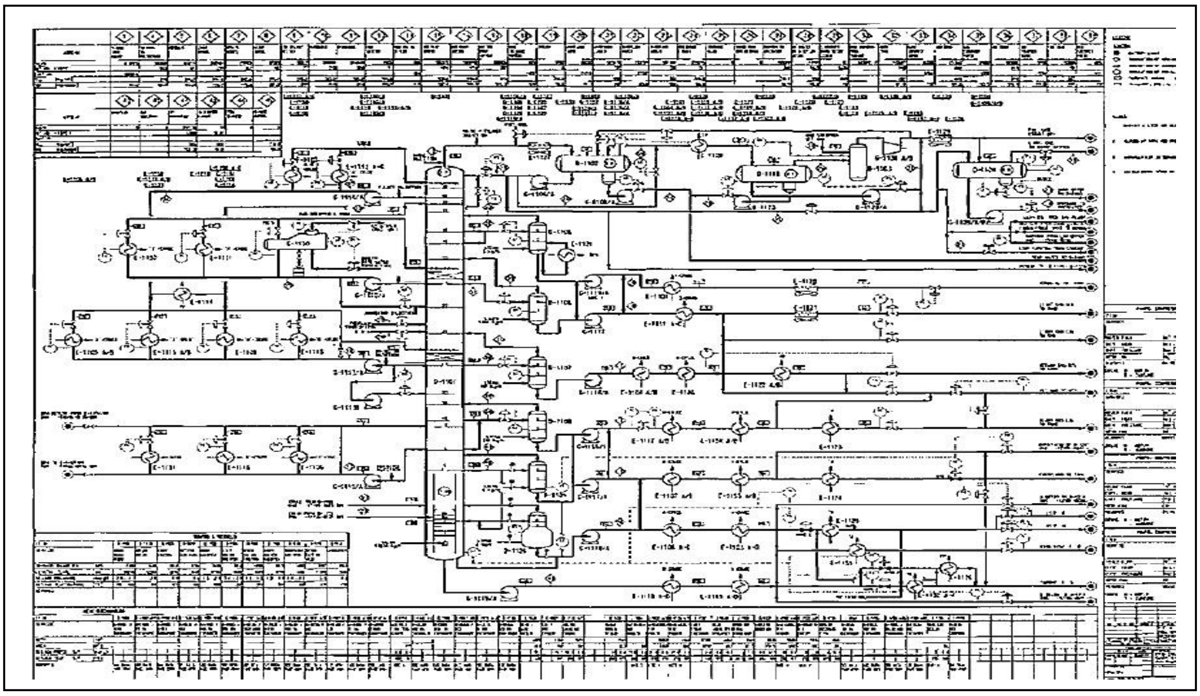
(The graphics are unclear because the original is over 3 square meters in size)

Process Descriptions

ATMOSPHERIC CRUDE DISTILLATION UNIT 1100



Unit 1100 Atmospheric Crude distillation footprint



Crude Distillation Unit 1100 PFD

The atmospheric crude distillation plant unit 1100 has a capacity of 275,000bbliday and is a Mobil design. It is capable of handling two trains of crude input and has some capability to process acidic crude oils by the use of corrosion additives and the addition of corrosion monitoring equipment. The refinery was limiting crude charge acidity to a maximum of 0.7 total acid number, although the additive provider claims that's its additive package should provide protection up to 2.0TAN. The crude unit has two separate parallel preheat trains, which are used to run different crude oils at the same time. The crude composition could be changed on a regular basis, typically every 24 to 36 hours. There are densitometers on both trains, which allow the ADMC control to adjust for the changing crude density. Once the crude

Two desalters fort the 2crude trains Preheat heat exchanger trains and fin fan coolers oil enters each train it passes through a series of heat exchangers that cross pump-a-rounds and draw streams from the top of the atmospheric tower. After desalting in D1112 & D1113, the hot crude enters the pre-flash drums D1110 for the A stream and D1111 for the B stream) where some of the light material is flashed and the vapor enters around Tray 22 of the atmospheric column D1101. The de-salters are 3.6m in diameter and 18.2m and 24.3m long with a design pressure of 27.7barg. The crude is then passed through a series of additional heat exchangers, which are crossed with the bottom pump-arounds and side draws of the atmospheric tower and the vacuum residue draw from the bottom of the vacuum tower.

The hot crude oil then enters the furnaces where additional heat is supplied to improve the separation of the raw crude. Within the furnaces, the exit temperature is controlled by adding either fuel oil or fuel gas, depending on various environmental and process constraints. Each furnace consists of two cells with two passes flowing through each. A crude heater has a duty of 58 Mal and B heater 78 MW. The pass flows are balanced to equal each other at all times. This balancing helps to reduce the possibility of hot spots occurring within each pass. The air addition to the furnaces is also controlled relative to achieving the corresponding environmental constraints such as NO_X, CO and SO_2.

Two pre-flash Towers

Furnaces for the two crude feed steams

From the furnace, the hot crude oil at about 370C enters the atmospheric tower above tray 4 near the bottom of the tower separately from each train. The tower operates at around 1 bar above atmospheric pressure. It is 7.5 m in diameter and 54.2m tall and has a design operating temperature of 345c to 400c. It has a total of 43 plates and two distributor sections. When the hot oil enters the column the lighter materials flash and begin traveling up the column. At the same time cooler condensed liquid products travel down the column contacting the vapor and cooling it.

Atmospheric crude distillation column

The lighter materials proceed up the column and the heavier materials tend to flow towards the bottom of the column. To help improve the separation of the components reflux, or cooled products, is introduced at different points down the column Pump-around side draws remove material from different stages within the column. The material then flows through various heat exchangers that transfer the heat from the side draw material to the colder crude oil flowing through one of the pre-heat trains. Overhead vapors are injected with amine and cooled through the crude tower fin fan condensers E1127 (which have a total surface area of and sent to the reflux accumulator drum D1102. Liquid is pumped back to the column and the vapors are further cooled and sent to drums D1118 and D1104. Gasoline product is pumped to unit 1400 for further processing and the non-condensed gases are sent to flare or other plants.

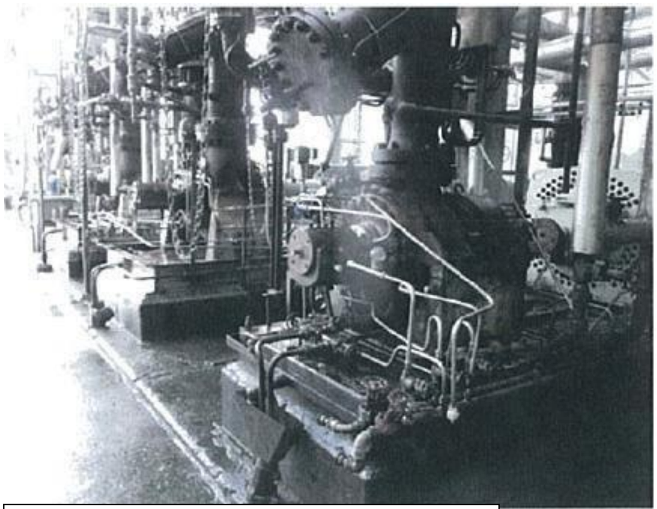
OVF from plate 6 is sent to the over flash stripper D1109 which has 4 plates and fed with 1300 Kg/hr steam. The vapours are sent back to the column and the liquid pumped through a series of pre heat exchangers to OVF oil blending storage.

Product from plate 4 is fed to the A stripper D1126 which has 6 plates and then pumped to storage for fuel oil blending. There is a HGO bottom pump around off plate 9 returned at plate 12. In addition HGO is fed to the HGO stripper D1108 which has 4 plates and is fed with 1330 Kg/hr steam. HGO off the bottom is pumped through a number of heat exchangers and sent to storage.

MGO off plates 21/22 are partially pumped around and the rest sent to the MGO stripper D110 with MGO product pumped to storage. LOGO off plates 29/30 are sent to the LGO stripper D1106 with LGO pumped to storage. Finally Kerosene off plate 35 is fed to the Kerosene stripper D1105 and kerosene product pumped to storage.

The heavy material in the bottom of the atmospheric column is then pumped to the vacuum column.

Side strippers



Pump alley

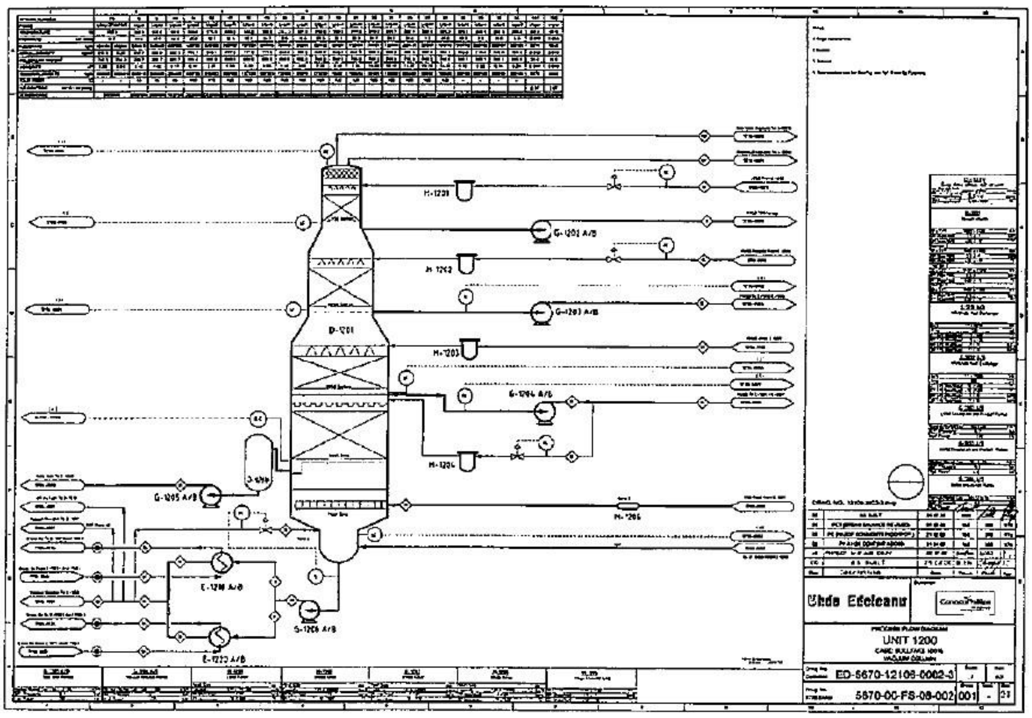
Vacuum Distillation plant

The Jeep-cut Vacuum Distillation Unit (VDU) was installed in 2004 to recover valuable distillates from atmospheric residue which is received from the crude distillation unit (CDU).

The VDU is designed for a feed rate of 575 std. m3/hr (110 Kbbld) and can be operated in the range of 50 to 110% of design feed rate. The design feed rate is based on 1500 std. m^3/hr Gullfaks_30 crude oil feed to the upstream crude distillation unit (CDU). It was designed by Fluor Daniels and Uhde Edeleanu.

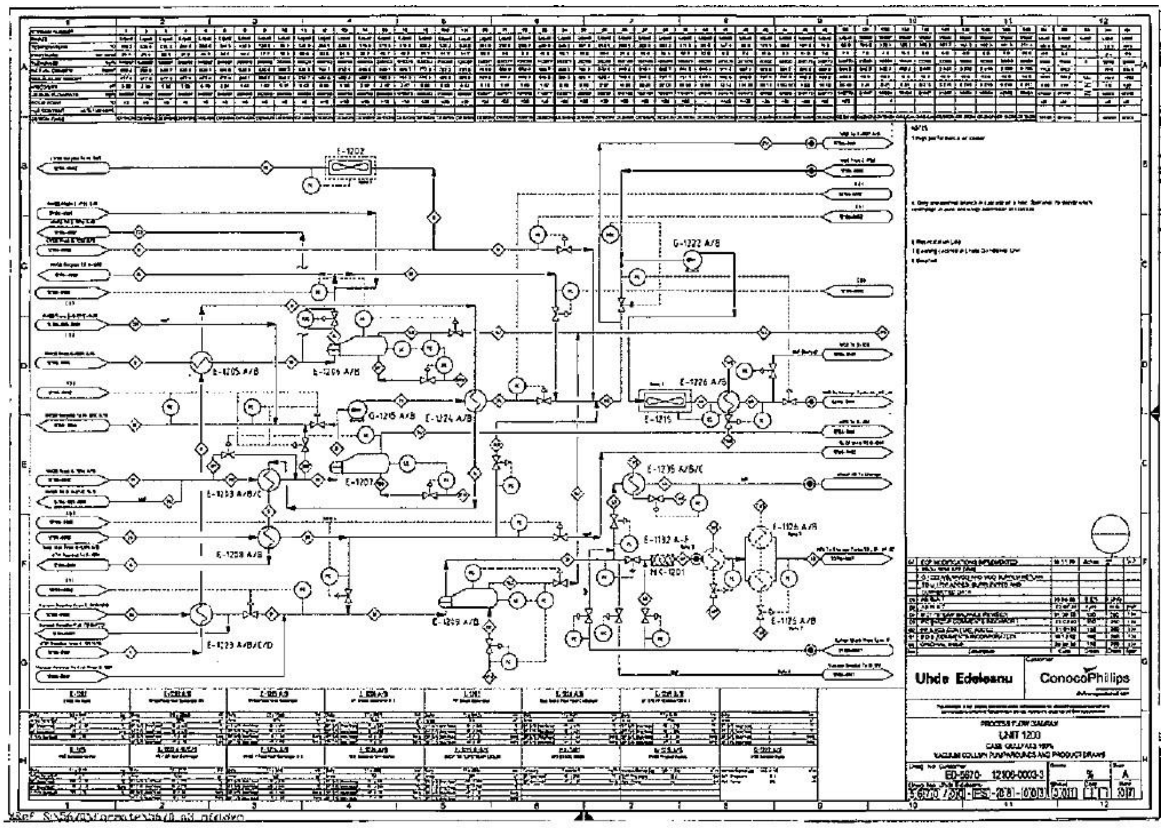
The design VDU feedstock is constituted by a combined A-bottoms, B-bottoms and over flash fraction of the CDU with an initial TBP (True Boiling Point) cut point of 390 °C.

The VDU is designed that the yield of the combined vacuum gas oil distillates (LVG) + MVGO + HVGO), VGO, reaches at least 65 % by volume based on VDU feed from Gullfaks_30 crude oil. This yield corresponds to a TBP cut point of the VDU feed of 550 °C.



Vacuum column PFD

The VDU feed received from the CDU is fed to the VDU feed drum D1211 and then preheated in a heat exchanger train (13 off shell and tube heat exchangers), in which the feed is heated against VDU product and recycle streams to about 190C, and through the vacuum furnace B1201 to 425C before the hot feed enters the vacuum distillation



VDU heat exchange PFD

CDU feed drum/furnace PFD

The VDU feed received from the CDU is fed to the VDU feed drum D1211 and then preheated in a heat exchanger train (13 off shell and tube heat exchangers), in which the feed is heated against VDU product and recycle streams to about 190C, and through the vacuum furnace B1201 to 425C before the hot feed enters the vacuum distillation column D 1201.

The furnace consists of a single cell with a design duty of 69.5 MW . The vacuum column has four main sections;

LVGO section which is 4.8m dia X 6.4m with a demister, distributor and packed section.

MVGO section which is 7.4m dia X 4.9m with a distributor and packed section

HVGO section which is 9.4m dia X 22m with a packed section, distributor, collection tray and a wash zone packed section Flash zone of 3.9m dia x 7.7m.

Temperature range across the column is 68C to 380C.

The feed from the furnace is fed to the flash section. The vacuum column separates the feed into light vacuum gas oil (LVGO), medium vacuum gas oil (MVGO), heavy vacuum gas oil (HVGO), slop wax (SW) and vacuum residue (VR). The purpose of drawing more than one vacuum gas oil distillate is to achieve a high degree of heat recovery and efficiency. Slop wax is pumped from below the wash section through heat exchanger E1208, E1209 to raise LP steam and further cooled to 85 C in E1235A/B/C to storage.

The vacuum residue bottoms are pumped through a series of heat exchangers and mixed with the slop wax from the VDO column and sent to storage. The HVGO cut taken drum the bottom of the HVGO packed section is partly recycled and the rest pumped through heat recovery shell and tube heat exchangers, the E1207 MP steam generator to storage. The MVGO is cut from the bottom of the MVGO packed section and pumped through a series of heat exchangers and the LP steam generator E1206 to storage. The LVGO cut is pumped to storage, with some cooled and recycled to the top of the column and the overhead vapours are drawn off by the 4 stage vacuum ejectors $J1201/^{2}/_{3}/4$ with their associated condensers. The mixed gases/liquid are fed to the ejector condenser effluent separator drum D1202 and the resultant waste gas removed.

Many of the shell and tube heat exchangers are manufactured from stainless steel as well as much of the pipework and vessels.

The distillates LVGO, MVGO and HVGO are combined to saleable vacuum gas oil (VGO). Slop Wax and vacuum residue are combined and blended with cutter stock to a saleable heavy fuel oil (HFO).

In order to meet the tight specification of the VGO, the vacuum column is equipped with a single wash bed which enhances the separation between HVGO and Slop Wax/VTB, i.e. minimizing contaminants in HVGO and finally in the VGO product. This wash bed is installed above the flash zone and consists of a high performance structured packing providing more than three theoretical separation stages. Another important design feature of the vacuum column is the gravity distributor above the wash bed which ensures adequate wetting of the wash bed packing over a range of 65 to 110% of design hot reflux rate.

A Brief summary of test results is shown in the following table. The achieved figures as shown in the table prove that the VDU clearly passes the requirements of the Functional Specification.

Parameter	Functional Specification	Achieved
Feed rate std. m^3/hr	575	587.8
VGO product rate std. ms/hr	375	389
VGO yield Vol.°/0	min. 65	70
TBP cut point °C	550	563

VGO quality	meet specified quality parameters	all quality Parameters met
Vacuum furnace thermal efficiency %	min. 90 (1)	91.3
Heat Input into crude MW	19.83	23.36
Electrical power consumption MW	max. 3.7	2.75
Low pressure steam export t/hr	max. 40	38
Regulatory and environmental standards NO_x emission mg/Nm^3 CO emission mg/Nms Particulate mg/Nm^3 SO_2 emission mg/Nm^3	(1) 450 175 50 1,700	(2) 391 56.1 48.9 858
VGO export loading rate std. m^3/hr	3,200	3,247
Fuel oil export loading rate std. m^3/hr	3,200	3,236
LCO import capacity std. ms/hr	800	984
Gasoline export loading rate std. m^3/hr	2,200	2,294 (3)

Catalytic Reformer Unit 1500

D1502 2.9m dia x 7.6m TT

D1503 2.9m dia x 8.9m TT

D1504 3.2m dia x 9.8m TT

The naphtha from unit 1400 is fed to the catalytic reformer for the processing of low octane naphtha from the crude unit, chemically reforming it into high octane gasoline blend stocks. The plant was commissioned in 1974 and based on a Mobil design. It has a capacity of 45,000 bbl/day of feed. It is a semi-regenerative unit using 4 off down flow fixed bed reactors each with about 25Te of catalyst. The reformer produces sufficient hydrogen used for the 1400 and 1600 units. It is a semi-regenerative reforming unit and all of the catalyst is regenerated in place every 2-3 years

The feed of 174Te/he is mixed with the recycle 112 from the recycle gas compressor G1502 and is preheated to 440C in E1501 and fed to the reformer charge furnace B1501 to 505C and then to the first reformer reactor D1502. The pressure inlet to the 1st reactor is 21.3bar and 17.2 bar exiting the last reactor. The vaporized reactants enter the first reactor. As the vaporized reactants flow through the fixed bed of catalyst in the reactor, the major reaction is the dehydrogenation of naphthenes to aromatics which is highly endothermic and results in a large temperature decrease between the inlet and outlet of the reactor. Depending on the feedstocks the decrease in temperature can vary from 50 to 90°c. To maintain the required reaction temperature and the rate of reaction, the vaporized stream is reheated in the second fired heater B1502 before it flows through the second reactor D1503. This is repeated in the 3rd and 4threactors (D1504 and D1505). As the vaporized stream proceeds through the three reactors, the reaction rates decrease and the reactors therefore become larger. At the same time, the amount of reheat required between the reactors becomes smaller reactor sizes.

The hot reaction products from the 4th reactor are cooled through the E1501 heat exchangers and E1502 fin fan cooler and pass to the reformer flash drum D1506. The reformate from the bottom of the flash drum is pumped to the depentansier. The gases off the top of the flash drum which are rich in H2 and Cl, C2, C3 pass to the off gas booster compressor and the recycle gas compressor. The off gas compressor G1503 which is 3100HP raises the pressure to 42 bar and feed H2 rich gas to other units. The remaining H2 rich gas off the flash drum are fed to the recycle gas compressor G1502 which is a 11,500 HP and raises the pressure back to 22.3 bar.
The compressor G1503 was being removed to storage for preservation. Both G1502 and G1503 are steam driven compressors. G1503 has 2 stages.


GI 502 recycle gas compressor   GI 503 off gas booster

The reformate from the flash drum is preheated to 170C in E1505 and E1503 heat exchangers and fed to plate 17 of the depentaniser D1507. This is 3.1m diameter and 30.2 m tall and operates at 15 bar. The column has 34 plates. A bottom pump around is reheated in the furnace B1505 to 250C.Overhead vapours and cooled and partially recycled with the rest fed to the gas plant. The heavy reformate from the bottom of the depentansier is pumped to storage after cooling.

Cycle	2	SOR	02.09.2007	Charge today (m^3/h)	165
04.02.2009	14:00			Cum. charge (m^3)	2265478

Liquid

Days on Stream:	522	Flow	Temperature					Corrected	Flow	
Liquid Stream	Indicator	Reading	Indicator	°C	Density	correction		m^3/h	kg/hr	LV %
Naphtha	FC14015.pv	182,7	TI14121.pv	66,3	754,2	1,0139		185,3	139728	-
DeC5 O/H Liquid	FC15024.pv	18,7	TI15061.pv	47,6	569,9	1,0306		19,3	10978	10.47
Reformate	FI15021.pv	131,5	TI15187.pv	20,4	819,7	1,0195		134,0	109863	72.34
Reformate CTS		134,8		15,0		1,0000		134,8	110496	72.76

Gas Streams

		Flow		Temperature		Pressure		Corrected	Flow	
Gas Stream	Indicator	Reading	Indicator	°C	Indicator	Reading	Dens.	Nm^3/h	kg/h	Nm^3/m^3
DeC5 O/H Gas	FI15025.pv	1341	TI15061.pv	47,6	PC15014.pv	12,6	1,391	1233	171€	6,81
Sep Gas to Fuel	FI15017.pv	2090	TI15053.pv	41,9	PC15009.pv	17,5	0,416	2086	868	
Booster Off Gas	FC15015.pv	40040	TI15053.pv	41,9	PC15009.pv	17,5	0,416	39493	1642€	
Anti Surge Gas	FI15016.pv	115	TI15054.pv	127,1	PI15021.pv	40,0	0,416	109	45	
Net Booster Gas								39384	16381	
Recycle Gas	FI15020.pv	306461	TI15053.pv	41,9	PC15009.pv	17,5	0,416	305940	12725(	

Reactor Temp Input

	Inlet temperatures		Outlet temperatures						
Reactor	Indicator	°C	Indicator	°C	Temp. Diff.	Av. Bed T	%DT		
D 1502	TC15011.pv	493,4	TI15008.pv	430,2	63,2	461,8	73,67		
D 1503	TC15012.pv	493,4	TI15016.pv	472,0	21,4	482,7	24,95		
D 1504	TC15013.pv	493,4	TI15024.pv	490,2	3,3	491,8	3,81	RONC (lab)	101,3
D 1505	TC15014.pv	490,5	TI15032.pv	492,6	-2,1	491,6	-2,46	RONC (C5+)	100,1
	WAIT	492,5		Total Delta T	85,8			R Gas H2	78,4 %
	dWAIT	-5,9			WABT	484,5		R Gas H2O	15,3 vol ppm

Calculation Results and Lab Results

Catalyst life =	13,9972	m^3/kg		LHSV =	0,950	/h				
	39,9242	BPP			Injection rates:	Water	Dh =	50	Wt. ppm =	4,00
H2/HC-ratio =	8,07	mol/mol		Cl Conc.:	10,4	Chloride	Dh =	50	Wt. ppm =	1,00

Wt. recov. =	100,06%	Wt recovery using tank-level:	100,51%	
C5+ =	76,71	LV %	p inlet last reactor =	18,46 bar g
Net H2 Yield =	181	Nm^3/m^3	Molwgt Feed =	105 kg/kmol

PONA

PONA (vol%)	P	O	N	A
Reformer Feed	45,44		40,29	14,27
Reformate	25,40	0,26	0,96	73,36

Feed S =	0,13	
N + 2A =	68,8	
RVP =	18,5	kPa

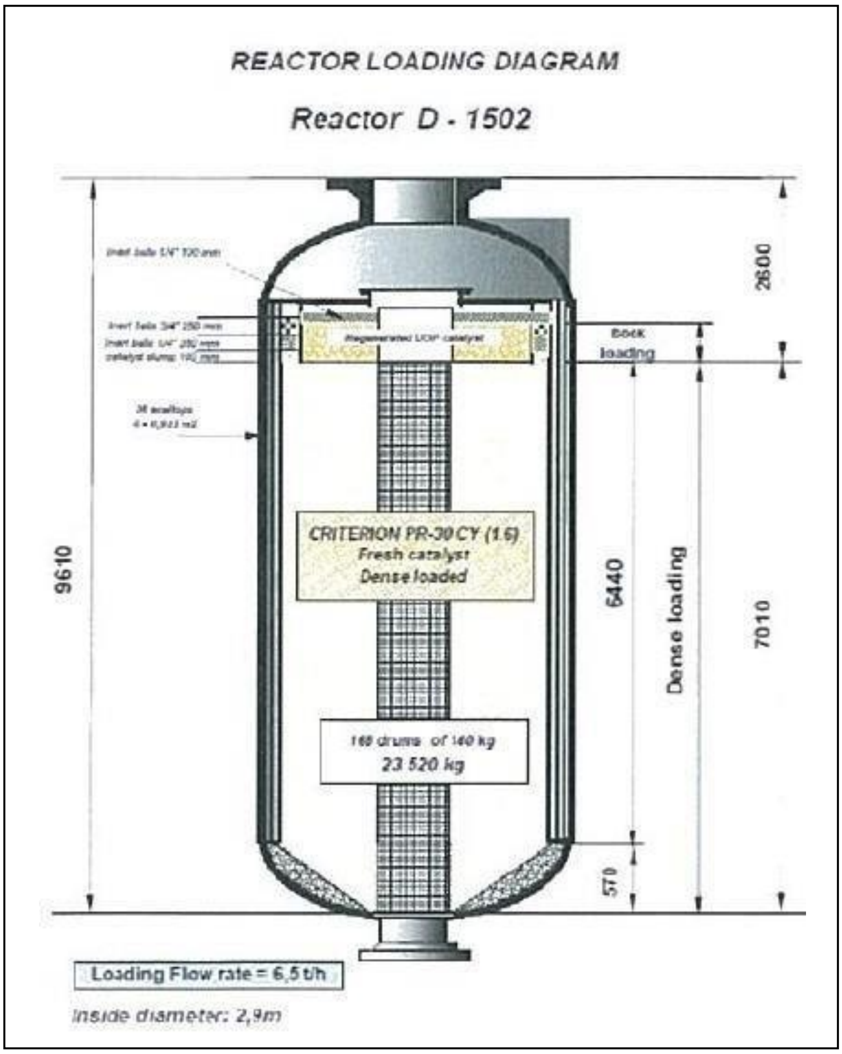
Benzene in reformate = 2,91 Vol %

Yields:	wt %	lv%
H2	2,16	
C1	1,64	4,13
C2	2,94	5,93
C3	4,80	7,12
iC4	2,65	3,54
nC4	3,53	4,55
iC5	3,74	4,50
nC5	2,50	2,99
C6 +	75,04	69,30
Total	100,00	101,99

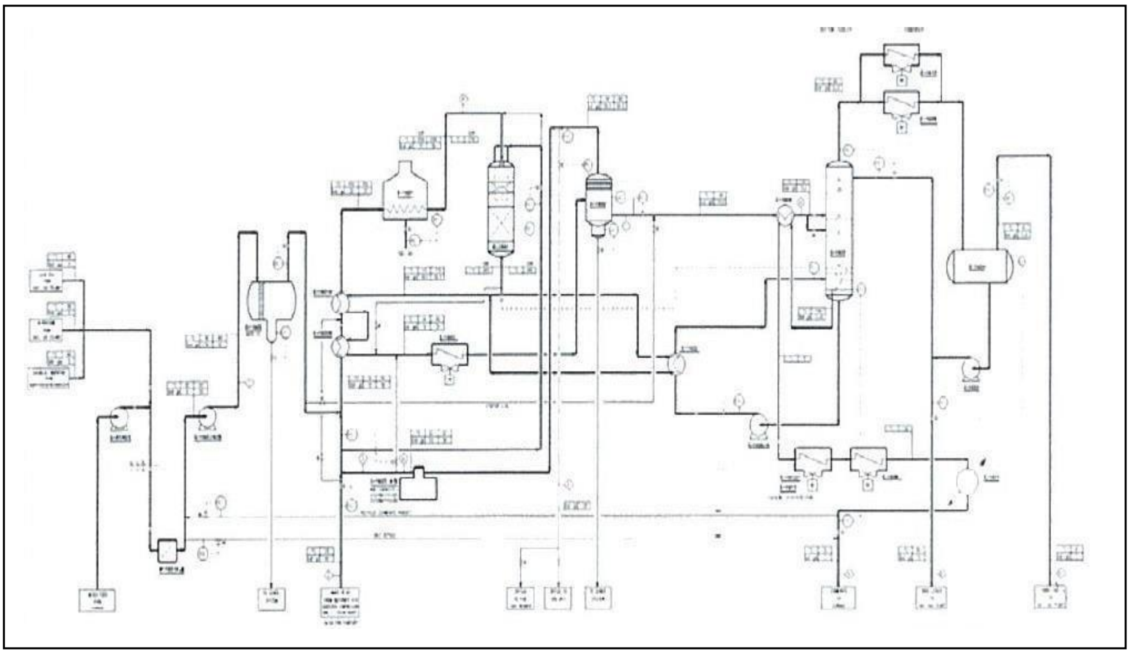
C5+ wt% 82,29

50% Feed = 116,9°C

Mass Balance for Unit 1500



The Isomerisation Plant Unit 1600 Plant



The isomerisation plant was installed in 1974. It has a capacity of 12,000bbl/day. The plant design is Mobil/UOP. There are 3 main sections to the plant;

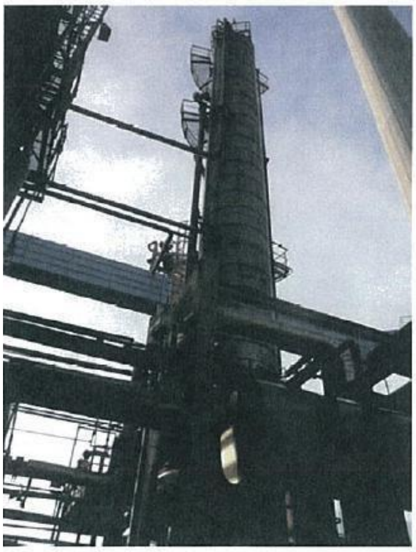
112 recycle feed and charge feed Reaction Stabilisation

The Unit 1600 consists of 1 heater, 1 reactor, a HD separator, a stabilizer column and a recycle gas compressor together with several pumps and heat exchangers. The main purpose of the unit is to increase the octane number of the lean oil stream (C5/C6). This is achieved with a zeolitic catalyst to process a reduction of the remaining aromatics in the stream. The vapor pressure of the isomerate product is adjusted within the stabiliser.

The amount and composition of the U-1600 lean oil stream depend highly an the feed streams from other process units. The process is closely controlled to manage the benzene & C7+ concentrations as the benzene and C7+ react exothermically.

Performance of Isomerisation U-1600

Several parameters influence the increase of the octane number: Reactor temperature, feed rate, composition of the lean oil and activity of the catalyst. High reactor temperatures have a lower result in octane but the conversion is nearer to the equilibrium. The higher the feed rate the shorter is the retention time and therefore the conversion is decreased. High amounts of benzene & C7+ influence the equilibrium in a negative way especially for the C5 conversion. The data below shows the overall mass balance.



Stabiliser colum D 1603

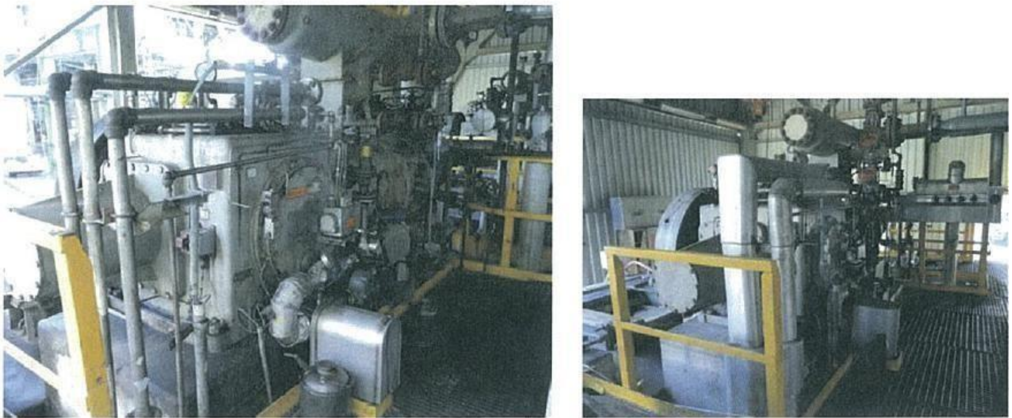
Reactor D1601 & furnace B 1601

Mass balance of U-1600

Component	Unit	Reactor					Stabilizer				
	Yield	M-1002	M-1507	M-1610	M-1005	Yield	M-1610	M-1609	M-1606	M-1604	Yield
		Isom liq. Feed	D-1506 Rec.	Stab. Feed	Offgas OHD		Stab. Feed	Stab. Prod	D1604 liq. OV	D1604 gas OVHD	
Corrections		1,000	1,000	0,950	1,000		0,950	1,000	1,000	1,000	
kg/h	0,98	23158	2190	24430	365	1,00	24430	22601	10	1376	0,98
C5+	0,96	23094	206	23323	97	1,01	23323	22368	4	1	0,96
H2	0,67	0	391	0	247	0,63	0	0	0	15	
C1	1,08	0	298	0	229	0,77	0	0	0	63	
C2	1,07	0	438	56	143	0,46	56	0	0	325	5,79
C3	1,33	0	512	327	93	0,82	327	0	2	566	1,80
C4= 1		0	0	0	0		0	0	0	0	
iC4	2,24	42	171	520	44	2,06	520	129	3	302	0,83
C4= 2		0	0	0	0		0	0	0	0	
nC4	0,79	32	180	200	11	1,01	200	104	1	62	0,78
C4= 3		0	0	0	0		0	0	0	0	
22DMC3		0	0	0	0		0	0	0	0	
C4= 4		0	0	0	0		0	0	0	0	
C5= 1		0	0	0	0		0	0	0	0	
2MC4	2,20	422	63	1097	19	2,21	1097	1086	3	3	0,99
C5= 2		0	0	0	0		0	0	0	0	
nC5	0,80	1140	46	901	11	0,77	901	938	1	0	1,04
C5= 3		0	0	0	0		0	0	0	0	
C5= 4		0	0	0	0		0	0	0	0	
C5= 5		0	0	0	0		0	0	0	0	
C5= 6		0	0	0	0		0	0	0	0	
22DMC4	2,47	843	0	2104	0	2,50	2104	2064	0	1	0,97
C5= 7		0	0	0	0		0	0	0	0	
C5= 8		0	0	0	0		0	0	0	0	
C5= 9		0	0	0	0		0	0	0	0	
C5= 10		0	0	0	0		0	0	0	0	
CyC5	0,90	707	0	630	0	0,89	630	633	0	0	1,00
23DMC4	1,44	999	0	1509	0	1,51	1509	1440	0	1	0,95
2MC5	0,98	5088	0	5224	0	1,03	5224	4967	0	1	0,95
3MC5	1,14	2857	0	3428	3	1,20	3429	3272	0	1	0,95
nC6	0,66	5419	78	3749	67	0,69	3749	3559	0	1	0,95
22DMC5	0,70	49	0	37	0	0,75	37	34	0	1	0,93
MeCyC5	1,06	2570	0	2911	3	1,13	2911	2771	0	1	0,95
24DMC5	0,13	88	0	17	0	0,19	17	16	0	1	0,93
223TMC4	0,73	9	0	7	0	0,79	7	7	0	1	0,93
Benzol	0,00	[illegible]	0	0	0	0,00	0	0	0	1	
33DMC5	0,98	14	0	15	0	1,08	15	14	0	1	0,93
CyC6	0,57	1372	0	818	0	0,59	818	775	0	1	0,95
2MC6	0,38	130	0	54	0	0,41	54	50	0	1	0,93
23DMC5	0,48	44	0	22	0	0,50	22	33	0	1	0,93
11DMCyC5	0,70	42	0	32	0	0,76	32	29	0	1	0,93
3MC6	0,59	104	0	66	0	0,63	66	61	0	1	0,93
c13DMCyC5	0,93	49	0	49	0	1,00	49	45	0	1	0,93
tr13DMCyC5	1,09	39	0	46	0	1,16	46	43	0	1	0,93
3EtC5+tr12D…	0,88	23	0	22	0	0,95	22	20	0	1	0,92
tr12DMCyC5+	0,88	48	0	44	0	0,95	44	41	0	1	0,93
224TMC5		0	0	0	0		0	0	0	1	
nC7	0,56	65	0	39	0	0,60	39	36	0	1	0,93
MeCyC6+c13…	2,11	89	0	190	0	2,22	190	181	0	1	0,95
113TriMeCyC…	2,93	5	0	15	0	3,16	15	14	0	1	0,93
22DMC6		0	0	0	0		0	0	0	1	
ECyC5	2,68	9	0	24	1	2,84	24	26	0	1	1,00
25DMC6+223…		0	0	0	0		0	0	0	1	
24DMC6		0	0	0	0		0	0	0	1	
1tr2c4TMCyC…	3,41	5	0	17	0	3,69	17	16	0	1	0,93
33DMC6		0	0	0	0		0	0	0	1	
1tr2c3TMCyC…	4,88	3	0	12	0	5,27	12	11	0	1	0,93
234TMC5		0	0	0	0		0	0	0	1	
Toluol+233TM	0,00	19	0	0	0	0,00	0	0	0	1	
others		39	0	195	0		195	183	0	1	
		23158	2190	24430	365	1,209	24430	22601	10	1376	1,115

	Tag	Flow [m3/h]	Tag	Temp [°C]	Tag	Press [barg]	Density [kg/m3]	Corr.	corr. Flow [m3/h]	Massflow [kg/h]
Feed from D1005	FC16002.pv	35,0	TI16001.pv	28,0			690,6	0,9630	33,7	23272
Recirculation	FC16007.pv	10,0	TI16034.pv	24,6			679,9	0,9758	9,8	6636
Fresh Feed + Recirc.							690,6		23,9	23160
Feeds from U1700	FC17201.pv	0,1	FC17200.pv	25,9			690,6	0,9710	0,1	38
Make-up gas	FC16013.pv	5006	TI15054.pv	113,3	PI15021.pv	39,0	0,406	1,0648	5430	2196
Recycle gas	FI16508.pv	23790	TI16030.pv	37,4	PI15021.pv	38,3	0,209	1,1287	26805	7217

The various charge feeds are mixed with the 112 rich recycle gas from the units recycle gas compressor and make up from the reformer booster compressor. The Thomassen recycle compressors (A and B) were built in 1974. This stream is preheated to 152C in the E1601A/B heat exchangers and further heated to 250C in the B1601 furnace at a pressure of about 33 bar. This is then fed to the top of the reactor D1601. The reactor is charged with a zeolite catalyst in the lower section and is 2.26m in diameter and 5.6m TT. Some gases off the top of the reactor are recycled to the inlet feed and the reacted product from the bottom of the reactor is cooled in a series of heat exchangers and fin fan coolers and fed to the separator drum D1602 at about 45C. The 112 rich gas is vented back to the recycle gas compressor G1603A/B at about 36 bar. Product from the separator is sent to the stabilizer column D1603 at 13 bar and 107C. The stabilizer column has 32 plates and is 2.2 m in diameter and 23.5m tall. Temperature across the column is 148C to 88C. The overheads are cooled and partly recycled back to the column with the rest sent to the gas plant. The uncondensed gases are also sent to the gas plant There is a reheated pump around an the bottom of the column and the isomerate product from the bottom of the column is cooled and sent to storage.



G 1603 Thomassen recycle compressor

Misc Pumps

NAPTHA HYDROTREATER UNIT 1400

The Unit 1400 consists of 3 heaters, 1 reactor, an HP and an LP separator, a depentaniser and a dehexanizer column together with several pumps and heat exchangers. The main purpose is the removal of catalyst poisons from the naphtha stream for the reformer unit.
The reformer feed requires a feed sulfur content of less than 0.3 ppm (w0/0).
Other constrains exist for the depentaniser overhead with a restriction an the C7+ concentration and a low concentration of benzene and its precursors in the reformer feed stream.
The naphtha treater can treat naphtha with high sulphur content (700ppm) and get the reformer feed sulphur levels down to less than 0.2 ppm. The reactor temperature can be increased to deal with higher sulphur levels. Typically the reactor bed temperature is about 295C.
The naphtha hydro treater was commissioned in 1974 and build to treat the naphtha fraction from the atmospheric distillation unit for the removal of sulphur and nitrogen. It has capacity of 88,000bbl/day and is a Mobil/ Haldor Topsoe design. Naphtha from the atmospheric distillation plant is fed through the naphtha coaleseer and E1401 and E1407 heat exchangers to the unit furnace B1401. H_2 rich gas from the reformer is mixed with the feed after the coalescer at about 41 bar pressure. The furnace has a design duty of 25.8MW pressure of 44.1 bar and temperature of 400C. After the furnace the feed is approx. 370C and 36 bar pressure. The feed passes through the down flow fixed bed reactor D1401 and after cooling passes to the high pressure flash drum D1402 at 50C and 31 bar pressure. The reactor is 3.94m in diameter and 6.2m TT.
Gases vent off the top through a demister and the treated naphtha pass to the low pressure flash vessel. Sour water off the bottom of the flash drums is sent to the sour water stripper and the product stream is fed the depentaniser. The catalyst is recharged about every 5 years.

Furnace B 1401 and reactor D1401 E1401 heat exchangers

G 1401 feed pump

The treated naphtha from the LP flash drum is preheated to 150C and fed to the middle of the depentaniser column D1404. This column is 4.8m in diameter and 30.3m tall with 35 plates. The column operates at about 14 bar and a temperature range across the column of 218C to 129C.

Furnaces B1402 and B1506

There is a bottoms pump around reheated in furnace B1402 to 235C.

Overheads are cooled in the E1405 fin fan coolers and fed to the drum D1405. Gases off the top of the drum are fed to the gas plant. The liquid stream is partly recycled to the column and the rest sent to the gas plant. The bottoms are fed through a series of heat exchangers to the dehexaniser column D1509 at 142C. Sweet naphtha from storage is also mixed and fed to the column. The dehexaniser column is 3.7m in diameter and 29.7m tall with 40 plates. The column operates at about 1.5 bar pressure and a temperature across the column of 158C to 106C. There is a bottoms pump around reheated to 158C through the furnace B1506. The overheads are cooled and partially recycled to the column. Gases off the separation drum D1510 are sent to flare and the chemical naphtha sent to storage. The C7+ off the bottom of the column is cooled and sent to the reformer or intermediate storage

Depentaniser column D1404 Dehexaniser D150

HYDRODESULPHURISATION PLANT UNIT 1900

The hydrodesulphurisation plant was built in 1974, but was revamped in 2009. It is based on Shell technology and can process 78,000bb1iday of feed. It produces ultra-low sulphur diesel (<10ppm).

The reaction feed section can take a number of different feed types, including kerosene, Light Gas Oil, Medium Gas Oil and Heavy Gas oil. These various types of feed can be routed directly to the deaerator, or part of the kerosene and light gas oil can bypass the feed and reactor section directly to the stripper.

The mix of various feeds in the feed header is pumped through the E1901 heat exchanger to the charge deaerator D1901. This is 2.3 m diameter at the top and 3.7 m diameter at the bottom and 15.5 m high, with 16 plates. The feed temperature into the

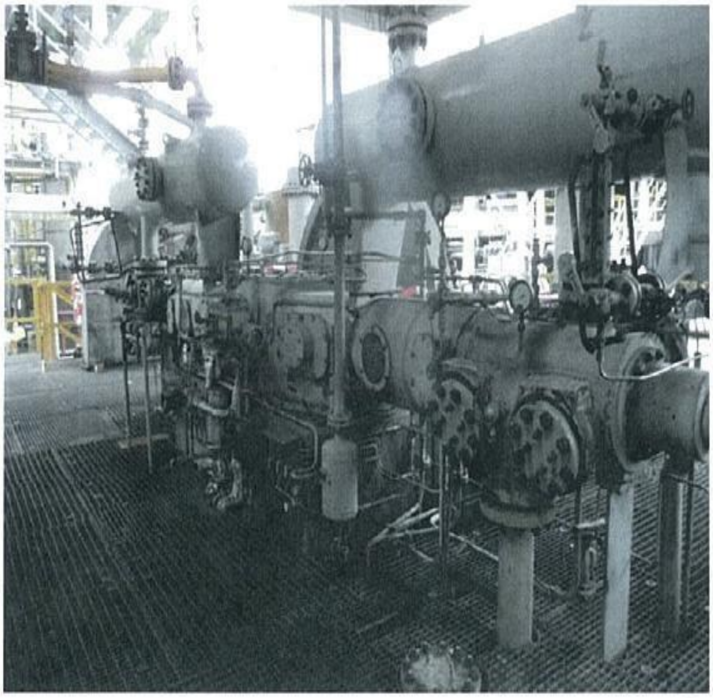
Left:
Deaerator D 1901 & LP amine absorber

Right:
HP amine absorber & HP seperator

deaerator is 71C and 5.5 bar.

Pressure is controlled on the deaerator by controlling the off gas pressure. The off gas is sent to the fuel gas system. From the deaerator the feed is preheated with the reactor effluent and pumped to the reactor furnace B1901. The furnace has a design duty of 25.3MW.The feed to the reactor is 352C and 28 bar pressure. There are 3 pumps available, 2 are generally in operation and one is in standby. Hydrogen is introduced with the feed in the reactor. The H_2 rich gas is fed from units 1400, 1600 and 200 as well as recycle gas

G1901 recycle gas compressor



Reactors D1901 & D1911	B1901 furnace

From the furnace the gasoil is sent through 2 desulpurisation reactors D1902 and D1911 and separated in a hot temperature separator D1903. The reactors are downflow fixed bad reactors that operate at about 345C and 28 to 24 bar. Reactor D1902 is 4.85m in diameter and 10.2m tall and reactor D1911 is 3.88m in diameter and 8.9m tall. The reacted gas oil liquid is introduced to the lower section of the stripper fractionator D1906 at about 260C., and the gasses are sent to the cold separator D1904. The remaining liquid is sent under level control to the top section of the stripper. The off gas from the cold separator is sent to the HP amine absorber D1905 where H2S is removed by washing with lean MEA solution. The off gases from the HP absorber are sent to the amine knock out pot. Some OF the gas is fed to the gas plant with the rest being recycled back to the reactors through the electrically driven G1901 gas recycle compressor, which increases the Pressure to 31 bars. The compressor was built **in 1979.** Remaining liquid from the amine knock out pot is sent to the amine plant 2200.

Stripper column D 1906 B1902 furnace

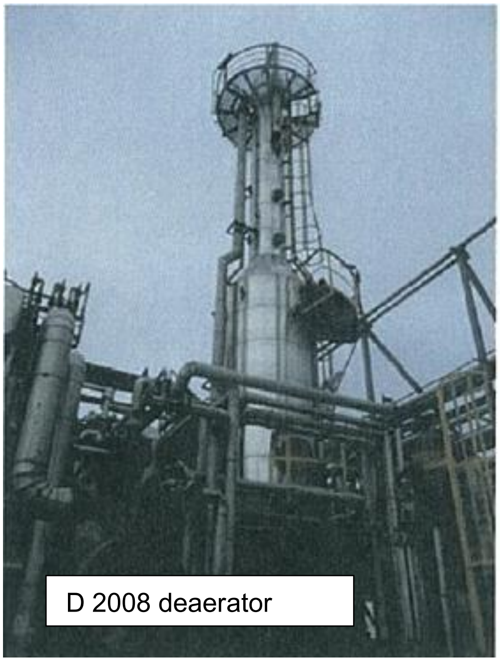
The stripper D1906 consists of 24 trays with a furnace B1902 providing heat for vaporization on the bottoms pump around. The furnace is a fuel gas fired furnace.

The overhead vapors are condensed with an air condenser and collected in the reflux drum D1907. Part of the liquid is returned under flow control as reflux to the column, the rest is sent back to the crude unit under reflux drum level control. The off gas from the stripper is processed in the LP amine absorber D1909 for removal of the H2S. The pressure in the stripper is set by the pressure controller in the amine absorber. Desulphurised product from the bottom of the stripper column is pumped through a series of heat exchangers for cooling and sent to storage.

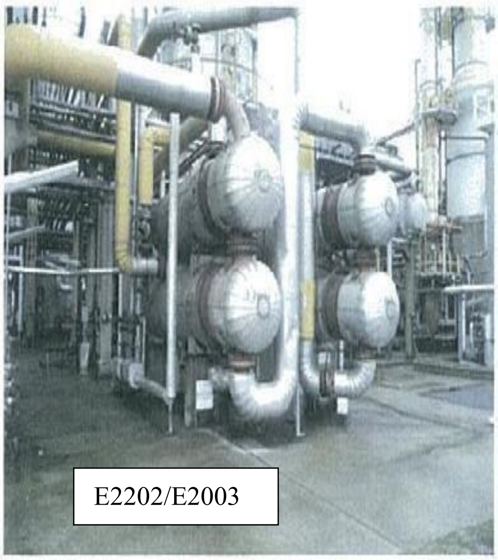
Page

Distillate hydrodesulphurisation plant unit 2000

The Naphtha hydrodesulphurisation plant was originally built as a dewaxer unit in 1974 and later changed to a hydrodesulphurisation plant. It was last revamped in 2009 and is based on Mobil and Albemarle designs/technology. It has a capacity of 15,000bb1/day.
HGO, MGO and LGO from the crude unit is fed to the plant through a coalescer D2001 to the charge stripper deaerator D2008. D2008 operates at 7 bar and 146C. Off gases are cooled and sent to the amine plant 2200.

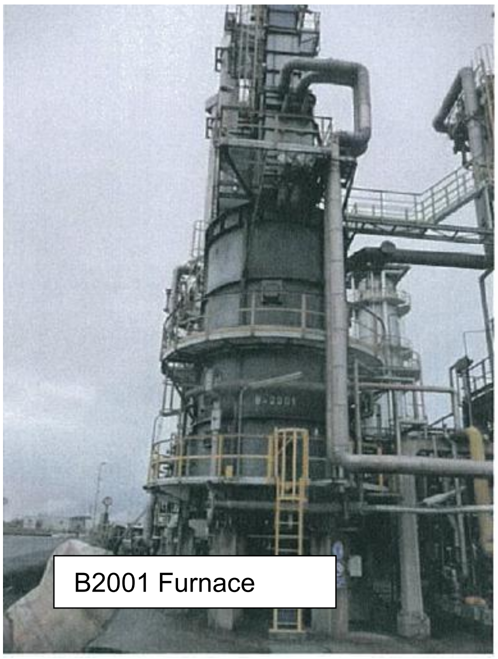


D 2008 deaerator



E2202/E2003

The charge feed from the deaerator is preheated through a series of heat exchangers to 320C and then further heated to 380C in the furnace B2001. B2001 has a design duty of 7.9 MW.



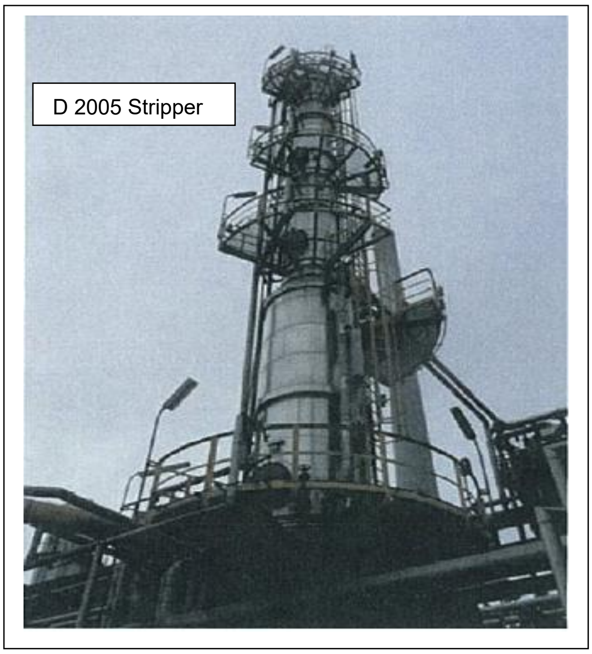
B2001 Furnace

G2007 recycle gas

The feed is mixed with H_2 rich gas from the recycle compressor and make up from other plants and fed to the reactor at 38 bar pressure and 380C. The reactor D2002 contains 55 m^3 of catalyst and is 2.6 m in diameter and 11.8 m tall.

D2002 Reactor Pre heat exchanger

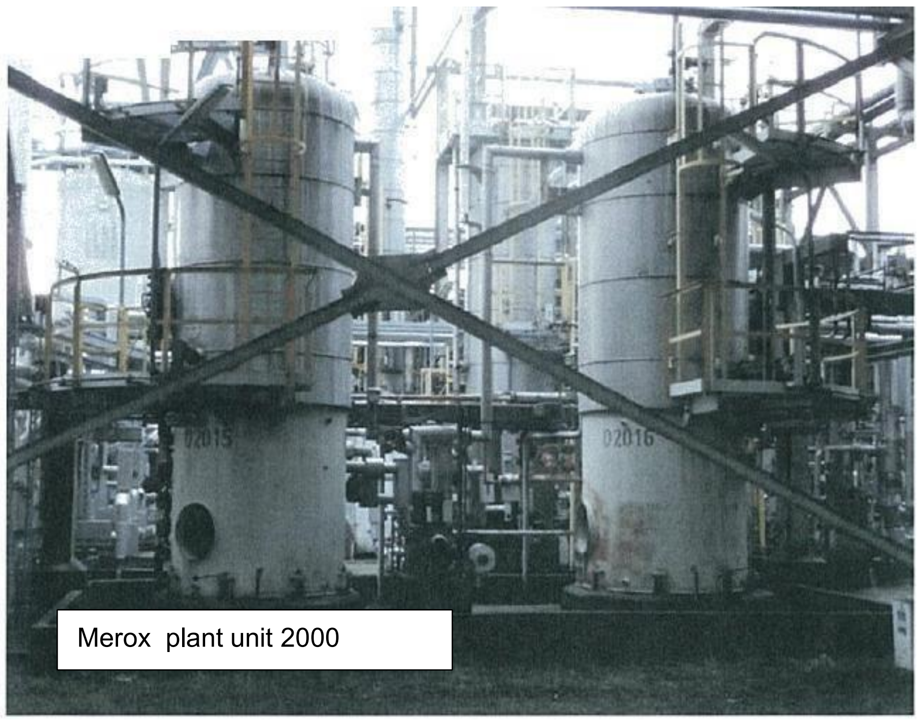
The reacted product is cooled in the preheat heat exchangers and sent to the high temperature separator. The H_2 rich gases are fed through the low temp separator to the recycle gas compressors G2006 or G2007, which raises the pressure back to 44 bar. The liquid product stream from the high temperature separator is fed to the middle of the product stripper D2005.



D2005 has 20 plates is 13.1m tall and operates at 5.4 bar. Temperature across the column is 243C to 114C. The stripper is fed with MP steam. The overheads from the stripper are cooled in the fin fan coolers E2007 and sent to the overhead accumulator drum D2006. The uncondensed fuel gases are sent to the amine plant for further processing. The unstabilised naphtha liquid is partially recycled to the stripper and the rest pumped back to the crude tower and to the naphtha stabilizer D2011. The desulphurised gas oil bottoms off the stripper are cooled in a series of heat exchangers and sent to the 1900 unit feed or to the D1906 stripper.
The naphtha stabilizer column is 1.07m in diameter and 21.5m tall with 31 plates and operates at about 16 bar and a temperature across the column of 90C to 173C. The overheads are cooled with the fuel gases sent to the amine plant and the liquid partially recycled to the column and the rest sent to either the Merox plant or back to the crude tower. There is a reheat pump around on the bottom of the naphtha stabilizer column. The stabilized naphtha form the bottom of the column is cooled and sent to the merox unit**.**

Naphtha stabilizer D2011

The naphtha from the stabilizer is mixed with air fed from air compressor G2016 and 2.4% sodium hydroxide solution and fed to the Merox reactor D2014. This is 1.6 m in diameter and 4.5m tall. It is a fixed bed reactor with activated charcoal packing. It operates at about 7 bar. The caustic solution is recycled back to the reactor and waste water is fed off the bottom of the reactor to waste. The naphtha product sent from the reactor is sent to storage.
The Merox plant removes mercaptan from naphtha.
An extension to the Merox plant (Merichem plant) has been installed recently and was in the stages of final commissioning.



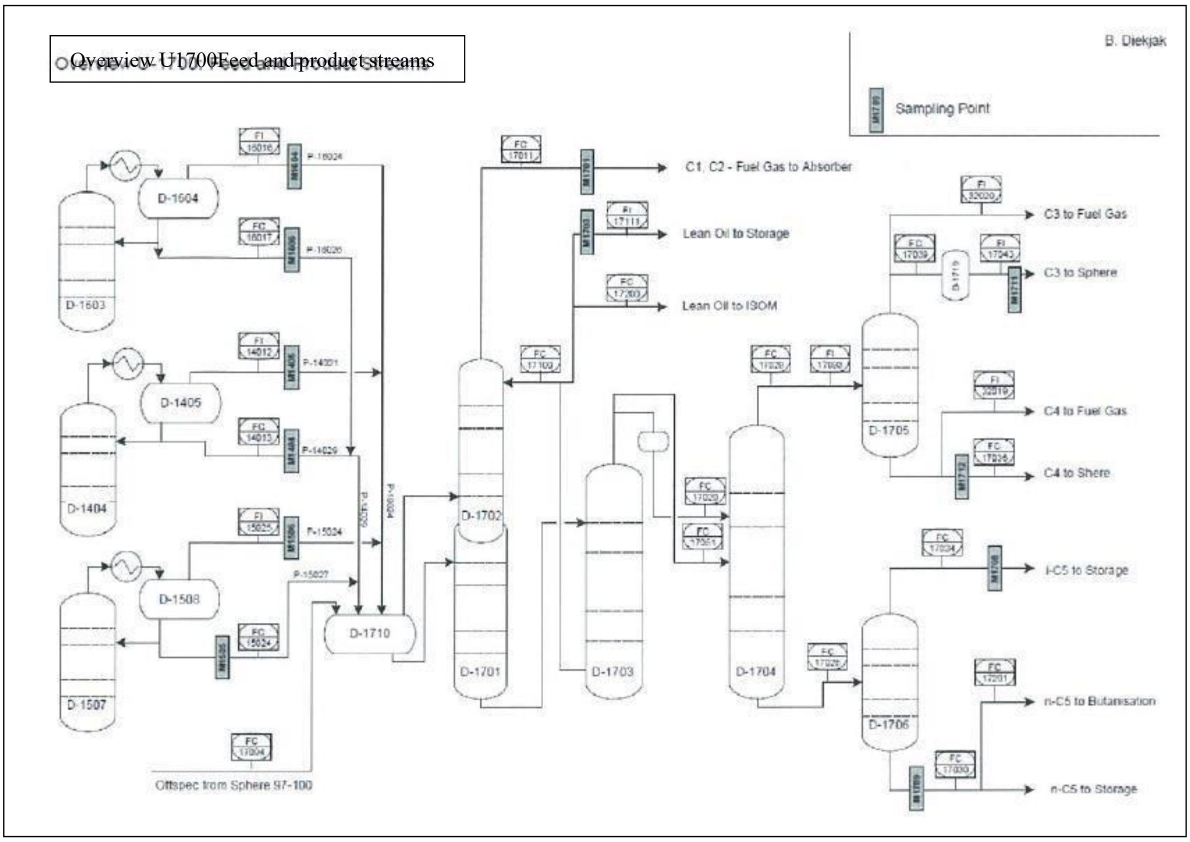
Merox plant unit 2000


New Merox plant

GAS SEPARATION PLANT UNIT 1700

The gas plant was installed in 1974 and last revamped in 1997. It has a capacity of 41,000 bbl/day.
With a feed rate of 125T/hr the plant would typically produce;

C3 11T/hr
C4 30T/hr
IC5 29T/hr
NC5 32T/hr
C6 60T/hr



The lighter ends from the Naphtha pre treater plant and the reformer plant are fed to the gasoline separation plant. This is fed to the deethaniser stripper. D1701. The stripper is 2.8m in diameter and 18.5m tall with 22 plates. There is a reheat bottoms pump around. The overheads are cooled and fed back to the feed surge drum

Deethaniser stripper and absorber (LHS) and debutaniser

Separation drums and debutanizer

Uncondensed vapours pass to the deethaniser absorber and the liquid phase is pumped back to the stripper. The bottoms of the stripper are pumped and reheated through E1706 heat exchanger to the depentaniser column D1703. The deethaniser absorber which is installed an top of the stripper is 1.4m in diameter and 12.3m tall. It has 26 plates and a top recycle stream. The C2- vapours from the top of the absorber are sent to the fuel gas absorber in unit 2200**.**

Propane/Butane amine scrubber
Fin fan air cooler

Depentanise

The depentaniser column D1703 is 3 m in diameter and 27.2 m tall. It has 30 plates and operates at 11 bar and a temperature across the column of 99C to 158C. The C6- off the bottom of the column is partially recycled back to the deethaniser absorber with the rest sent to Chemical naphtha storage. There is a reheat bottoms pump around on the depentaniser which is reheated through furnace B1701. This has a design duty of 15.8 MW. The C5- overheads are partially sent to the debutaniser and partially cooled and fed to the depentaniser reflux drum. Fuel gas is vented off to the fuel gas system and the condensed liquid is recycled to the depentaniser and also fed to the middle of the debutaniser column D1704. D1704 is
3m in diameter and 27.2 m tall with 40 plates. It operates at 11 bar and a temperature across the column of 70C to 125C. The column has a reheated bottoms pump around reheated through heat exchanger E1708 and MP steam. The C4- off the top of the column is cooled and sent to the debutaniser reflux drum. The condensed C4- is partly recycled back to the debutaniser and the rest sent to the propane/butane amine scrubber.
The C5+ plus from the bottom of the debutaniser is fed to the middle of the deisopentaniser D1706. This column is 3.6m in diameter and 45.9m tall with 65 plates. It operates at 1.8bar and a temperature across the column of 57C to 70C. It has a reheated bottoms pump around. The overheads are cooled and partly recycled back to the deisopentaniser and the rest to isopentane storage. The N Pentane from the bottom of the column is cooled and pumped to storage.

Desitopentansier D1706	Desitopentansier D 1705

The C4- from the debutaniser is pumped to the propane/butane amine wash column D1718. This is 2.4m diameter in the lower section and 3.9m at the higher section. It is 16.2m tall and has 2 packed sections. Lean DEA from the amine plant is sprayed above the top packing section. The feed flows countercurrent and exits the top of the column. The rich DEA from the bottom is sent back to the amine plant. The washed C4- is pumped and heated in E1711 to the depropaniser column D1705. This is 2.8m in diameter and 26.7m tall. It has 36 plates and operates at a pressure of 15.5 bar and a temperature across the column of 50C to 99. A bottoms pump around reheat system provides the heating through heat exchanger E1710 using MP steam. Butane from the bottom of the column is cooled and pumped to storage. Propane from the top of the column is cooled and pumped to the fuel

Mass balance of U-1700

An example of a mass balance. of the gas saturation plant is given in Figure 2 and Figure
Further details of all probes are given in a separate document

Mass Balance Saturate Gas Plant U-1700 **SHEET 1 (Bulk Mass Balance)**

Date 27.07.2009 06:00 / 27.07.2009 12:00

				Flow		Temp		Press	Density	corr	Corr. Flow m³/h	Flow kg/h
U1700 (in)												
	M1405	G from D1405	FI14012.pv	10080	TI14023.pv	91,5	PI14011.pv	10,9	2,40	0,7311	7369	1766
	M1404	L from D1405	FC14013.pv	93,5	TC14041.pv	49,3			646,0	0,9559	89,4	5775
	M1506	G from D1508	FI15025.pv	1081,5	TI15059.pv	36,9	PC15014.pv	11,4	1,41	0,8870	959,2	135
	M1505	L from D1508	FC15024.pv	22,0	TI15061.pv	37,2			554,9	0,9795	21,5	1195
	M1604	G from D1604	FI16016.pv	789,5	TI16041.pv	25,7	PC16006.pv	10,4	2,50	0,7537	595,0	148
	M1606	L from D1604	FC16017.pv	0,0	TI16041.pv	25,7			540,8	0,9997	0,0	8
		Off spec repro	FC17094.pv	0,0					0,0		0,0	0
U1700(out)												
	M1701	G to fuel gas	FC17011.pv	4966,6	TI17021.pv	93,5	PC17009.pv	9,2	1,44	0,8364	4153,9	599
	M1711	C3	FI17043.pv	15,8	TI17054.pv	20,3			508,5	0,9640	15,3	776
		C3 to fuel gas	FI32020.pv	6,2	TI17053.pv	31,0			508,5	0,9661	6,0	305
	M1712	C4	FC17036.pv	42,6	TI17046.pv	23,6			576,5	0,9595	40,8	235
		C4 to fuel gas	FI32019.pv	0,0					576,5		0,0	0
	M1708	i-C5	FC17034.pv	19,9	TC17029.pv	30,3			624,7	0,9487	18,9	1178
	M1709	n-C5	FC17030.pv	26,0	TI17038.pv	36,7			641,1	0,9476	24,7	1582
	M1703	Lean oil	FI17111.pv	0,2	TI17014.pv	37,1			682,7	0,9356	0,2	138
		Lean oil	FC17200.pv	30,8	TI17014.pv	37,1			682,7	0,9367	28,8	1968

90213 kg in 87786 kg out 97,31% closure

Heat exchansers	Mix pumps

Mass balance of U-1700

Mass Balance Saturate Gas Plant U-1700 — **SHEET 2 (Material Balance)**

COMPOSITION

	U1700 (in)							U1700 (out)						
	R14012	RC14013	R15025	RC15024	R16016	RC16017	RC17049	RC17011	RC17034	R17036	RC17034	RC17030	R17111	R17200
	G from D 1608	L from D 1608	G from D 1608	L from D 1608	G from D 1604	L from D 1604	off-gas	G to fuel gas	C3	C4	iC5	nC5	Lean oil	Lean oil
	wt %	wt %	wt %	wt %	wt %	wt %	...	wt %	wt %	wt %	wt %	wt %	wt %	wt %
H2	0,00	0,00	1,47	0,00	0,00	0,00		0,23	0,00	0,00	0,00	0,00	0,00	0,00
C1	0,52	0,00	5,48	0,20	0,06	0,32		5,53	0,00	0,00	0,00	0,00	0,00	0,00
C2	7,15	0,00	34,34	4,62	2,08	4,62		62,87	0,00	0,05	0,00	0,00	0,00	0,00
C3	30,52	0,05	37,80	22,38	24,57	28,87		26,32	98,75	2,98	0,03	0,01	0,00	0,00
iC4	12,30	2,58	12,78	18,91	54,09	51,69		0,03	1,22	37,97	0,05	0,01	0,00	0,00
nC4	28,84	12,23	13,83	28,80	7,74	10,75		0,08	0,03	58,31	0,80	0,03	0,00	0,00
iC5	8,32	18,50	3,28	17,09	8,42	2,87		0,78	0,00	0,71	96,95	4,01	3,04	3,04
nC5	10,81	30,83	1,14	7,80	2,20	0,82		3,30	0,00	0,01	2,22	85,88	18,84	18,84
C6+	2,77	34,42	0,09	0,22	0,84	0,07		6,25	0,00	0,00	0,00	10,25	77,70	77,70
total	98,83	100,00	100,00	100,00	100,00	100,01	0,00	98,38	100,00	100,00	100,01	99,99	98,98	98,98

MATERIAL BALANCE

	U1700 (in)							U1700 (out)									
	R14012	RC14013	R15025	RC15024	R16016	RC16017	RC17049	RC17011	RC17034	R17036	RC17034	RC17030	R17111	R17200	Total in	Total out	
	G from D 1608	L from D 1608	G from D 1608	L from D 1608	G from D 1604	L from D 1604	off-gas	G to fuel gas	C3	C4	iC5	nC5	Lean oil	Lean oil			
	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	kg/h	
H2	0,7	0,0	19,8	0,0	0,1	0,0	0,0	13,8	0,0	0,0	0,0	0,0	0,0	0,0	20,5	13,8	H2
C1	80,0	0,0	73,9	23,9	0,9	0,0	0,0	330,7	0,0	0,0	0,0	0,0	0,0	0,0	190,8	330,7	C1
C2	1265,8	0,0	326,5	550,1	31,0	0,4	0,0	3786,3	0,0	0,0	0,0	0,0	0,0	0,0	2175,7	3786,3	C2
C3	5381,6	375,4	507,4	2671,9	365,7	3,3	0,0	1217,3	10684,8	304,8	3,5	1,6	0,0	0,0	9314,3	12211,8	C3
iC4	2173,8	1490,0	172,4	2258,7	805,1	4,0	0,0	1,8	132,1	6584,5	5,9	1,6	0,0	0,0	6604,8	6725,9	iC4
nC4	4706,8	7057,1	186,8	3441,5	115,2	0,8	0,0	4,8	2,9	10084,7	94,3	4,7	0,0	0,0	15507,9	10191,4	nC4
iC5	1468,2	11261,4	44,3	2042,2	125,3	0,2	0,0	46,7	0,0	167,8	11430,2	604,5	5,0	716,5	14942,5	12960,8	iC5
nC5	1874,5	17806,1	15,4	932,1	32,7	0,1	0,0	187,7	0,0	3,5	261,8	12267,4	23,7	3098,1	20240,8	17710,1	nC5
C6+	885,5	19877,8	1,2	26,3	12,5	0,0	0,0	374,4	0,0	0,0	0,0	1821,9	107,2	15284,5	20580,4	17396,1	C6+
															87192,6	87743,8	
															97,29	% closure	

Amine plant view

Amine plant PFD

Fuel gas absorber D2205 and amine reactivator D2206

Gases containing H_2S from various units across the refinery are piped to the Amine plant where the gases are washed with DEA. The gases will also contain some Cl and C2 gases

These gas streams are fed to the fuel gas absorber D2205. This has 20 plates and is 1.3m in diameter and 16.5m tall. It operates at 4.5 bar. The lean DEA solution at 15% wt DEA is pumped to the top of the column counter-current to the flow of gases. Any gases are taken off the top of the column to fuel gas. The washed gases from the bottom of the column are fed to the rich amine settler drum D2208 and then pumped through a series of heat exchangers to the top of the amine reactivator column D2206. This column is 2.4m in diameter and 17.7m tall with 20 plates. The column operates at 0.9bar and a temperature across the column of 90C to 120C. Vapours off the top of the column are cooled and sent to the reactivator reflux drum. Part is recycled to the column and the H2S gases are sent to the Sulphur unit 2300. Separated sour water is sent to the sour water stripper. Recovered amine soln from the bottom of the column is reheated in the reactivator reboiler fed with MP steam. Some of the amine solution is pumped back through heat exchangers to the fuel gas absorber.

ORIGINAL DESIGN CLAUS UNITS

In the table below the original design capacities of the two sulfur plants are given including the amounts of amine acid gas and its original design compositions:

Table 2.1 : Original SRU feed gas specification

		SRU-2300 B	SRU-2300 A
Sulfur capacity	t/day	50	10
Amine Acid Gas Composition	kmol/h		
H_2S C_2H_6 H_2O		65.38 2.27 5.2	13.10 0.45 1.04
Total flow Total sulfur	kmol/h kmol/h	72.84 68.56	14.59 13.73
Temperature Pressure	°C barg	50 0.6	50 0.6

Feed gas

Feed gas to the SIRIU consists of amine acid gas from regeneration_ SM off gas from The sure water stripper is fed to the incinerator burner.

[In the H2S knock-out drum the entrained water is separated from the acid gas_

Thermal Stage

The .air blower supplies the air to the burner. The air to the burner is exactly sufficient to accomplish the complete oxidation of all hydrocarbons present in the feed gases and to burn as much H_2S as required to obtain a H_2SO_4 to SO_2 ratio of 211 at the outlet of the 3rd·
reactor.

The combustion air to the burner consists of two parts: a feed forward and a feed back part.

The required quantity of air is controlled in ratio with the acid gas flow and multiplying this flow (feed forward control). The ·Now control system is also adjusted by the H_2SO_4 analyzer controller (teed back control) located in the process line downstrearn of the 3rd catalytic stage. It ensures the required H_2S to SO_2 ratio in the process gas, in order to obtain the obtain the optimum sulfur recovery efficiency of the unit.

To remove the heat generated in the burner the process gas passes through the tube bundle located in the waste heat. Boiler feed water is introduced to the shell side of the waste heat boiler and generating low pressure steam la saturated. The process gas is cooled further in the thermal reactor off-gas cooler, so that sulfur in the process gas can be condensed on removed from the stream. Liquid sulfur from the thermal reactor off-gas cooler is directed to the sulfur day pit.

Claus stages

The gas stream from the thermal reactor off-gas cooler is heated again in the reactor reheater, to obtain the optimum temperature for the catalytic conversion in the reactor # 1 reheater reactor, the H_2S and SO_2 in the process gas react over the catalyst until equilibrium is reached. More over the temperature. In the reactor is such that a favorable COS and CS_2

Sulfur plant view

conversion is achieved. The process gas from the reactor passes into the reactor #1 off-gas cooler and is cooled so that the sulfur in the gas is condensed and removed from the sulfur from the reactor #1 off-gas cooler directed to the sulfur day pit.

The gas stream from the reactor #11 off-gas cooler is healed again in the reactor #2 reheater, to obtain the optimum temperature for the catalytic conversion in the reactor #2. In the reactor, the H_2S and SO_2 in the process gas react over the catalyst until equilibrium is reached. The process gas from the reactor passed into the reactor #2 off-gas cooler and is cooled so that the sulfur in the gas la condensed and removed from the stream. Liquid sulfur from the reactor #2 off-gas cooler is directed to the sulfur day

The steam flow to the reactor #3 reheater adjusted by a temperature controller at the inlet of the reactor. The inlet temperature is lower than in the first reactor to promote a. high conversion of H_2S and SO_2 into sulfur.

The process gas from the reactor #3 flows to the downstream of the gas knock out drum, provided with a demister pad, in which the last traces of liquid sulfur are separated from the gas and the liquid sulfur is send to sulfur day pit.

The tail gasses from both Claus trains are routed to incinerator B-580 1/A via a common tail gas header. The off gasses from the sulfur pita from both trains are also routed to the incinerator via this header.

2300B Sulfur plant

B2301 burners

SOUR WATER STRIPPER UNIT 2400

HIGH PRESSURE PART

Sour water from the upstream refinery facilities is fed to the sour water surge drum D-2401. Dissolved gases and hydrocarbons are permitted to flash out to the indicator under Pressure control. Liquid hydrocarbons will settle in a top layer and are separated via an overflow Raffle. The sour water is pumped out of the surge drum on flow control by SWS feed pumps G-2401. The sour water is purnped through the feed/effluent exchanger E-2401 where heat is exchanged with the stripped water stream from the bottorn of sour water stripper D-2401.

In the sour water stripper hydrogen sulfide and am ammoniac are stripped from the sour watersteam produced by reboiler E-2402. Alternately direct MP steam injection can be used The sour water stripper incorporates in total 16 sieve trays with single straight down comers. The feed enters at the top tray. H_2S and NH_3 rich gas is mixed with the flash gas from D-2401 and is routed to the incinerator.
The stripped water is discharged to the desalter feed water tank F-2401 by stripper bottams pump G2405A.

LOW PRESSURE PART

Stripped water is pumped from the desalter feed tank F-2401 by desalter feed pump G-2402. The sour water is pumped through the feed/effluent exchanger E-2403 where heat is exchanged with the stripped water stream from the bottom of sour water stripper D-2402. The majority of the sour water is discharge to the desalter unit. The surplus is mixed with the sour water from the desalter unit. The mixed stream is fed to the sour water surge drum D-2405. Dissolved gases and hydrocarbons are permitted to flash out to the incinerator under pressure control. Liquid hydrocarbons will settle in a top layer and are separated via an overflow baffle. The sour water is discharged from the surge drum on flow control by maintaining a sufficiently high pressure.

In the sour water stripper hydrogen sulfide and ammonia are stripped from the sour water by MP steam, which is directly injected into the stripper column. The sour water stripper incorporates in total 16 sieve trays with single straight down comers. The feed enters at the top tray. H2S and NH3 rich gas is mixed with the flash gas from D-2405 and is routed to the incinerator.

The stripped water is discharged to the water treatment unit by stripper bottoms pump G2405 after excess heat is removed in air cooler E-2404.

Sour water plant Unit 2400

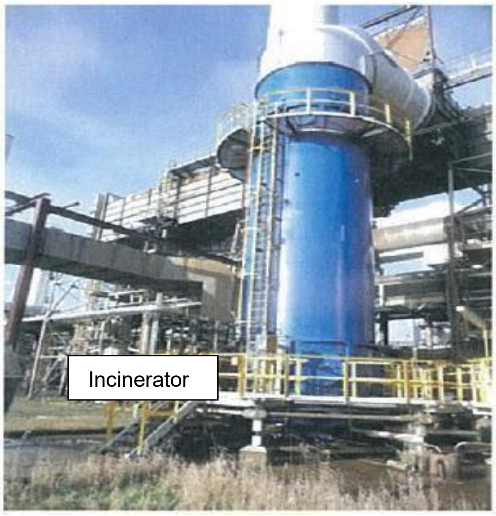
GENERAL SITE FACILITIES/UTILITIES

The Refinery's utility systems are as follows:
Utilities systems capacities
Design Capacity
Two steam boilers 2 x 110 t/h 40barg
Boiler feed water 3 x 240 t/h;
Cooling water 3,400 m^3/h 1700 22 °C
Nitrogen (own production) 295 Nm^3/h 8-10 barg
Nitrogen (vaporizer from N2-tank) 8000 Nm^3/h 8-10 barg
Instrument Air 2,200 Nm^3/h 7 barg
Power Import Facility 36 MW 20 / 6 / 0.4 kV

The high pressure (40barg) steam for the Refinery is generated in two boilers, each capable of producing 120 t/h of steam, although the normal load is 50 t/h. There are three steam levels: 40barg, 10barg and 3barg. One boiler is dual-fired with refinery gas and residue. The other boiler is gas-fired. The type of fuel burned in the boilers is dictated by prevailing economics and may be refinery gas (methane and ethane), residue, propane or butane.
Electricity is purchased from EWE. It is delivered by two main trunk lines and is stepped down to 20kV in two onsite transformers leased by the Refinery from EWE. There is an emergency back up from a battery system for administrative requirements. The wastewater treatment plant includes three API separators, surge tanks and two biological treatment units. Cooling water for the Refinery is obtained from local suppliers and is treated in a series of contactors and filters. The Refinery utilises a single forced-air circulation cooling tower.


Site steam boiler


Feed pumps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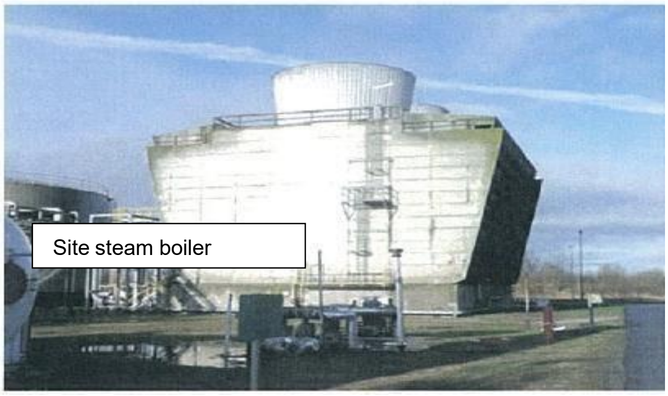
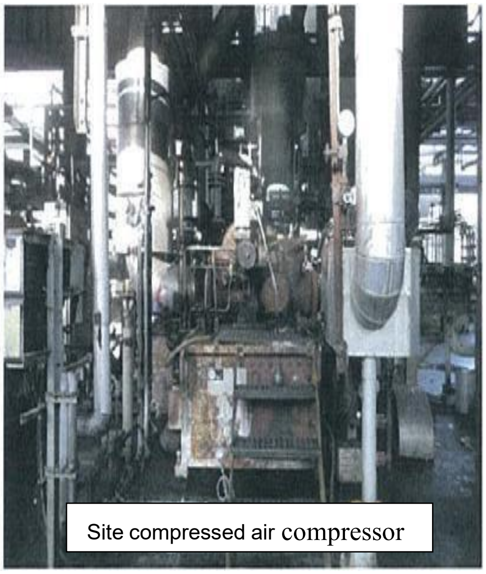

Site steam boiler

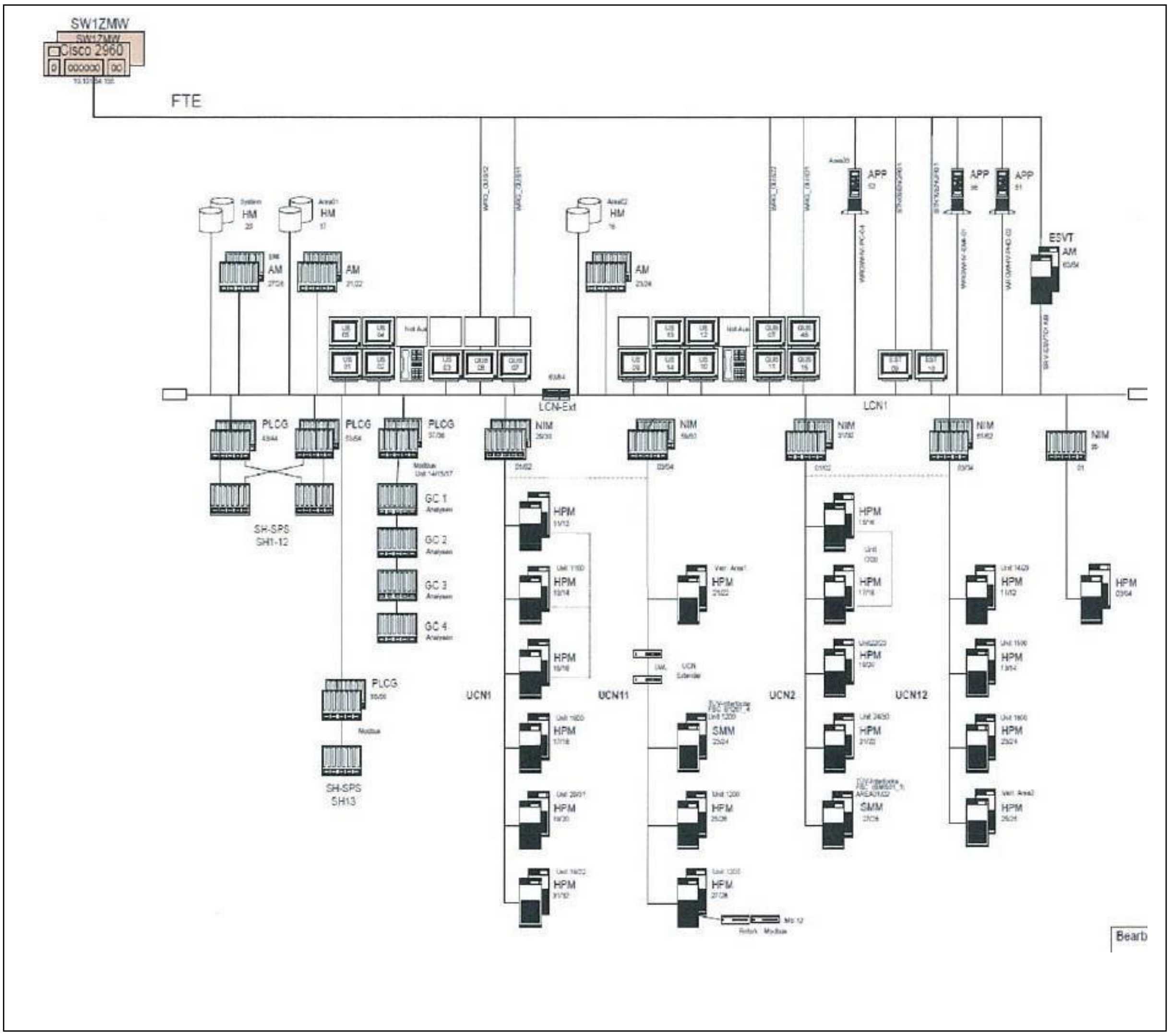


Documentation

- Most of the documents produced since the commissioning of the refinery have been saved as hard copy.

- The most important ones are also available in electronic format:
  PFDs, P&IDs, Isometrics & specification sheets
- Operating manuals and operating procedures are available in German
- On a regular basis, test runs were performed on the different units.
  Records are available as electronic files

APPENDIX 2 — Control systems



APPENDIX 3 – TURNAROUND WORK/COSTS

year	T/A	T/A costs	T/A days	Cat change costs [2]	Cat change days	Economic downtime days	Maintenance downtime days	TOTAL Tunaround & Catalyst Change Out
2010	T/A post	3.519 [1]	0	0	0	Jan-Apr	May- now (Fire-Repair)	3.519
2009	T/A	41.247	Oct - Dec 76 days	8.356	May (16 days)	Dec (14 days)	-	49.603
2008	T/A-preparation	3.107	0	6.406	Sep-Oct (20 days)	Oct (10 days)	-	9.513
2007	-	0	0	1.961	Aug (26 days)	-	-	1.961

year	T/A	T/A costs [3]	T/A days	Cat change costs [2,4]	Cat change days	Economic downtime days	Maintenance downtime days	TOTAL Tunaround & Catalyst Change Out
2006	T/A	6.788	Sep-Oct (28 days)	1.120	Sep-Oct (28 days)	Oct (5 days)	-	7.908
2005	Insp.	3.644	0	0	Jun-Jul (21 days)	-	-	3.644
2004	Insp.	1.955	0	0	Nov-Dec (20 days)	-	-	1.955
2003	T/A	11.180	Jun-Aug (43 days)	0	-	-	-	11.180
2002	Insp.	1.251	0	0	Jun-Jul (11 days)	-	-	1.251
2001	Insp.	3.118	0	0	-	-	-	3.118
2000	Insp.	2.261	0	0	-	-	-	2.261
1999	Insp.	1.341	0	0	-	-	-	1.341
1998	Insp.	1.271	0	0	-	-	-	1.271
1997	T/A	5.941	May-Jun (41 days)	0	-	-	-	5.941

APPENDIX 4 _ CAPITAL PROJECTS

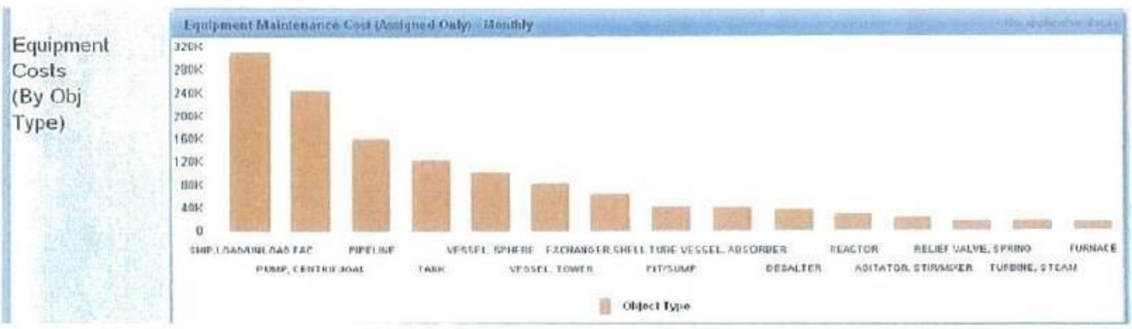
WILHELMSHAVEN REFINERY CAPITAL PROGRAM
EUR THOUSAND

Project	Category	2011 Long Range Plan: 2011	2012	2013	2014	2015	Total	Prior Yrs	Total	ERM Report: EUR MM	Description
VRU Jetties (completion of island jetty VRU only)	Non-Discretionary	15.469					15.469	1.011	16.480	18,0	Offshore VRU
TA-Luft Tank & Pump Programs	Non-Discretionary	1.770	1.828	1.254	966	552	6.370	6.094	12.464	13,2	Installation of tertiary seals at tanks
Subsequent decision to handle as operating expense.							0	0	0	5,5	Upgrade of pumps
Waste Storage Area	Non-Discretionary	510					510	0	510	3,5	Inspection & repair work to reduce leakages and prevent mixing of wastewater streams
Waste Water Unit	Non-Discretionary	1.416	703	697			2.816	0	2.816		
Replacement of High Voltage Motors at CDU	Discretionary	95					95		95	0,0	Not in ERM report
Small Projects	Discretionary	770	770	769	765	761	3.834	N/A	3.834	1,5	Overfill protection systems on gasoline tanks
Tank Repair Program	Discretionary	1.835	2.159	1.394	1.380	1.379	7.947		7.947	0,0	Not in ERM report
Motor Control Centers in Switch Houses	Discretionary	1.062	703	453			2.218		2.218	0,0	Not in ERM report
IT-Budget	Discretionary	319	70	70	69	69	597		597	0,0	Not in ERM report
Oil Accounting	Discretionary				248		248		248	0,0	Not in ERM report
Terminal Conversion	Discretionary	7.695					7.695		7.695	0,0	Not in ERM report
Jetty Infrastructure (piping, loading arms)	Discretionary	283					283		283	0,0	Not in ERM report
Upgrade VGO tank roof	Discretionary	177					177		177	0,0	Not in ERM report
Forecast		31.191	6.233	4.637	3.428	2.761	48.249	7.105	55.354	40,7	Total - Refinery & Terminal
											Refinery Only Projects
										2,5	Replacement of burners with low Nox burners
										1,5	Replace pentane sphere safety valves
										2,5	Inspect and repair pipe work
Number of Tanks in Program		6	3	3	2	5	19			47,2	Total - ERM Final Report January 2011 - Page V

Information Memorandum
Table 5-7: Terminal Financial Projections, page 64

	2011	2012	2013	2014	2015
Sustaining	8.037	6.233	4.637	3.428	2.761
Terminal Conversion	23.154	0	0	0	0
Subtotal (tie to LRP)	31.191	6.233	4.637	3.428	2.761
Subsequently Removed for LRP - Terminal Conversion	450	0	0	0	1.400
Subsequently Added for LRP - Sustaining	-5.500	-3.600	-2.000	-700	0
Total (tie to IM)	26.141	2.633	2.637	2.728	4.161
CAPEX - Sustaining	2.537	2.633	2.637	2.728	2.761
CAPEX - Terminal Conversion	23.604	0	0	0	1.400

APPENDIX 5 _ MAINTERNACE COSTS



Daly production level in January 2009