CHALMERS

Pinch analysis at Preem LYR II

Modifications

ANDERS ÅSBLAD
EVA ANDERSSON
KARIN ERIKSSON
PER-ÅKE FRANCK
ELIN SVENSSON
SIMON HARVEY

Department of Energy and Environment
Division of Heat and Power Technology
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Department of Energy and Environment
Division of Heat and Power Technology
Chalmers University of Technology
SE-412 96 Göteborg
Sweden
Telephone: + 46 (0)31-772 1000

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EVA ANDERSSON, PER-ÅKE FRANCK, ANDERS ÅSBLAD, KARIN ERIKSSON, ELIN SVENSSON AND SIMON HARVEY

Department of Energy and Environment
Division of Heat and Power Technology
Chalmers University of Technology

ABSTRACT

This energy inventory and pinch analysis of the Preem, Lysekil refinery is a part of the Preem – Chalmers research cooperation and has been carried out by CIT Industriell Energi AB. This report is Part II of the report “Pinch analysis at Preem LYR”. The aim with the first part was to supply the researchers at Chalmers with energy data from the refinery in a form that is suitable for different types of pinch analysis. Furthermore, the aim was to make an analysis to establish the possible energy saving potentials in the refinery at various levels of process integration constraints.

In this report, “Pinch analysis at Preem LYR, Part II”, we have applied pinch analysis methods such as the “Matrix Method” and “Advance Composite Curves” to find concrete improvements in the heat recovery network.

The process units of the refinery have a net heat demand of 409 MW (for the operation case studied) which is supplied by firing fuel gas. Steam is generated in the process by cooling process streams. Most of the generated steam is used in the process units (167 MW) and the remainder (17 MW) is used for other purposes.

The energy saving potential, that is the theoretical savings that are achievable, depends on the constraints put on the heat exchanging between process streams in the refinery. Three levels have been analysed:

A: There are no restrictions on the process streams that may be heat exchanged in the refinery. In this case the minimum heat demand is 199 MW giving a theoretical savings potential of 210 MW.

B: All streams within each process unit can be exchanged with each other, but direct heat exchange between process units is not permitted. In this case the minimum heat demand of each process unit must be calculated. The total savings potential, 146 MW, is calculated by adding the savings potential for the separate units.

C: Heat exchange between process units is allowed for those streams which are heat exchanged with utility today (e.g., steam, air, cooling water). However, it is not allowed to modify existing process to process heat exchangers. The scope of the analysis is limited to only consider the 5 largest process units. This group of units are using ~90 %, 363 MW, of the added external heat. It is possible to reduce the external heat demand with 57 MW to 306 MW.

In this report, part II, we give results of possible modifications identified in two process areas, ICR 810 and MHC 240. These areas were selected for further analysis due to their large energy savings potentials. Another area with high potential was CDU+VDU. However, improvements in this area were made during the 2013 turnaround.

To reach the savings potential calculated in Part I, a Maximum Energy Recovery (MER)-network must be constructed. This will however involve a large number of new and modified heat exchangers. It is unlikely that a MER design would be economical in a retrofit situation. Therefore, the trade-off between capital costs and energy savings in a retrofit situation must be evaluated. However, this analysis is not yet done.

The modifications suggested in this study include different levels of increased heat integration. The result of the suggested modifications is presented in the table below.

Modification in ICR 810	New heat exchangers	Heat supplied by H-8101 [MW]	Heat supplied by H-8120[MW]	Steam production [MW]
Present situation	–	4.2	40.5	26
1 Use heat from R-8102 to heat fractionator feed (generate less steam)	1	8	19	12
2 Split exit stream from R-8102 to enable improved heat recovery	1	18.5	0	0
3 Radical makeover	9	6.6	0	0
Modification in MHC 240	New heat exchangers	Heat supplied by H-2401[MW]	Heat supplied by H-2403[MW]	Savings [MW]
Present situation	–	17.1	9.4	
1 Use heat in flue gases to heat feed to T-2408	1	17.1	6.1	3.3
2 Use heat currently removed in air heat exchangers to heat the cold feed to unit	5	6.6	9.4	10.5

Key words: *Pinch analysis, Process integration, Stream data extraction*

Contents

1 Introduction

This report considers a process integration project within the Preem – Chalmers research cooperation. In the beginning of 2013, the project issued a report concerning the Lysekil refinery titled "Pinch analysis at Preem LYR". The material covered by that report will hereafter be referred to as Part I. It forms the basis for the continuing project work that is the topic of this report and that correspondingly will be referred to as Part II.

Before the main findings from Part I are summarised in section 2, some essentials concepts in Pinch Analysis are presented below.

1.1 Basic concepts used in Pinch Analysis

Pinch Analysis is based on the concepts of streams and composite curves. From an energy or heat recovery point of view, a process consists of streams that either undergoes heating or cooling. A stream is characterised by a start temperature, a target temperature and a heat load. Streams that needs to be cooled are called *hot streams* (regardless of absolute temperature), and streams that needs to be heated are called *cold streams*.

If all hot streams are combined into one hypothetical stream (with respect to temperatures and loads), the so called *hot composite* is obtained. Similar, the *cold composite* is obtained by combining all cold streams. The composites represent the accumulated cooling and heating demands. If the composites are plotted on a *temperature versus heat load* graph, the so called composite curves are obtained.

From the composite curves, the maximum thermodynamically possible amount of heat recovery can be identified. The curves are separated by the minimum temperature difference, which is the minimum approach temperature for heat exchanging. This location is called the *pinch*. A low temperature difference (small temperature approach) increases the possibility for heat recovery, thus lowers the utility demands, but increases the required heat exchanger area.

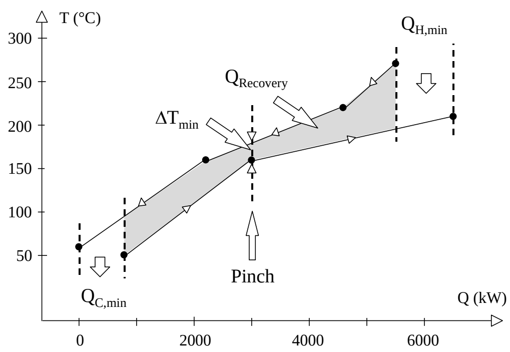


Figure 1 Composite curves

The pinch divides the system into two parts. Above the pinch, we have a heat deficit area, while below the pinch we have an area with heat surplus. Therefore to obtain a system with minimum utility usage we shall not we violate the *pinch rules*, such as; *we shall not place a cooler above the pinch*. Cooling of the hot streams above the pinch shall be accomplished by process-to-process heat exchange. Analogous, *we shall not place a heater below the pinch*. Heating of the cold streams below the pinch shall be accomplished by process-to-process heat exchange. Additionally, *we shall not transfer heat downward through the pinch*.

The *grand composite curve* – also called the heat surplus diagram – shows the *net heating or cooling demand* on a temperature scale.

2 Background: Project Part I

This section describes the approach used and the main results obtained in Part I and documented in the report *Pinch analysis at Preem LYR* (Andersson, et al., 2013).

The Preem refinery in Lysekil has a capacity of processing about 11.5 million tonnes of crude oil, corresponding to about 35 000 m^3/day. The plant is organized into 18 different process units. Service areas and tank farm are not included in the inventory or the subsequent analysis.

2.1 Energy inventory

Process flow diagrams for all units of the plants were used to identify streams that were to be included in the energy inventory. Data for these streams was extracted from the following sources:

- Process flow diagrams, PFD
- Screenshots from the process information system
- Internal studies at Preem
- Contact with process engineers at Preem and access to present and historical data from the process information system

All screenshots were taken on the same day, 2010-04-23, and as close in time as possible. At the time, the plant operating conditions was considered stable and representative.

The data was processed and arranged in a format suitable for pinch analysis and within Part I, analysis of the refinery's possible energy saving potential was conducted (see further below). In addition, the comprehensive data obtained from this energy inventory was supplied to researchers at Chalmers for use in related research projects. The outcome of these projects includes one PhD thesis (Johansson, 2013) and one licentiate thesis (Brau, 2013) both presented in 2013

2.2 Energy balances

It was establish that the process units in the refinery had a heat demand of 409 MW. This result relates to the time of the energy inventory and to the operation case studied[1]. In the analysis of saving potentials presented in this report, this will be referred to as the *present* heat demand. The heat demand is supplied by firing fuel gas. Total fuel gas supplied to the process (boilers not included) was 543 MW.

[1] The case was selected by Preem, but we do not have information on the type of crude oil or the product mix at the time for data collection.

Some of the process cooling demand is satisfied by generating steam. The major part of this steam, 167 MW, is used within the process and the remainder, 17 MW, is expanded in backpressure turbines and used for heating purposes outside process. In Figure 2 an illustration of the energy balance for the total refinery is given.

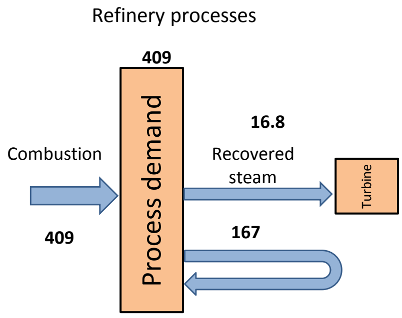


Figure 2 Present heat balance for the refinery

The same type of representation as in Figure 2 has been used for individual process units as well, two examples are shown in Figure 3. It can be seen that the integrated process units for crude distillation and vacuum distillation, CDU + VDU, has a present heat demand of 181 MW. Here, steam corresponding to 36.2 MW is generated. Within the process 6.2 MW steam is used and the rest, 30 MW, is exported to other process units. When establishing the present heat demand for individual process units, no credits are given for steam export. For the process, steam generation is merely a utility cooling.

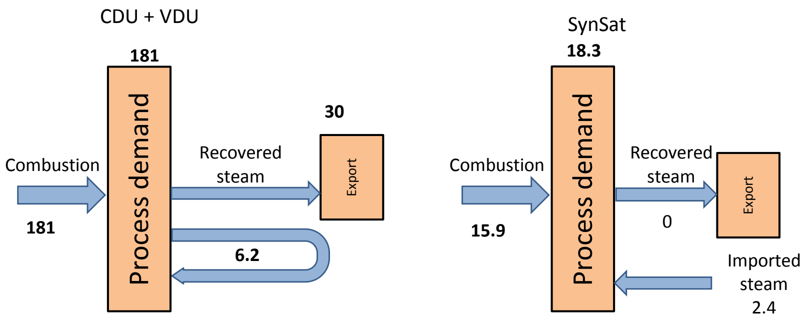


Figure 3 Present heat balances. Left: CDU + VDU; right: SynSat

For the other example, the SynSat unit, there is a deficit of steam and 2.4 MW has to be imported. In the analysis of individual process units, the present heat demand for the SynSat unit will be considered as the sum of heat demand from combustion of fuel gas and the heat demand from imported steam, i.e. 15.9 + 2.4 = 18.0 MW. Consequently, the sum of the present heat demands for all individual process units will exceed the present heat demand value obtained when considering the entire refinery (Figure 2). The difference represents the current level of steam utility integration.

2.3 Energy saving potential at different levels

Theoretical energy saving potentials for the refinery was calculated using pinch analysis. Three cases (levels) were considered and they differed with regards to the constraints applied for heat exchanging between process streams and thus the amount of rearrangement allowed/required in the heat exchanger network. In short, the cases can be described as follows:

Level A — No restriction in heat exchanging between streams within or between different process units. The necessary rearrangements in the heat exchanger networks may be considerable.

Level B No restriction in heat exchanging between streams within a process unit. No direct heat exchange between process units is allowed. The necessary rearrangements in the heat exchanger networks may be considerable.

Level C Streams heated or cold by utility can exchange heat between process units indirectly (using a heating media or utility) or within process units directly. However, all existing process to process heat exchangers are accepted. The rearrangements in the heat exchanger networks should therefore be limited.

2.3.1 Level A

Level A gave a minimum heat demand of 199 MW, see Figure 4. This corresponds to a theoretical savings potential of 210 MW. Given that no restrictions are considered in this case, these figures represent the theoretical upper boundary.

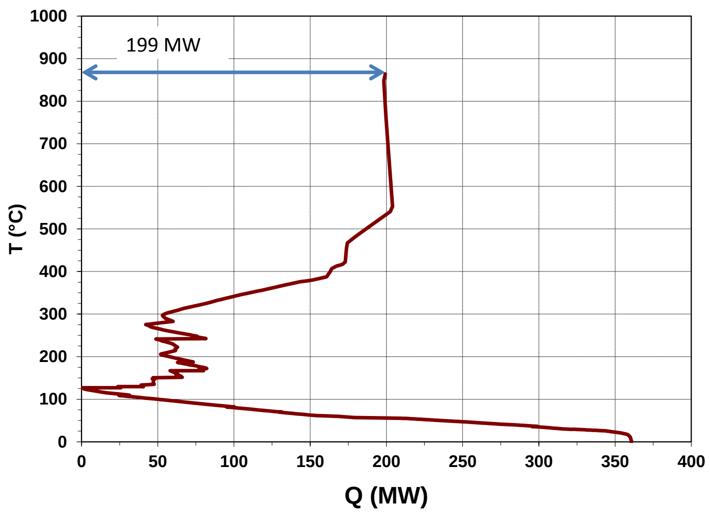


Figure 4 Grand Composite Curve for the processes considered at level A

2.3.2 Level B

At Level B constraints were introduced such that heat exchanging was only allowed between process streams belonging to the same process unit. Consequently each process unit was analysed separately. The difference between the actual and the minimum utility requirements from a pinch analysis equals the sum of pinch violations (see section 1.1), that is:

- heating below the pinch
- heat down through the pinch
- cooling above the pinch

Below the pinch, we have a net heat surplus that could be used for, for example, steam generation. Above the pinch, only process-to-process heat exchangers and heaters are allowed according to the pinch rules, thus steam generation above the pinch (by cooling of process streams) increases the utility demand for the unit.

Some process units use steam for heating purposes. This steam may or may not be generated within the process unit. For the refinery as a whole, the steam production within the process units exceeds the demand (see Figure 2), but in individual units this may not be true. When the actual heat demand for a process is determined, both steam heaters and process heaters fueled by gas are included in the demand. Steam generated in the unit is subtracted from the actual heating demand. Thus, the sum of heating demands for the individual process units exceeds the current consumption for the refinery as a whole.

At Level B the actual heat demand, established in the manner described above, was 504 MW. The minimum heating demand was 358 MW and the savings potential 146 MW.

2.3.3 Level C

At Level C, heat exchange between different process units was allowed, but only for streams which, at the time of the analysis, were heat heated or cold by utilities such as steam, air or cooling water. In other words, heat exchangers not using utility were not to be modified.

The Level C analysis was limited to the five largest process units. The present heat demand for these units was summarized to 363 MW, i.e. almost 90% of the present heat demand of the entire refinery.

The heat recovery opportunities for Level C can be identified in Figure 5. The source and sink curves in the figure are the composite curves for the stream segments heated and cold by utilities. The source curve includes cooling demands satisfied by generating steam and also fictive flue gas coolers representing the possible cooling of flues gases down to 125°C. The sink curve includes heat demands currently heated by steam. (Steam is generated both in the process and in boilers at the refinery.) The current level of heat recovery via the steam system is represented by the overlap between the source and sink curves at a hot utility demand of 363 MW.

In Figure 5, the minimum hot utility demand is given as 306 MW. Thus the saving potential amounts to 363 – 306 = 57 MW.

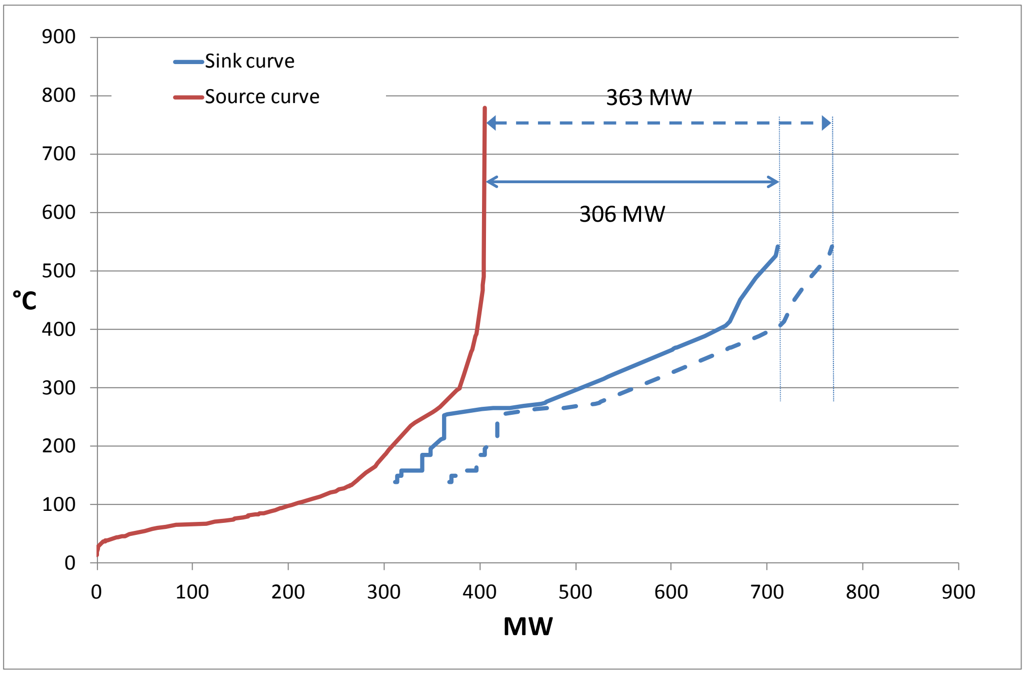


Figure 5 Present and maximum heat recovery for the five largest process units considered at Level C

2.3.4 Summary of savings potentials for the different heat integration levels

Table 1 summarizes the results for the different levels studied. Note that the given percentage figures are related to different system boundaries and therefore different minimum utility requirements.

Table 1 Summary of identified savings potential at different levels

	Savings potential	
	[MW]	Percentage of current consumption
Level A	210	50%
Level B	146	30%
Level C	57	15%

2.4 Analysis of selected process units

The analysis at Level C focused on five specific process units and it was decided that also the further analysis of Level B should focus on the same units. These units were:

1. Crude distillation and vacuum distillation, CDU + VDU
2. Naphtha hydrotreating, NHTU
3. Catalytic reforming, CRU
4. Mild hydrocracker, MHC
5. Hydrocracker, ICR

As already mentioned, together they made up for almost 90% of the present heat demand of the refinery. It can be noted that the CDU and the VDU are two process units that are already integrated. It was therefore natural to consider them together.

2.4.1 Heating demand for the selected process units

For evaluation and comparison of heating demands as well as savings potential for different process units, results are presented in Table 2. In addition, Grand Composite Curves for all five process units are shown in Figure 6.

Firstly, it can be concluded that the five process units did, besides accounting for the main part of the refinery's heat demand, also include the main part of the energy saving potential. The energy saving potential for the selected units amounts to 135 MW compared to 146 MW for all process units at Level B.

At Level B, each unit was considered separately as only heat exchanging within units was allowed. This implied that, for each unit, the external heat demand from fuel gas and any demand of steam imported from other units must be included, as illustrated in Figure 3.

The units CRU and NTHU, both have a steam deficit and have to import steam from other units at the refinery. From the units CDU + VDU, MHC and IRC, steam can be exported to other processes as the amount generated exceeds the demand within these units. If a steam balance for all five units is considered, there will be a steam surplus of 63.8 – 24.3

= 39.5 MW. Thus, from a refinery perspective none of the steam used in the selected process units needs to be produced in utility boilers.

Table 2 Heat demand and saving potential for five selected process units used for analysis at Levels B and C. Minimum heat demands from pinch analysis.

	CDU + VDU	NTHU	CRU	MHC	ICR	Sum
Present heat demand	180.7	45.4	78.4	36.7	45.6	386.8
supplied by fuel gas	*180.7*	*23.3*	*76.2*	*36.7*	*45.6*	*362.5*
supplied by imported steam	*0*	*22.1*	*2.2*	*0*	*0*	*24.3*
Minimum heat demand	133.0	40.4	61.5	11.4	5.8	252.1
Energy savings potential at Level B for selected units	47.7	5.0	16.9	25.3	39.8	134.7
Recovered steam, exported from units	*28.5*	*0*	*0*	*6.4*	*28.9*	*63.8*

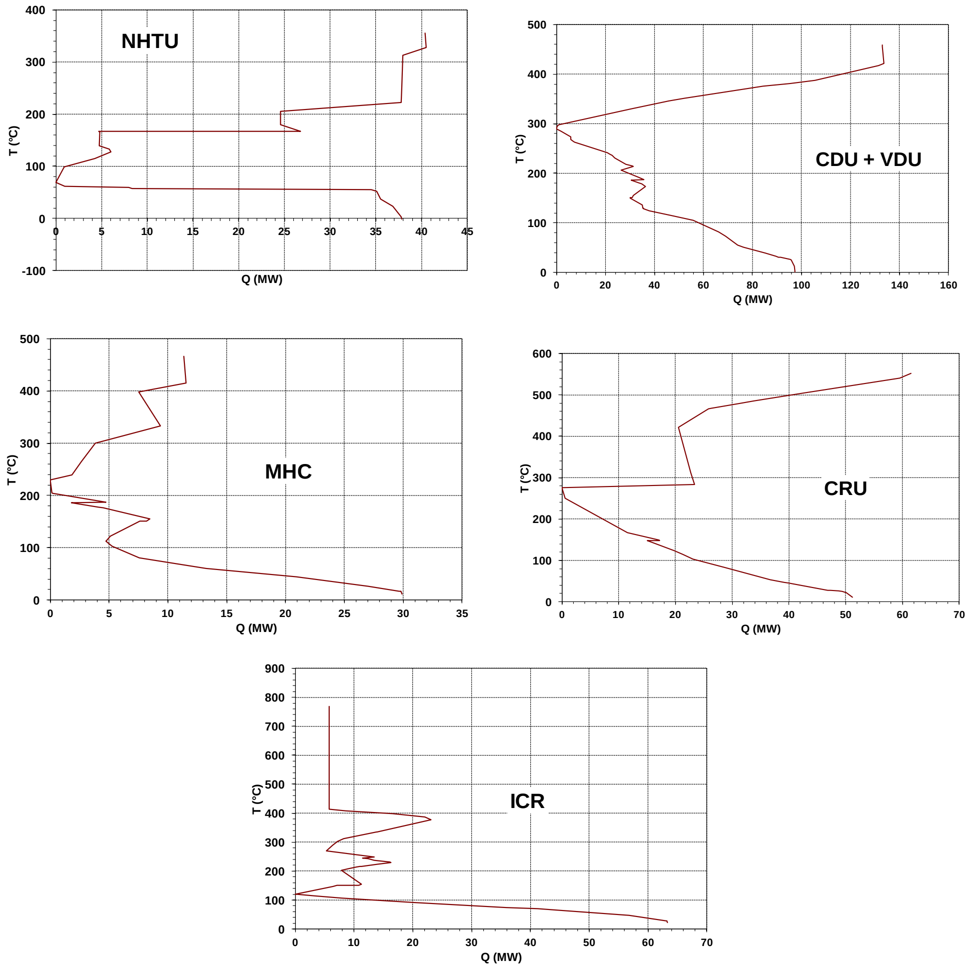


Figure 6 The Grand Composite Curves for the process units with the largest heat demands: CRU + VDU, NHTU, CRU, MHC and ICR.

2.4.2 Quantifying pinch violations for the selected process units

In section 2.4.1, demands and savings potentials were determined by pinch analysis. To identify concrete energy savings measures, the existing heat exchanger network must be analysed.

Thus, as part of the analysis at Level B, all present pinch violations in the present heat exchanger network were analysed and evaluated for the selected units, see Table 3. Considering all five units together, cooling above the pinch is the largest pinch violation, followed by heat down through the pinch and lastly heating below the pinch. Table 3 shows the sum of pinch violations as 148.3 MW.

Table 3 Pinch violations in the five process units identified from the existing heat exchanger networks.

Process unit	Pinch temperature [°C]	Cooling above the pinch [MW]	Heat down through the pinch [MW]	Heating below the pinch [MW]	Total pinch violations [MW]
CDU + VDU	298	10.6	15.6	22.7	48.9
NHTU	69	8.6	0	2.3	10.9
CRU	275	10.0	11.1	2.2	23.3
MHC	230	4.2	14.2	6.2	24.6
ICR	120	32.0	8.6	0	40.6
Sum		65.4	49.5	33.4	148.3

As discussed before, the sum of pinch violations equals the savings potential. The savings potential from Table 3 is 148 MW. This could be compared to the savings potential of 135 MW from Table 2. The difference reflects the precision in extracted stream data. Table 3 is based on heat balances for existing heat exchangers. Very seldom, if ever, heating and cooling loads for a given heat exchanger adds up, due to imperfect instrumentation and estimation of thermochemical properties.

When comparing results for the different units, it was found that the total pinch violations were largest for CDU + VDU, see Table 3. In addition, it was earlier established that the CDU + VDU had the largest present heat demand as well as the largest minimum heat demand, see Table 2.

3 Analysis of Selected of process units

In Part II, we will discuss modifications in the heat recovery system that partly can realise the energy saving potential discussed in Part I.

3.1 Scope and limitations

The suggested modifications presented in Part II are based on the collected stream data during the first part of the project (Part I). Furthermore, the analysis is based on the reasoning behind the “Matrix Method” (Carlsson, et al., 1993) and the “Advanced Composite Curves” (Nordman & Berntsson, 2009). Furthermore, MER networks have been derived.

The analysis is focused on heat recovery possibilities. We have not evaluated the profitability of the proposals or checked whether existing equipment can be reused in the new configurations.

The data are some years old and a snapshot of the refinery at steady state operation. Process demand at start-up has not been considered.

3.2 Selection of process units for further analysis

The three units with the largest savings potentials according to Table 2 are CDU + VDU, ICR and MHC.

The unit CDU + VDU is not chosen for further analysis since it was modified to improve the energy efficiency at the recent (2013) planned maintenance turnaround.

The unit ICR (isocracker) has a notably large energy savings potential of about 40 MW of its present heat demand 45.6 MW (Table 3). This makes it a good candidate for further analysis to find concrete modification suggestions. Also the MHC unit (mild hydrocracker) has presently a large heat demand compared to minimum demand. Thus the units ICR and MHC have been selected for further analysis.

4 Detailed analysis of the Hydrocracker unit ICR 810

The hydrocracker is amongst the newest units at the refinery. It was part of the major refinery upgrade in 2006 called the Gas Oil Project (GOP). The throughput is high and it thus has a large energy demand. Since it is new it might be expected that integration possibilities is already accounted for, but as it turns out this is a type of process where energy saving potentials often can be identified: (Canmet Energy, 2003): ”*PI studies will usually identify energy-saving opportunities on reaction processes (such as hydroformers, hydrotreaters and hydrocrackers). However, much of the equipment in the loops of these processes operates at high pressures and high temperatures; and, in retrofit situations, new equipment and piping changes are often very expensive.*”

That the energy saving potential was substantial was established in Part I. Here, in Part II, we will analyse the possibilities to realise it.

4.1 General process description

This general description is based on (Alfke, et al., 2007), (Speight, 2005), (Colwell, 2009) and (European IPPC Bureau, 2013).

In the Hydrocracker unit sulphur is removed and the vacuum gas oil from the Vacuum distillation unit, VDU, is converted to lighter components in a cracker. Hydrocracking is a catalytic process in which hydrogenation accompanies cracking. The hydrogenation process is exothermic and favoured by high temperatures and high hydrogen partial pressures. Operating conditions in the reactor section are usually about 400°C and 8 – 15 MPa.

Advantages of the hydrocracking include a high flexibility towards product yields. Another attractive feature of hydrocracking is the low yield of gaseous components, such as methane, ethane, and propane, which are less desirable than the gasoline components.

The common process layout includes two reactors in series and equipment needed for separation, fractionation and recycling. The following steps are performed:

1. Water is removed by passing the feed stream through a silica gel or molecular sieve dryer
2. The feed is mixed with hydrogen and preheated
3. The mixture passes through a multi-bed reactor with interstage hydrogen quench for hydrotreating
4. Between the two reactors intermediate cooling is performed and additional hydrogen is introduced
5. The mixture passes through a second multi-bed reactor with quenches for hydrocracking
6. Reactor effluents are cooled and pass through high and low pressure separators. H_2S is separated from the product stream with amine solution.
7. The product stream is reheated and fed into product fractionators where products are drawn from the top, sides, and bottom.

In the first reactor stage, organic nitrogen compounds and organic sulphur in the feedstock are converted to hydrocarbons and to ammonia and hydrogen sulphide by hydrogenation and mild hydrocracking. Most of the hydrocracking is accomplished in the second reactor stage.

The heat generated in the exothermic reaction is partly used to generate medium pressure (MP) steam. However, whether or not steam generation is the best way to use this heat depends on the refinery's steam balance. If there is an excess of steam at times it is energy wise better to use the heat internally in the process and control the steam production in a boiler and thereby reduce fuel gas use.

4.2 Specifics for ICR 810 at Preem LYR

At the time of the energy inventory, the feed to the ICR unit at Preem LYR was about 7500 tonnes/day. In Table 4, the corresponding specific fuel consumption and amount of steam produced is presented together with intervals for these utilities according to a Best Available Technology (BAT) report issued by (European IPPC Bureau, 2013) . As seen, the ICR unit at Preem LYR is found in the middle of the fairly large ranges given in Table 4. Anyway, based on Table 4 there could be room for improvement in ICR810, although the consumption is highly dependent on the specific refinery situation.

Table 4 Specific fuel consumption and steam production in the hydro cracking unit ICR compared with utility consumption of hydrocrackers in BAT report (European IPPC Bureau, 2013)

	Specific fuel consumption [MJ/tonne]	Specific steam production [kg/tonne]
ICR at Preem LYR	620	140
BAT report by European IPPC Bureau 2013	400-1200	30-300

The present heat demand for this unit is 45.6 MW according to the analysis performed in Part I. If we assume that there are no constraints in how heat exchanging can be performed, the potential savings amounts to about 85% or 39.8 MW (see Table 2). If all the heat is produced with fuel gas the potential economic savings can be estimated to 100 MSEK per year[2].

There is quite a difference in pressure for the different equipment in the unit. High pressure implies that investment costs can be high and large difference in pressure between streams might make heat exchanging expensive.

It could therefore be interesting to identify the penalty resulting from forbidding matches between high and low pressure streams. If only streams with the same pressure level can exchange heat with each other, the minimum heat demand is 44 MW and the savings potential is only 1.6 MW for the ICR unit. This is quite a penalty.

Hence the high potential that was established in Part I assume that process streams can be heat exchanged regardless of difference in pressure. Operations like this can be both feasible and motivated. One example in the present design is E-8106 (fractionator preheating with reactor interstage cooling) where the pressures of the streams are 140 bar and 20 bar respectively.

4.2.1 Existing heat exchanger network

The process flow diagram is shown in Figure 7.

[2] Boiler efficiency 80 %, 8400 hours pr yr, 250 SEK/MWh fuel gas

Figure 7 Process flow diagram of the ICR process unit.

Firstly, the feed is heated from 76 to 228°C before the pressure is raised and hydrogen is added. Heat from hot products going to tank is used for this preheating. The corresponding heat exchangers are labelled E-8101, E-8102, E-8103 and E-8104.

Secondly, to raise the temperature to the specified inlet temperature of the first reactor, the feed is heated by the second reactor outlet to 368°C in E-8105 A-C. Lastly, the feed is raised another 20°C in the process heater H-8101.

The cooler between the reactors, E-8106, is heating the product going to the fractionator tower T-8121. In E-8106 the pressure difference between streams is high (on one side the pressure is 140 bar and on the other side 20 bar).

The outlet from the second reactor is cooled in E-8105 A-C (above) from 422 to 300°C. Thereafter, further cooling takes place in E-8107 where MP steam is generated. The process side pressure is 130 bar and the BFW (Boiler Feed Water) pressure is 27 bar. The heat exchangers E-8107 and E-8106 (above) are the only heat exchangers with a pressure difference between streams that exceeds 100 bar.

The separation of products starts by a liquid-vapour separation at a pressure of about 130 bar in V-8102 (the hot high pressure separator, HHPS). The hydrogen rich vapour stream is cooled to 164°C by a heat exchanger, E-8109, and that heat is recovered by preheating hydrogen to the reactors. Wash water is injected before the stream is cooled further in an air heat exchanger, EA-8101, before sent to the cold high pressure separator (CHPS) and further amine treatment (in T-8101).

The liquid from V-8102 is let down to 16 bar fed into V-8103 (the hot low pressure separator, LHPS) where liquid and vapour are separated. The liquid is pumped to product stripper, T-8120. The vapour flows from both V-8103 and T-8120 are cooled by air exchangers, EA-8102 and EA-8120.

After H_2S removal (accomplished by the equipment described above together with V-8105 and amine absorber T-8124) the product stream is reheated in two heat exchangers and process heater. Firstly, from 193 to 219°C with heat from the UCO (unconverted oil) product stream in E-8120. Secondly, heated from 219 to 254°C with heat from the reactor intercooler E-8106, and lastly the product stream is heated from 254 to 368°C in process heater H-8120 and fed to the product fractionator T-8121.

After fractionation the products are cooled. Some of that heat is recovered, as mentioned, by heating the feed and some of the heat is used in side stripper reboilers. (The strippers T-8122 and T-8123, T-8122 and T-8123 are not shown in the PFD, but are located at the T-8121.) Some final cooling is accomplished with air coolers.

4.2.2 Stream data

Stream data for all streams used in the analysis are listed in Table 5. Heat in flue gases leaving the stack has not been included. Steam used as stripper steam is not included as a heat demand but must be included in the overall steam balance of the refinery.

Table 5 Stream data for all streams that are heated or cooled in the ICR 810 unit.

Stream description		Tin [°C]	Tut [°C]	Duty [MW]
R-8101 to R-8102	Hot	422	402	7.9
R-1802 to V-8102	Hot	423	240	92.4
V-8102 OH	Hot	224	164	14.9
V-8102 OH, air cooler	Hot	130	57	23.3
V-8103 OH	Hot	234	75	2.1
T-8120 OH	Hot	82	38	11.2
T-8121 OH	Hot	131	80	27.8
Gasoil pump around	Hot	285	247	16.3
Kerosene to tank	Hot	231	38	5.3
Diesel to tank	Hot	318	43	16.8
UCO to tank and to FCC	Hot	351	162	19.0
UCO to tank	Hot	162	83	2.5
Feed to V-8101	Cold	76	228	32.2
R-8101 Feed	Cold	227	388	64.6
T-8120 to T-8121	Cold	193	368	56.5
V-8105 to T-8120	Cold	60	193	5.3
Hydrogen from compressors to R8101	Cold	93	201	14.0
T-8122 reboiler	Cold	231	233	3.0
T-8123 reboiler	Cold	298	320	3.0

4.2.3 Energy savings potential using advances curves

The so called “Advanced curves” were developed during the first decade of the century. The aim was (Nordman & Berntsson, 2009):

- To identify heat recovery projects that reduce the problem size, and are economically feasible, prior to detailed design.
- To identify temperature levels where usable excess heat can be extracted and used by other processes (confer total sites, eco-cyclic industrial parks).

Here, only the first point is discussed. The curves should be used as a screening tool to identify interesting retrofit alternatives. The advanced curves are four: utility curves, actual load curves, theoretical load curves and extreme load curves. Here, we are going to make use of the actual load curves in a bit simplified manner.

The actual heating load curve (AHLC) is a composite of the process stream segments in the existing heaters. Correspondingly, the actual cooling load (ACLC) curve is a composite of the process stream segments in the existing coolers. In the original method (Nordman & Berntsson, 2009), the advance curves were plotted together on special type of graph. Since we here only will use the actual load curves, we can instead plot them as ordinary composite curves in the same way as the source and sink curves in Figure 5.

The actual load curves for ICR 810 are shown in Figure 8. The AHLC represents the heat added in the unit's two furnaces. The largest part, about 40.8 MW, is added in H-8120 to increase the temperature of the flow to the fractionator. The other furnace, process heater H-8101, adds about 4.8 MW to the reactor feed.

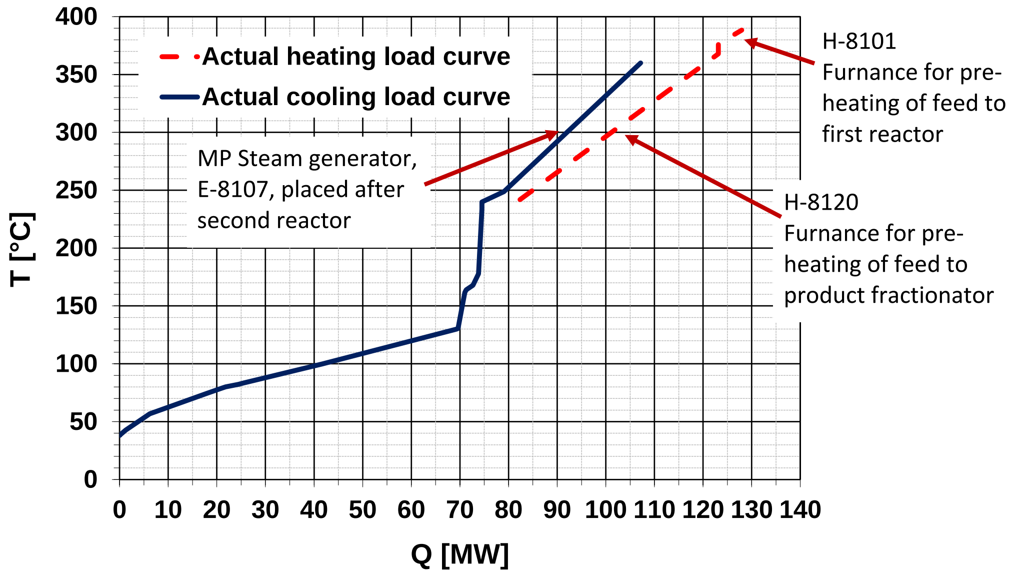


Figure 8 Actual heating and actual cooling load curves for ICR 810.

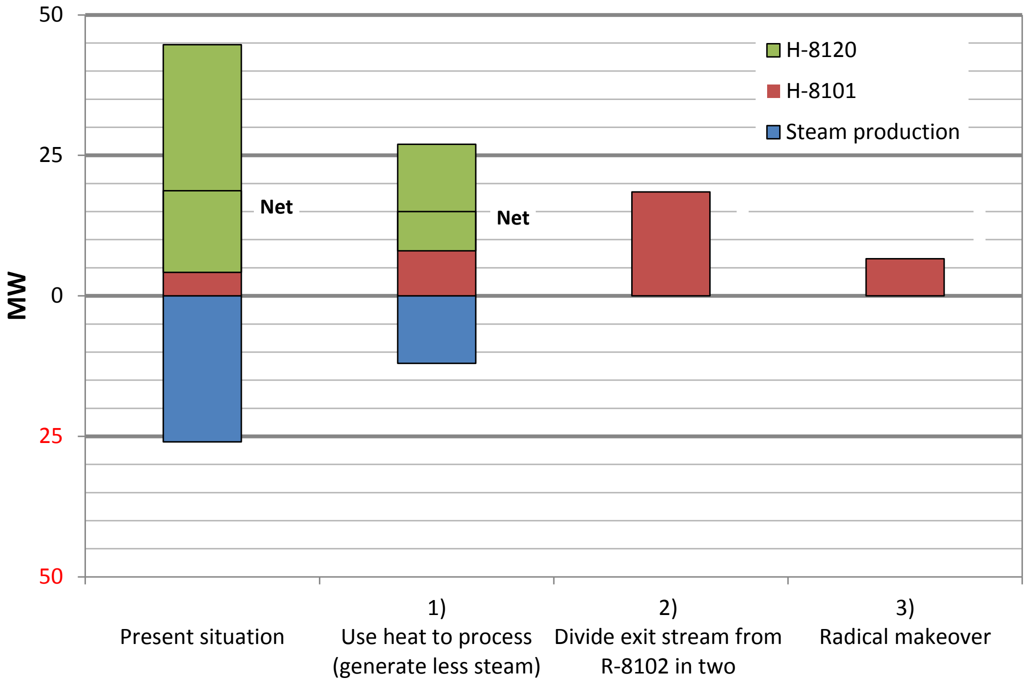
The reactor outlet is cooled by generating stem in in E-8107. This heat could instead be used to partly unload the furnace H-8120. From the overlap between the curves it is found that there is a potential to heat exchange 24.8 MW.

Naturally, the introduction of a heat exchanger here would decrease the amount of generated MP steam. This is commented on in the following section, where the proposed modifications of the process units are presented.

4.3 Possible modifications to ICR 810

A result of the pinch analysis is the minimum heat demand and thus Maximum Energy Recovery (MER). In an unconstrained process, a MER-network can always be obtained if process integration design principles are observed. However, a MER-network is unlikely to be economical in a retrofit situation and there must be an evaluation of the cost of revamp and the savings in energy. For Preem, this evaluation should also include a system analysis for the whole site that considers the reduction in steam production.

For the isocracker unit, we have evaluated three levels of modification with three levels of savings as shown in Figure 9. The modifications are described below.



Modification	New heat exchangers	Heat supplied by H-8101 [MW]	Heat supplied by H-8120[MW]	Steam production [MW]
Present situation	–	4.2	40.5	26
1 Use heat to process (generate less steam)	1	8	19	12
2 Split exit stream from R-8102 in two	1	18.5	0	0
3 Radical makeover	9	6.6	0	0

Figure 9 Three levels of process integration modifications. The horizontal line "Net" shows the heat consumption with credit for steam generation.

4.3.1 Modification 1

This modification decreases the amount of cooling (by generating steam) above the pinch[3]. In Modification 1, Figure 10, the outlet from the second reactor R-8102, presently used for MP-steam generation in E-8107, is used to preheat the feed to the product fractionator T-8121. The fuel gas demand in the process heater H-8120 is thereby reduced. It should be noted that there is a considerable difference in operating pressures between the streams, but a similar match (E-8106) already exists in the unit so this problem is possible to solve.

The heat supplied by processes heaters decreases with almost 18 MW. However, the steam generation is reduced with 14 MW. Thus the net savings – if the steam production is deemed necessary – amounts to about 4 MW. Nevertheless, in line with the discussion in 4.1, steam savings could be advantageous if there is an excess at times.

[3] Cooling above the pinch is a pinch violation, see section 1.1.

Figure 10 Modified process flow diagram of ICR 810, Modification 1

Although only one new heat exchanger is introduced in the network, adjacent exchangers in the network are affected due to changed temperature levels. Based on stream data, new temperatures have been estimated and are shown in Figure 11.

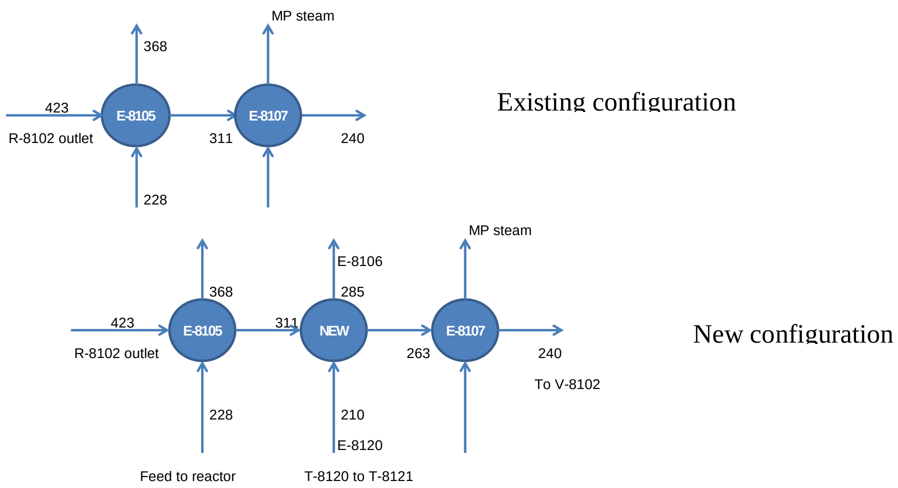


Figure 11 Use part of heat from R-8102 outlet to heat the feed to T-8121.

4.3.2 Modification 2

The feed to T-8121, before the heater H-8120, can reach a higher temperature than in modification 1 if the R-8102 outlet is split into two streams and the new heat exchanger heating the feed to T-8121 is placed in parallel with E-8105 heating the reactor feed, see Figure 12. Better use can be made of the high temperature in the reactor outlet and more heat can be recovered to the fractionator feed. However, the load on E-8105 decreases, thus increasing the load on process heater H-8101, and no MP steam is generated. Nevertheless, modification 2 recovers more heat in total than modification 1. It is even possible to make the heater H-8120 redundant for the case evaluated (as indicated in Figure 12 by the dashed contour), see the table in Figure 9.

Figure 12 Modified process flow diagram of ICR 810, Modification 2

In Figure 13, the new estimated temperature in E-8105 is shown together with the proposed new heat exchanger.

About 26 MW less heat is delivered from process heaters, compared to the present configuration, by this modification. But at the same time, no MP-steam is produced, see Figure 9. (The loss in steam production is the same amount as the decreased load on process heaters.) Whether this is beneficial depends on the refinery’s steam and fuel gas balance which we have not studied in total.

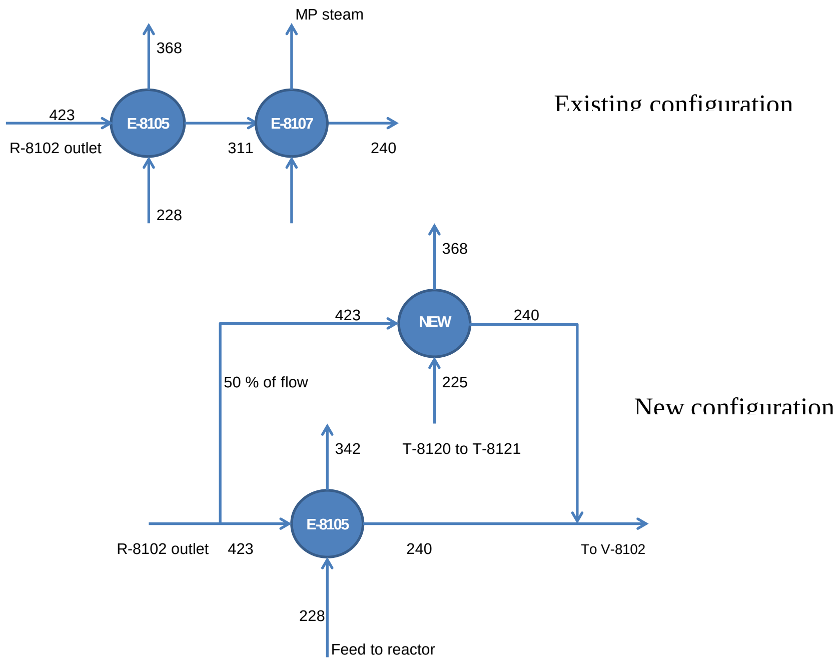


Figure 13 The stream from R-8102 is today used to heat the feed to the reactor. If the stream is split into two it can raise the temperature of the feed to T-8121 so that no heat must be added in H-8120. (The temperature in to T-8120 was 368°C at the time of data collection.)

4.3.3 Modification 3

To reach the minimum heat demand, that is construct a MER-network, the heat exchanger network will have to be rearranged in a more complicated (and expensive) way. Modification 3 is an example of how this can be accomplished and eliminates practically all existing pinch violations; cooling above the pinch, and heat down through the pinch.

Nine new heat exchangers are added, and many of the present heat exchangers must be used in new configurations and heat exchanger areas, pressure and temperature ratings, etc., must be checked.

This make-over is so radical that is not possible to show the changes based on an existing PFD like Figure 12 (configuration is available on request).

Compared to modification 2, the load on process heaters are decreased with 12 MW, and if this saving could motivate such a radical revamp of the heat recovery network has to be examined. No steam MP steam is generated in the unit. Again, this could be beneficial depending on the refinery's steam and fuel gas balance.

5 Detailed analysis of the Mild hydro Cracker unit MHC 810

The mild hydrocracker (MHC) functions both as a hydrotreater to the fluid-catalytic-cracker (FCC) by desulphurizing the feed and a cracker to increase the yield of naphtha and gasoil.

5.1 General process description

The process is rather similar to the isocracker unit, section 4.1, but with a lower hydrogen partial pressure. The absolute pressure in the reactors is about half the reactor pressures in ICR 810.

5.2 Specifics for MHC 240 at Preem LYR

The capacity of the unit is similar to the isocracker ICR 810, and at the time of the inventory the throughput was roughly 6400 ton/day.

The present heat demand for this unit is 36.7 MW according to the analysis performed in Part I. If we assume that there are no constraints in how heat exchanging can be performed within the unit, the potential savings amounts to about 70% or 25.3 MW (see Table 2).

As in all units of this type, there is high difference in operating pressure in the different process equipment. (However, the maximum operating pressure is considerably lower than in the ICR-unit.) The penalty for not allowing direct heat exchange between high and low pressure streams is about 20 MW, thus decreasing the theoretical energy saving potential to about 10% of the present consumption[4]. To achieve significantly savings we should thus not impose this restriction on the analysis.

A PFD for the existing layout is shown in Figure 14.

[4] It should be noted that the existing layout includes heat exchangers exchanging heat between the two pressure levels.

Figure 14 Process flow diagram of Mild hydrocracker, MHC 240

5.2.1 Existing heat exchangers

Equipment numbers are shown in Figure 14.

Firstly, the liquid feed to the reactors is heated from 45 to 315°C before hydrogen is added. Heat from the reactor outlet is used to heat the feed streams. The heat exchangers are E-2401 and E-2403. The hydrogen feed is heated from 88 to 397°C in E-2402 and E-2404. In the process heater H-2401 the temperature is raised to 400°C.

The reactor outlet stream is cooled by the feed streams from 409 to 165°C.

The separation of products starts by a liquid-vapour separation in the hot high pressure separator (HHPS), S-2402. Wash water is injected to the hydrogen rich vapour stream which is cooled in EA-2405 to 26°C. Sour water is drawn from the cold high pressure separator (CHPS) S-2401 while liquid products are mixed with liquid outlet from S-2402. The recycle hydrogen stream from S-2401 is treated in an amine absorber.

Lighter products are stripped off in three strippers, T-2407 and, at reduced pressure, in T-2408 and T-2409. The liquid from S-2402 is reheated from 189 to 225° C by using heat from products going to tank and also by a pump-around in T-2408 before entering T-2407.

Between T-2407 and T-2408 the temperature is raised from 220 to 323°C by using heat from the bottom products of T-2408 and process heater H-2403.

A side flow from T-2408 is fed to T-2409 where more light products are stripped off. The bottom product from T-2409 (GO product) is cooled in E-2415 (heating the feed to T-2407). However, the major part of the heat is cooled off in an air heat exchanger, EA-2417.

The bottom products from T-2408 (VGO product) not only recovers heat to the process as described above, but also generates some low pressure (LP) steam in E-2418 before air-coolers EA-2422 and EA-2412 cools down the product to tank temperature.

5.2.2 Stream data

Stream data for all streams used in the analysis are listed in Table 6. Heat in the flue gases leaving the stack of H-2401 is included in stream data since one of the modifications will include use of this heat. However, it should be noted that today this heat is partly used for preheating combustion air in a common stack/air-preheater for H-2403, H-2801, H-2803, and H-2401.

Steam used as stripper steam is not included as a heat demand but this consumption must be included in the overall steam balance of the refinery.

Table 6 Stream data for all streams that are heated or cooled in the MHC 240 unit.

Stream description		Tin [°C]	Tut [°C]	Duty [MW]
Feed to R-2401	Cold	45	400	70.3
Hydrogen feed to R-2401	Cold	88	400	15.7
Reactor outlet to S-2402	Hot	409	165	66.3
Hydrogen recycle, S-2402 to S-2401	Hot	91	26	11.0
S-2401/2 to T-2407	Cold	189	225	5.6
T-2407 BTM to T-2408	Cold	220	323	20.0
T-2407 OH to V-2413	Hot	132	36	4.5
T-2408 BTM to tank	Hot	315	59	34.2
T-2409 GO to tank	Hot	193	26	3.6
T-2408 OH	Hot	123	27	9.6
T-2408 reflux	Hot	220	192	3.0
H-2401 Flue gas, after heater	Hot	476	125	6.1

5.2.3 Energy savings potential using advances curves

The analysis using actual load curves (see section 4.2.3), Figure 15, shows that the remaining heat in flue gases of H-2401 can cover some of heat demand currently supplied with utility, that is fired process heaters. Approximately 3 MW can be recovered. If the flue gases are not considered, no heat can be covered by direct replacement of utility heaters and coolers by process-to-process heat exchangers.

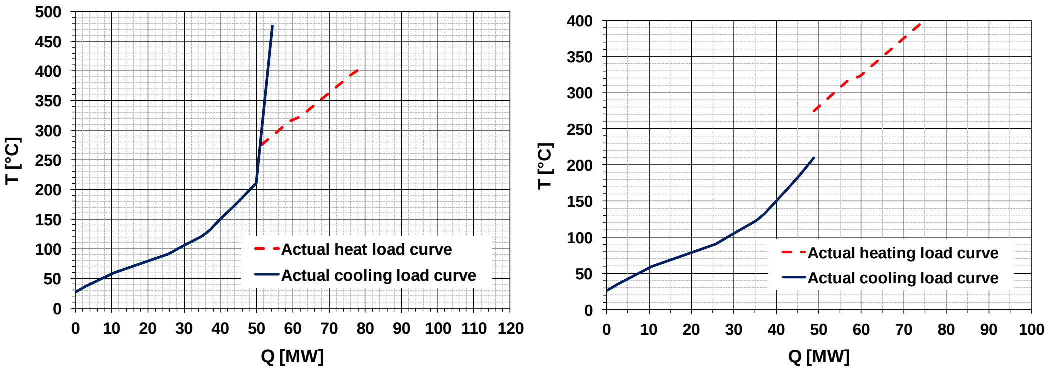


Figure 15 Actual heat load curve and actual cooling load curve with H-2401 flue gases (left) and without the said flue gases (right).

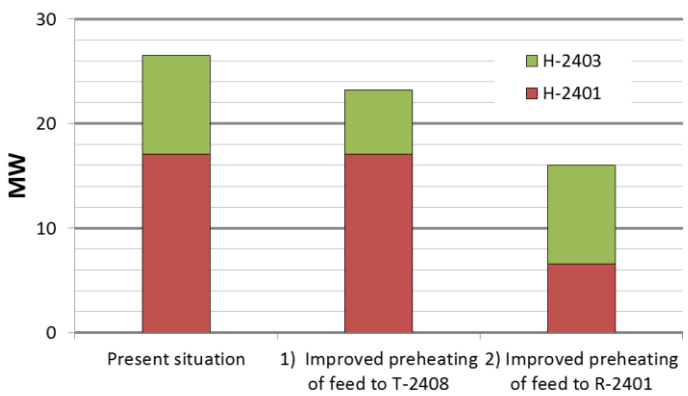
5.3 Possible modifications to MHC 240

Here we will discuss two possible modifications to the mild hydro cracker MHC 240 based on the process integration study.

The first modification is based on the actual load curves shown in Figure 15. The exit temperature for flue gases from process heater H-2401 was at the time of the inventory 476°C. This is represented in the actual cooling load curve as a fictional cooler from 476 to 125°C, the assumed lowest acceptable stack temperature.

The second modification use heat removed in condensers of T-2407, T-2408 and other product coolers, to heat the incoming feed to the unit.

Generation of LP-steam is not affected and in contrast to ICR, this generation is not a cooling above the pinch since the pinch is the pinch temperature is 230°C.



Modification	New heat exchangers	Heat supplied by H-2401 [MW]	Heat supplied by H-2403 [MW]	Savings [MW]
Present situation	–	17.1	9.4	
1) Improved preheating of feed to T-2408	1	17.1	6.1	3.3
2) Improved preheating of feed to R-2401	5	6.6	9.4	10.5

Figure 16 Two different process integration modifications. The modifications cannot be combined.

5.3.1 Modification 1

In this first modification, Figure 17, the hot flue gases from H-2401 is used to heat the feed to T-2408 and thus reduce the fuel gas demand in H-2403. This modification will reduce the load on the charge heater by 3 MW and thus the fuel gas demand by 4 MW.

Figure 17 Process flow diagram of MHC 240 with the changes in modification 1 included.

5.3.2 Modification 2

Modification 2 shown in Figure 18 includes use of heat removed in air coolers to heat the liquid feed and gas feed to the unit. Three new heat exchangers are used to collect this heat, NEW1-3. E-2408, used to cool pump-around of T-2408, can be exchanged with the gas feed in NEW4 to raise the temperature.

Since the feed streams are preheated before they are heat exchanged with R-2402 outlet, they can reach a higher temperature before entering the process heater H-2401.

A new heat exchanger, NEW5 will make it possible to raise the temperature of the liquid feed and further reduce the fuel demand in H-2401.

Modification 1 and modification 2 cannot be combined since modification 2 reduces the heat demand in H-2401 and thus will reduce the heat available to heat the feed to T-2408.

Figure 18 Process flow diagram of MHC 240 including modification 2.

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