



CHALMERS
UNIVERSITY OF TECHNOLOGY



Temperature Evaluation of Furnace for Crude Oil Refining

Master's thesis in Sustainable energy systems

JANNE SCHILLING

DEPARTMENT OF SPACE, EARTH, AND ENVIRONMENT

CHALMERS UNIVERSITY OF TECHNOLOGY
Gothenburg, Sweden 2026
www.chalmers.se

MASTER'S THESIS 2026

Temperature Evaluation of Furnace for Crude Oil Refining

JANNE SCHILLING



CHALMERS
UNIVERSITY OF TECHNOLOGY

Department of Environmental and Energy Sciences
Division of Energy Technology
CHALMERS UNIVERSITY OF TECHNOLOGY
Gothenburg, Sweden 2026

Temperature Evaluation of Furnace for Crude Oil Refining
JANNE SCHILLING

© JANNE SCHILLING, 2026.

Supervisor: Judit Fortet Casabella, Department of Environmental and Energy Sciences

Supervisor: Tharun Roshan Kumar, Department of Environmental and Energy Sciences

Supervisor: David Börjesson, St1 Refinery

Examiner: Martin Seemann, Department of Environmental and Energy Sciences

Master's Thesis 2026

Department of Environmental and Energy Sciences

Division of Energy Technology

Chalmers University of Technology

SE-412 96 Gothenburg

Telephone +46 31 772 1000

Typeset in L^AT_EX
Gothenburg, Sweden 2026

Abstract

Crude oil distillation is an energy intensive process and the bulk of the energy is provided in the preheating system for the crude oil. A furnace is the final unit operation in the preheating train where the crude feed temperature is raised to undergo atmospheric distillation. The furnace outlet temperature is a significant parameter to achieve proper separation of components in the crude oil. This master thesis explores the possibility to decrease the furnace outlet temperature while maintaining separation of crude components in the distillation column. A decrease of furnace outlet temperature is expected to cause a decrease in fuel consumed in the furnace, which in extension is expected to cause a decrease in CO₂-emissions. To estimate the effect of decreased furnace outlet temperature, the process was modeled in Aspen HYSYS and the model was validated with operational data from a refinery. Experimental test runs were performed at a refinery where furnace outlet temperature was gradually decreased. The result was a 15% decrease in fuel demand in the furnace and related emission of CO₂, and an increase in stripping steam demand in the distillation. The separation of the crude oil components was not impacted. When the impact of change in stripping steam demand was converted to equivalents of fuel gas, the combined impact on fuel demand for the furnace and for steam production was 15% reduction for a temperature decrease of 10°C compared to normal operating conditions. The economic potential for such a temperature decrease is 23 MSEK/year in fuel cost and EU ETS cost. The reduction in fuel gas demand for the investigated CDU corresponded to 2.2% of the total fuel gas consumed at the refinery.

Keywords: crude distillation unit (CDU), crude oil furnace, furnace outlet temperature.

List of Acronyms

Below is the list of acronyms that have been used throughout this thesis listed in alphabetical order:

APC	Advanced Process Control
ASTM	American Society for Testing and Materials
bf.	before
cbm	Cubic meter
CDU	Crude Distillation Unit
EU ETS	European Union Emissions Trading System
FBP	Final Boiling Point
HP	High Pressure
HX	Heat Exchanger
IBP	Initial Boiling Point
LPG	Liquefied Petroleum Gas
LP	Low Pressure
MK1	MiljöKlass 1
MP	Medium Pressure
P&ID	Process and Instrumentation Diagrams
TBP	True Boiling Point

Nomenclature

Below is the nomenclature of indices, parameters, and equipment terms that have been used throughout this thesis.

Indices

i	Index for component i
n	Index for equilibrium stage number n

Parameters

K	UOP K factor
L	Liquid flow
V	Vapor flow
x	Liquid molar fraction of most volatile component
y	Vapor molar fraction of most volatile component
h	Enthalpy
Q_{mixing}	Heat of mixing
$Q_{external}$	Heat supplied from external source
Q_{loss}	Heat loss
K_i	Equilibrium coefficient
ΔG	Gibbs free energy
ΔH	Change in enthalpy
ΔS	Change in entropy

Equipment Terms

F101	Furnace
F101A, F101B	Sections of furnace F101
V102	Desalter
V103	Pre-flash
C101	Atmospheric distillation column
C102	Side stripper for kerosene
C103	Side stripper for light gasoil
C105	Side stripper for heavy gasoil
P101	Pump after storage tanks
P102	Pump after pre-flash
100-unit	Crude distillation unit
CDU 3	Crude distillation unit

Contents

List of Acronyms	viii
Nomenclature	xi
List of Figures	xv
List of Tables	xvii
1 Introduction	1
1.1 Background	1
1.2 Crude Oil Distillation	2
1.3 Aim	2
1.4 Scope	2
2 Theory	3
2.1 Crude Oil Characterization	3
2.2 Modeling of CDU	3
2.3 Crude Oil Fouling	4
3 Description of Proposed System	7
3.1 Crude Oil Blending	7
3.2 Desalter	8
3.3 Pre-flash	9
3.4 Furnace	9
3.5 Atmospheric Distillation	11
3.6 Side Strippers	12
3.7 Steam Generation System	12
3.8 Product Sampling & Analysis	12
4 Method	15
4.1 Process Simulation	15
4.1.1 Data Collection	17
4.1.2 Model Validation using Base Case	17
4.1.3 Crude Feed Composition	18
4.1.4 Pre-flash	19
4.1.5 Furnace	20
4.1.6 Atmospheric Distillation Column	20

4.2	Test Runs	22
4.3	Performance Indicators	23
4.3.1	Product Separation	23
4.3.2	Model Accuracy	23
4.3.3	Fuel Consumption in Furnace	24
4.3.4	Steam Consumption	24
4.3.5	CO ₂ -emissions	25
4.3.6	Economic Evaluation	25
5	Results & Discussion	27
5.1	Model Validation	27
5.2	Source of Error during Test Run 1	28
5.3	Product Separation	29
5.4	Fuel Consumption & Steam Consumption	31
5.5	Emissions	32
5.6	Economic Evaluation	33
5.7	Outlook	34
6	Conclusion	37
	References	39
A	Appendix 1	I
A1	Results from Test Run 1	I
	Product Analysis	I

List of Figures

1.1	Crude distillation unit at St1 Refinery	2
2.1	Stage in equilibrium	3
3.1	Overview of investigated CDU system	7
3.2	Crude oil blending	7
3.3	Desalter	8
3.4	Pre-flash	9
3.5	Furnace in process overview	10
3.6	Furnace layout splitting crude feed into four loops divided into two sections	10
3.7	Atmospheric distillation column in process overview and tray numbering	11
4.1	Process parameter variation around base case	18
4.2	Crude oil blending in Aspen Hysys	18
4.3	Pre-flash in Aspen Hysys using TBP cut points to steer partition of vapor and liquid.	19
4.4	Furnace set up in Aspen Hysys	20
4.5	First version of distillation column in Hysys with only top and bottom products	21
4.6	Second version of distillation column in Hysys with side draws added	22
4.7	Final version of distillation column in Hysys with side strippers added	22
4.8	Set up of test runs	23
5.1	Comparison of product flow rates between the measured data and simulated data for base case	27
5.2	Comparison of product flow rates between the measured data and simulated data for T2 and T3 from test run 2	28
5.3	Setpoint vs flow rate in loop C, F101B	29
5.4	Setpoint vs flow rate in loop A, F101B	29
5.5	Boiling point range for products during test run 2	30
5.6	Cloud point of heavy gasoil test run 2	31
5.7	Specific fuel gas consumption test run 2	31
5.8	Specific steam consumption test run 2	32
5.9	Specific CO ₂ -emissions test run 2	33
5.10	Yearly cost for fuel to furnace & emissions	34
A1	Product quality analysis test run nr 1	I

List of Figures

A2	Specific fuel gas consumption	I
A3	Specific steam consumption	II

List of Tables

4.1	Simulation specifications for base case	16
4.2	Parameters for crude blend used in base case	19
4.3	Specified and calculated parameters for the column in the different column versions	21
4.4	Furnace outlet temperature during test runs	23
4.5	Parameters for CO ₂ -emissions calculation	25
4.6	Parameters for variable operating cost calculation	25
5.1	Error analysis of model in base case	27
5.2	Reduction of fuel gas during test run 2 compared to base case	32
5.3	Sensitivity analysis of changes to fuel cost and CO ₂ allowance cost . .	34

1

Introduction

1.1 Background

St1 Refinery AB owns and operates a refinery in the harbour of Gothenburg since December 2010. The refinery was previously owned by AB Svenska Shell, who acquired the refinery from Koppartrans AB in 1964 [1]. Koppartrans AB originally purchased the refinery from a partly emptied military depot south of Los Angeles just after the end of the second world war. At the time, the refinery consisted of two crude oil distillation units (CDU) and two units for thermal reforming of gasoline. A third crude oil distillation unit was added in the 1970's [1], pictured in Figure 1.1.

Manufacturer's instructions for the CDUs are outdated in some of the aspects, because the composition of the crude oil feedstock has changed from fifty years ago [2]. Today, the furnaces upstream of the CDUs are more or less operated in the same way regardless of the crude feed composition. Distillation feed stream temperature is a result of the operating temperature in the furnace, marked with a red circle in Figure 1.1. Without specific insight into the limitations of the feed stream temperature, there is a risk that furnaces are operated at too high a temperature. Consequences of excess temperature in the furnace include thermal cracking of crude oil and coke deposition, high consumption of fuel gas and excess emissions of CO₂ and NO_x emissions [2].

The refining of crude oil is one of the most common processes in the world that make use of distillation technology. It is also one of the most energy intensive separation processes at refineries, affecting the whole plant's energy consumption by up to 24% [3]. This makes it a prioritized process for optimization.

The European Union launched an Emissions Trading System (EU ETS) for carbon in 2005 where a limit for the total amount of greenhouse gas (GHG) emissions is set and allowances to emit are auctioned and traded by polluters [4]. Since then, the system has stepwise expanded its reach to more polluters and tightened the limit of emission allowances available.

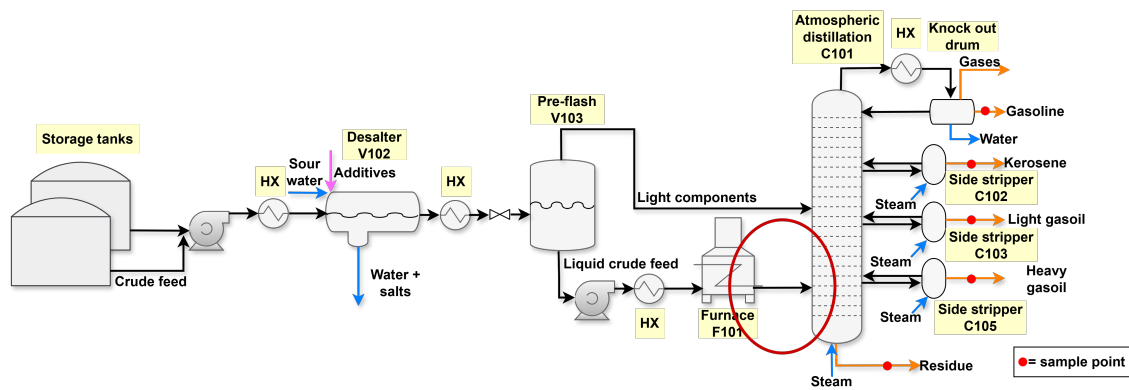


Figure 1.1: Crude distillation unit at St1 Refinery

1.2 Crude Oil Distillation

Distillation is a separation process in which components of a mixture are separated on the basis of a difference in relative volatility. In a distillation column, vapor flows upward while liquid flows downward. Proper mixing between the vapor phase and liquid phase over trays in the column is required to allow transfer of mass and energy. Lighter components accumulate at the top of the column while heavier components accumulate at the bottom.

1.3 Aim

The aim of this master thesis is to determine the effect that decreased furnace outlet temperature has on product separation, fuel gas consumption and CO₂-emissions. To achieve that, a model of the process will be created in Aspen Hysys and will be validated with operational data from the refinery. The aim of the model is to test the lower limit of furnace outlet temperature and to evaluate the impact on the product yield distribution.

1.4 Scope

This master's thesis uses the the newest CDU at the Gothenburg refinery from the 1970's - locally called CDU 3 or the 100-unit, as a case study. The Aspen Hysys model primarily includes the crude blending, the pre-flash, the furnace and the atmospheric distillation tower. The desalination equipment and the downstream vacuum distillation units are not included in the simulation. The simulation is performed in steady state. The thesis also includes test runs at the CDU 3 where operating conditions are monitored and samples of the distillation products are taken and analyzed.

2

Theory

2.1 Crude Oil Characterization

Two of the most important characteristics of crude oils are density and sulfur content. Density ranges from light to heavy. Sulfur content is characterized as sweet (low sulfur content) or sour (high sulfur content) [5]. Watson K factor, or UOP K factor, is an approximate index of paraffinicity. High values correspond to high degrees of saturation [6]. The UOP K factor, K , is defined according to Equation 2.1.

$$K = \frac{1.8Tb^{1/3}}{SG} \quad (2.1)$$

where Tb is the average boiling point (K) and SG is the specific gravity defined by Equation 2.2.

$$SG = \frac{\rho_{crude}}{\rho_w} \quad (2.2)$$

where ρ_{crude} is the density of the crude oil at 60°F and ρ_w is the density of water at 60°F. Cloud point is the temperature below which longer paraffinic molecules starts to precipitate as a wax out from the fuel [7].

2.2 Modeling of CDU

Commercial software such as Aspen Hysys are based on rigorous models of thermodynamic calculations, material balance, equilibrium, summation and heat balance (MESH-equations) [8].

Consider a theoretical ideal stage as shown in Figure 2.1 with incoming and outgoing vapor flow and liquid flow.

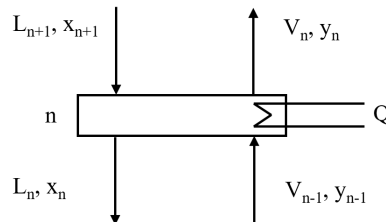


Figure 2.1: Stage in equilibrium

where L is the liquid flow rate, V is the vapor flow rate, x is the liquid molar fraction of the most volatile component, y is the vapor molar fraction of most volatile component, and n is the number of stage in the distillation column.

The material component balance over stage n for steady state simulation can be expressed as:

$$L_{n+1}x_{n+1} + V_{n-1}y_{n-1} - L_nx_n - V_ny_n = 0 \quad (2.3)$$

The heat balance or energy balance can be defined as:

$$L_{n+1}h_{n+1} + V_{n-1}h_{n-1} - L_nh_n - V_nh_n + Q_{mixing} - Q_{external} - Q_{loss} = 0 \quad (2.4)$$

where h is the enthalpy of the corresponding stream, Q_{mixing} is the heat of mixing, $Q_{external}$ is heat supplied from external source, and Q_{loss} represents heat losses.

For theoretical ideal stages, the outgoing streams from each stage are assumed to be in vapor-liquid equilibrium described by the following expression:

$$y_i = K_ix_i \quad (2.5)$$

where K_i is the equilibrium coefficient for component i . The equilibrium coefficient is defined from the chosen thermodynamic model. Typical models include Peng-Robinson equation of state, of Braun K10 for petroleum components [9].

The distillation column environment in Aspen Hysys solves the MESH equations when enough information is provided and the degrees of freedom for the system are zero. The number of constraints must equal the number of unknown parameters. Each parameter specification reduces the degrees of freedom by one [10].

2.3 Crude Oil Fouling

It is not uncommon for refineries to experience problems in operations of CDUs caused by crude oil fouling in the furnace upstream of the distillation column. Upon maintenance stops, operators may identify black deposits inside the tubes in the furnace, and attribute it to coke deposition. However, there are more processes which cause the deposition than coke, and understanding the mechanisms can aid in mitigation and cleaning strategies.

Crude oil is a complex mixture of many components, and when exposed to heat, the species in the mixture can react in different ways to cause deposition of material. The following three main reaction pathways have been identified as causes of crude oil fouling [11]:

1. Dissolved metal ions can react with compounds containing sulfur and lead to corrosion fouling due to iron sulfide.

2. A series of autoxidation reactions of hydrocarbon compounds in the presence of dissolved oxygen can lead to the formation of insoluble gums. This is a free-radical chain reaction mechanism. The process is exothermic and can occur at relatively low temperatures [12].
3. Thermal decomposition or cracking of hydrocarbon compounds can lead to formation of coke when temperatures are high enough. During thermal cracking, larger hydrocarbons are broken down into lighter alkanes or alkenes. While the reaction mechanism is endothermic (positive ΔH), the breakup of large compounds into smaller ones results in a large increase of entropy (ΔS). The resulting Gibbs free energy ΔG can thus be negative, which favors the process at high temperature above 650 K.

$$\Delta G = \Delta H - T\Delta S \tag{2.6}$$

The long interval in between planned maintenance stops, often up to two years, makes it difficult to assign the resulting crude oil fouling in the furnace to a specific crude oil, as the crude oil feedstock composition is varied much more frequently.

3

Description of Proposed System

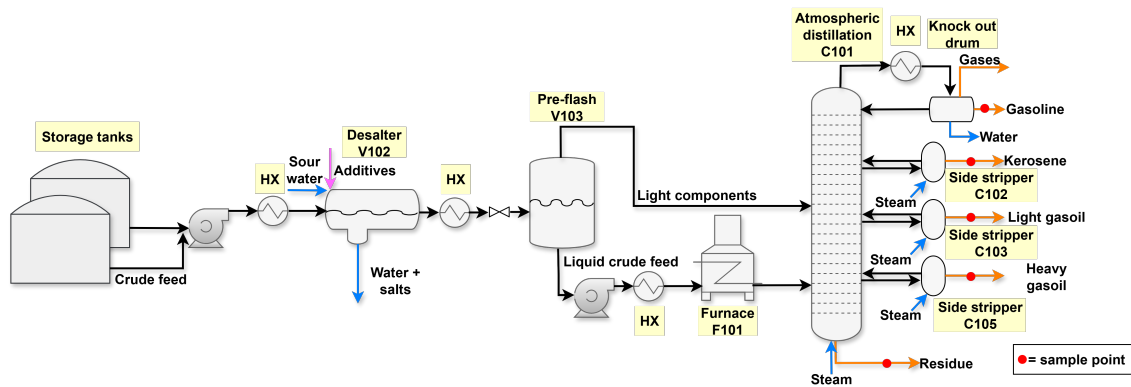


Figure 3.1: Overview of investigated CDU system

3.1 Crude Oil Blending

The crude oil blending is represented in Figure 3.2.

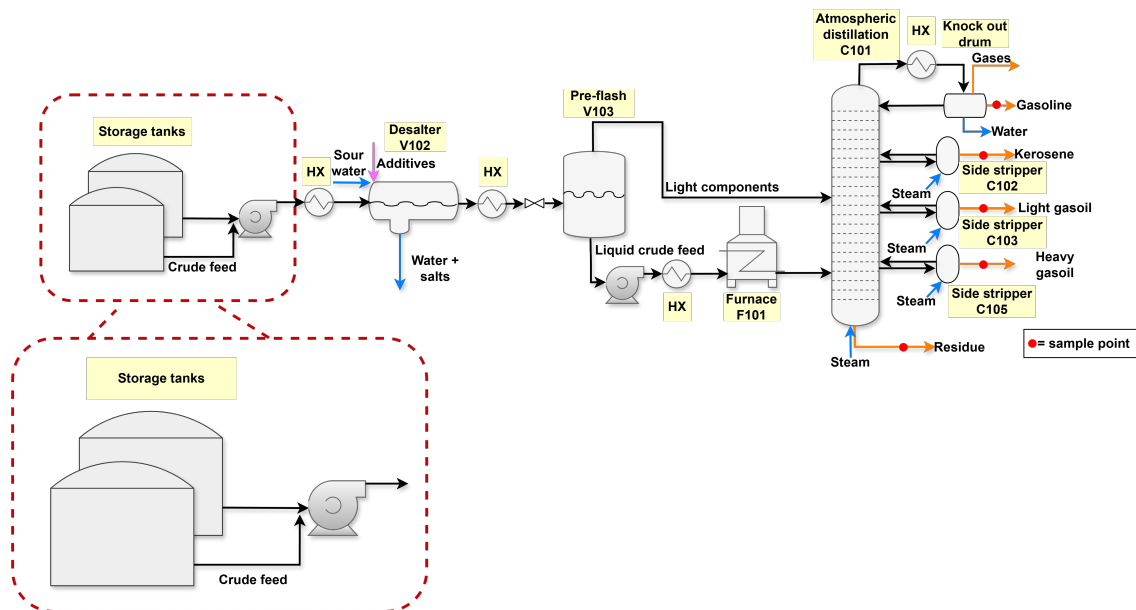


Figure 3.2: Crude oil blending

Multiple crude oils are typically blended to achieve a crude feed composition which matches the specifications of the refinery. The crude oils are stored in caverns and pumped into a feeding tank in the desired ratios. The blended crude feed is then pumped through a series of heat exchangers (HX) where the temperature is increased, on to the desalter.

3.2 Desalter

The desalter is shown in Figure 3.3.

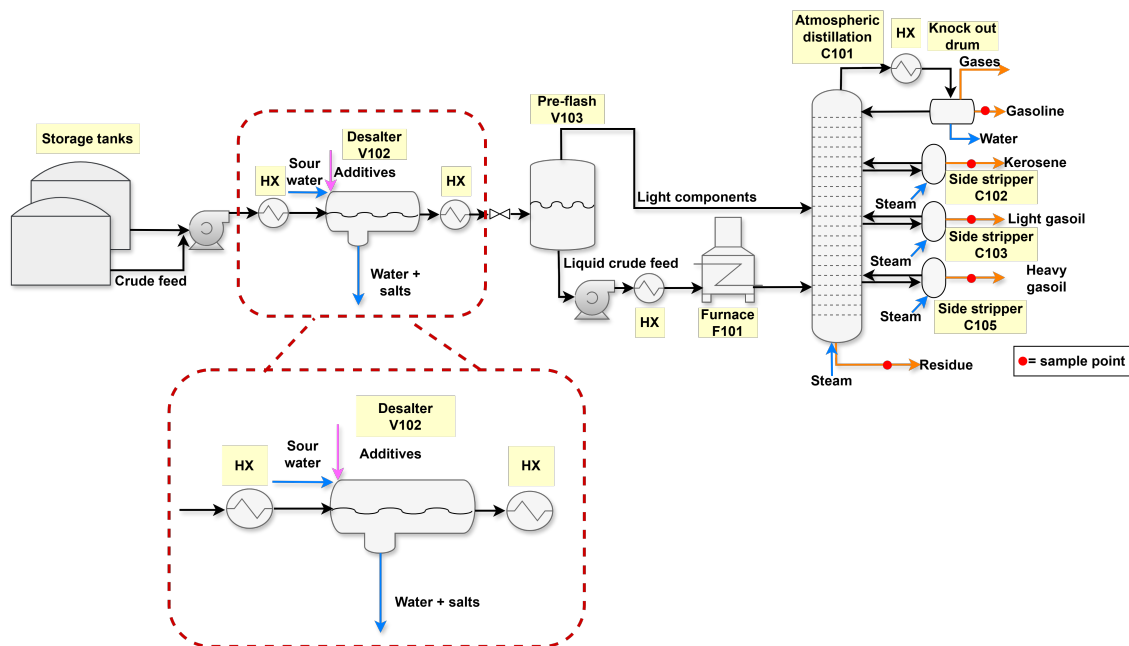


Figure 3.3: Desalter

The purpose of the desalter is to remove dissolved salts that are present in dispersed water droplets in the oil. The ingoing crude feed is mixed with process water in a turbulent manner. An emulsion-breaking additive is added to then allow the water phase to separate from the oil phase. Additionally, iron grates installed inside the vessel are used to apply an electric field, which further drives the phase separation [13]. The temperature inside the desalter is an important parameter for the efficiency. Too low temperature results in poorer phase separation between water and oil, while too high temperature results in higher solubility of water in oil which also gives lower efficiency of the desalter [14]. The water that remains in the crude feed after the desalter will be separated out in the top vessel of the distillation tower. The crude feed continues from the desalter through a series of heat exchangers where the temperature is raised before entering the pre-flash.

3.3 Pre-flash

The pre-flash is shown in Figure 3.4.

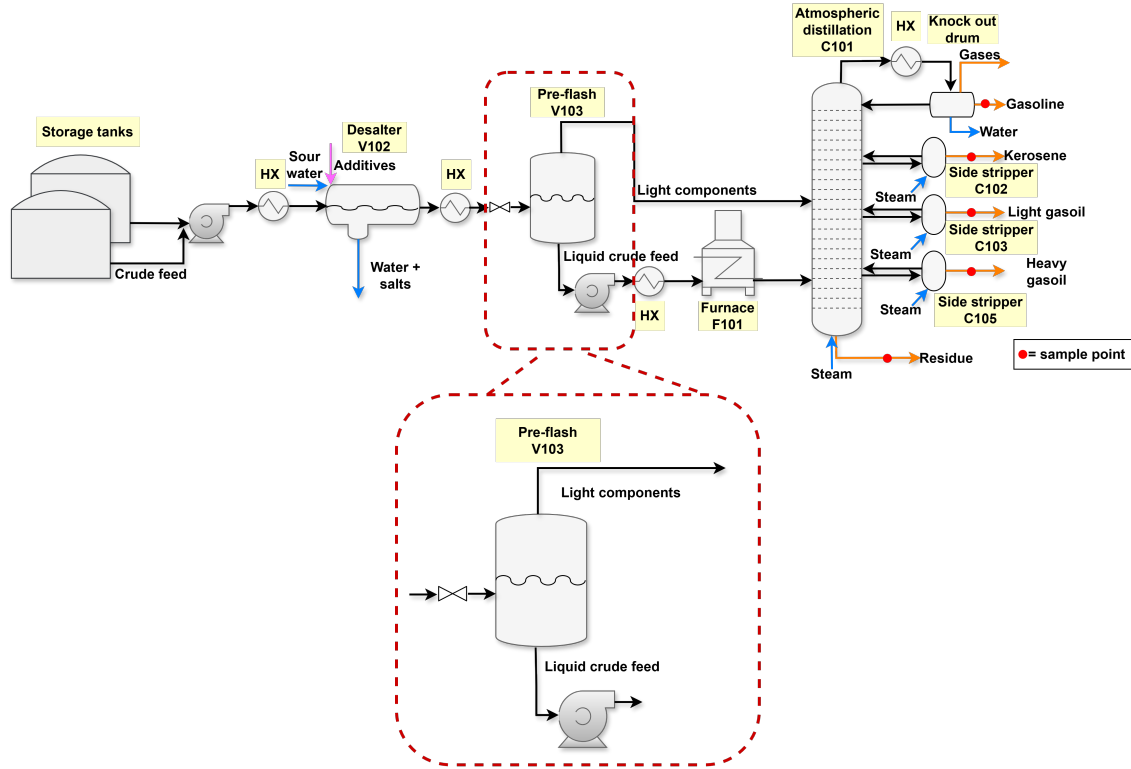


Figure 3.4: Pre-flash

The purpose of the pre-flash vessel is to flash off part of the crude feed. The vessel operates at lower pressure than the preceding units, which causes the lighter components of the crude feed to evaporate. Approximately 5-10% of the crude feed should evaporate here [14]. The vapor phase from the pre-flash vessel is fed directly to the distillation tower to tray 11. This is done to reduce the heat duty in the furnace and also to improve the hydraulic performance of the heat exchanger network [15]. The main bulk of the crude feed which is not evaporated in the pre-flash is fed through a series of heat exchanger where the temperature is raised before entering the furnace.

3.4 Furnace

Figure 3.5 shows the furnace in the process overview.

3. Description of Proposed System

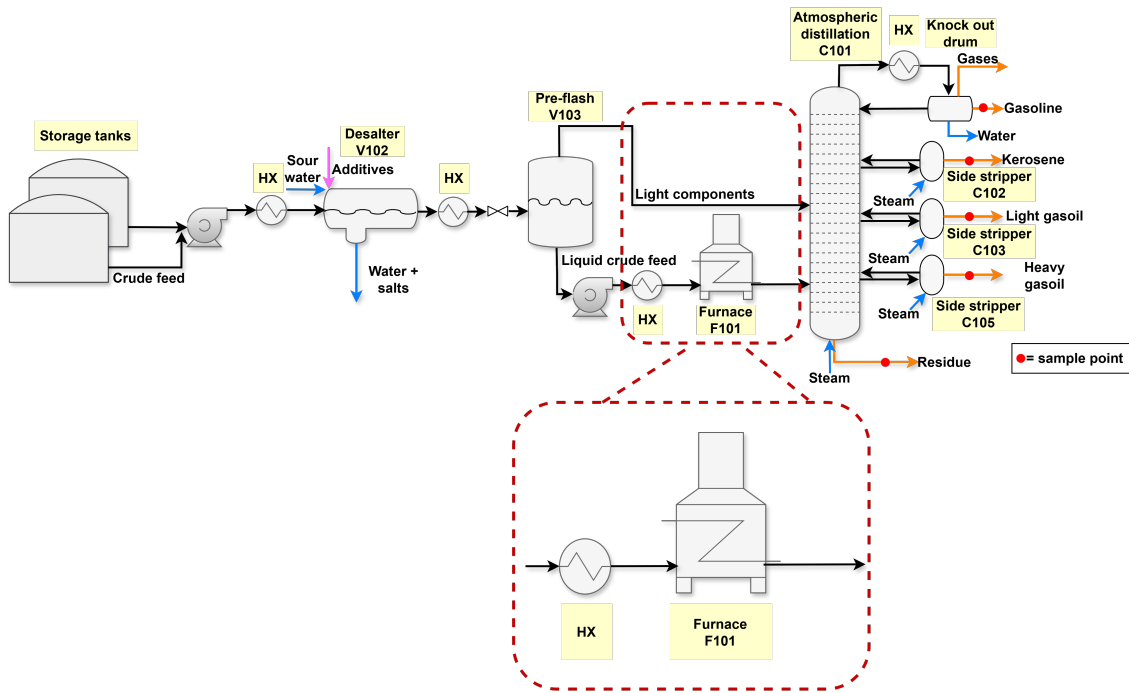


Figure 3.5: Furnace in process overview

The furnace, F101, consists of two radiant sections, F101A and F101B. The crude oil is split into four loops with two loops going through each furnace section, as depicted in Figure 3.6. An Advanced Process Control (APC) system is managing the crude flow rate through the loops with control valves. The APC system aims for the two loops exiting the same furnace section to have the same outlet temperature [14]. The fuel for the furnace is refinery fuel gas, which is produced in multiple units in the refinery and collected in a joint fuel gas system for the whole refinery.

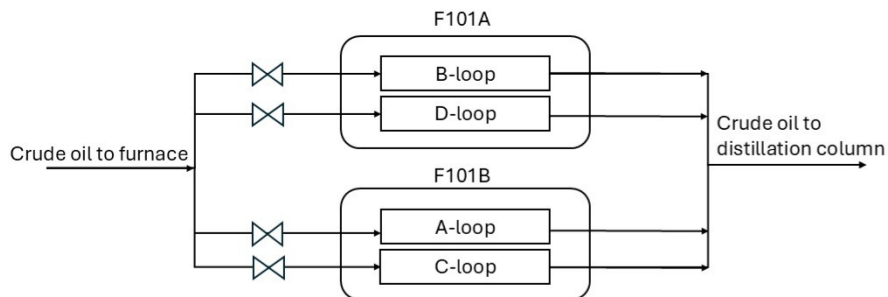


Figure 3.6: Furnace layout splitting crude feed into four loops divided into two sections

3.5 Atmospheric Distillation

Figure 3.7 shows the atmospheric distillation column in the process overview as well as the set up of the trays used in this case study.

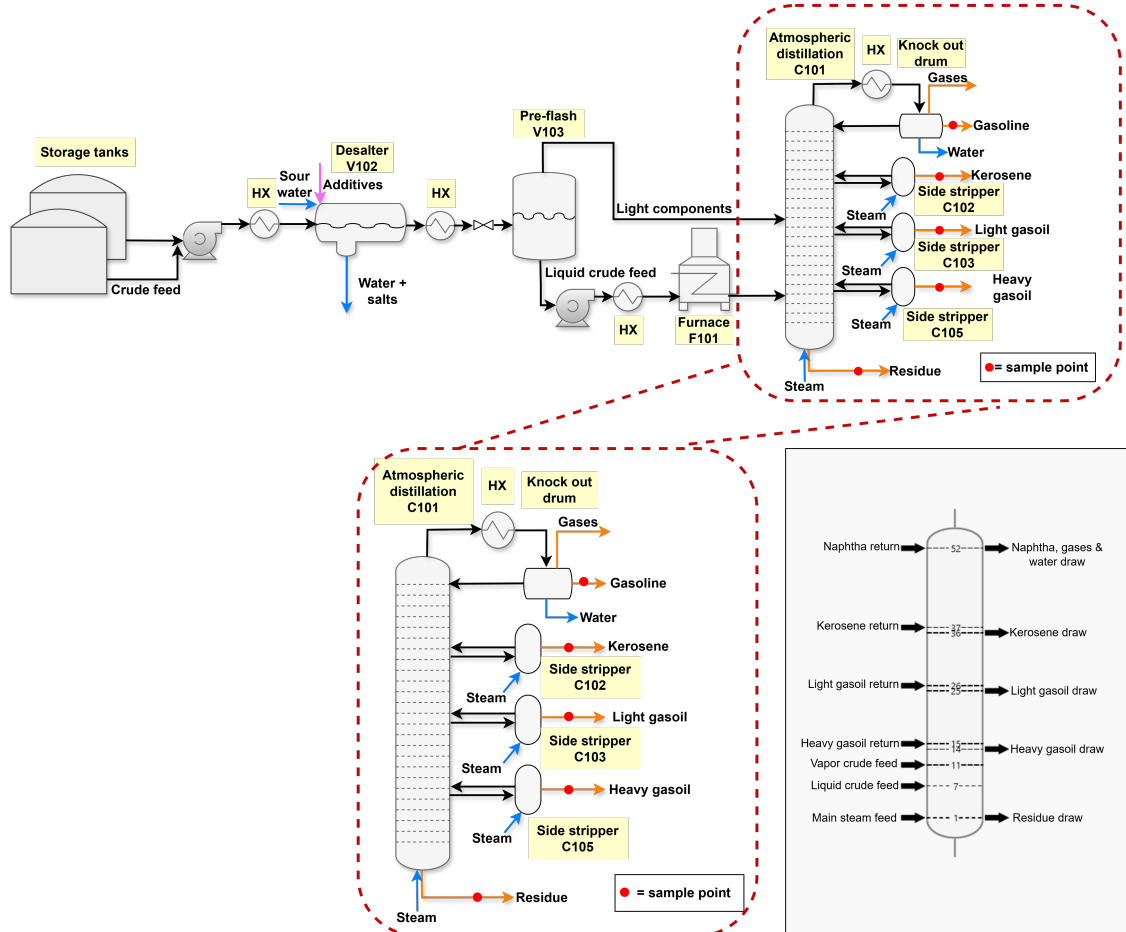


Figure 3.7: Atmospheric distillation column in process overview and tray numbering

The light components of the crude feed which were evaporated in the pre-flash enter the distillation column at tray 11 [14]. The liquid crude feed from the pre-flash has undergone partial evaporation in the furnace, and enters the distillation column as a mixture of liquid and vapor at tray 7. The components in the mixture which are in vapor state at the main feed tray temperature and pressure rise upward in the column, while the liquid residue passes down in the stripping section of the column. It is met by stripping steam that is added at the bottom of the column. The stripped residue is taken out at the bottom of the column. Heavy gasoil is drawn at tray 14 and passed through the heavy gasoil side stripper, from where a part of the stream

in returned to the main column at tray 15. Light gasoil is drawn from the tower at tray 25 and passed on to the light gasoil side stripper, from which part of the stream is returned to the main column at tray 26. Kerosene is drawn at tray 36 and partly returned to the column at tray 37 after passing the kerosene side stripper. The column top vapors are condensed before entering the overhead accumulator. From there, gasoline is partly returned to the top of the distillation column as cold reflux. The uncondensable gas stream (C1-C4 hydrocarbons) from the overhead accumulator is passed on for further processing in a gas treatment plant. Water originating from the stripping steam added to the distillation column is also drawn from the overhead accumulator and passed on to a water treatment system [13].

3.6 Side Strippers

For each side stream there is a side stripper, i.e for kerosene, light gasoil and heavy gasoil. Stripping steam is added to the side strippers to take off some of the lighter fractions in each stream and which is returned to the main column. The result is more precise separation of the product fractions. A lack of stripping steam can lead to too low flash-points in the products, while too much stripping steam increases the load at the top of the distillation tower and downstream equipment [14].

3.7 Steam Generation System

Low pressure (LP) steam och medium pressure (MP) steam can be used as stripping steam. Both LP steam and MP steam is produced in waste heat recovery steam boilers at the refinery. Normal production of LP steam is 48 tons/day and normal production of MP steam is 190 tons/day. If needed, high pressure (HP) steam can be let down to MP or LP steam. HP steam is produced in a gas fired steam boiler, or in a heat recovery steam generator. Normal production of HP steam is 4,500 tons/day [14].

3.8 Product Sampling & Analysis

Samples are taken daily of the gasoline, kerosene, light gasoil and heavy gasoil. Residue can also be sampled but is not part of the daily sampling and testing routine. The sampling point for gasoline is located after the overhead accumulator. Kerosene, lights gasoil and heavy gasoil sampling points are located after each respective side stripper. The residue stream sampling point is located after the bottom draw. For the gasoline fraction, the FBP should be below 180°C [16]. The kerosene fraction may be used either for jet production or production of Swedish diesel, "MK1". For MK1, the IBP should at least 180°C. However, the MK1 is blended of more components than just the kerosene fraction, so the kerosene fraction itself does not necessarily need to have an IBP above 180°C. 165°C IBP for the kerosene is acceptable. For jet, there is no limit for IBP. For the light gasoil fraction, the 95%

evaporated boiling point is the main parameter of interest, and the target is 290°C. For the heavy gasoil fraction, the cloud point is the main parameter of interest, and the target is below 3°C.

4

Method

4.1 Process Simulation

The parameters in Hysys were divided into two groups - fixed parameters and free parameters. The **fixed parameters** included the input parameters, organized in an "Input spreadsheet" directly connected to the model. These input parameters included flow rates of the crude oils, water to the desalter and stripping steam to the main column and side strippers, feed stream temperature, temperature before the pre-flash vessel, temperature of the furnace feed stream, temperature of furnace outlet stream, condenser temperature, and pressure at the bottom stage of the distillation column and the condenser pressure. Product quality indicators were also input parameters. Input parameters were changed by the user when simulating a new operating case. Model parameters such as equipment efficiency and component specifications were kept constant during different simulations. The fixed parameters are gathered in Table 4.1 with input values for the base case, and model parameters are denoted with *. The **free parameters** were the unknown values calculated by the model like fuel consumption and product flow rates.

Table 4.1: Simulation specifications for base case

Process Unit	Parameter	Value	Reference
Crude oil mixer	Crude feed temperature	16.6°C	[17]
	Crude feed pressure	24 bar	[17]
	Crude feed composition	See Table 4.2	Aspen Hysys
	Total feed rate	5,947 ton/d	[17]
HX series bf. desalter	Pressure drop*	0 bar	
	Temperature increase	112.1 °C	[17]
Pre-flash V103	Outlet pressure vapor*	2.45 bar	[14]
	Outlet pressure liquid*	2.49 bar	[14]
	Outlet temperature vapor	132 °C	[17]
	Initial cut point, vapor	-162 °C	Iteration process
	Initial cut point, liquid	48 °C	Iteration process
	Portion flashed off	3.34wt.%	[17]
HX series bf. furnace	Pressure drop*	0 bar	
	Temperature increase	138.4°C	[17]
Furnace F101	Air temperature*	25 °C	
	Air pressure*	1.013 bar	
	Fuel gas composition*	80wt.% CH ₄	
		20wt.% C ₂ H ₄	
	Fuel gas temperature*	25°C	
	Fuel gas pressure*	1.013 bar	
	Furnace outlet temperature	353°C	[17]
	Efficiency*	63.5%	
	Excess air percent*	10%	
Distillation column C101	Condenser pressure	1.3 bar	[17]
	Bottom stage pressure	2.45 bar	[17]
	Condenser temperature	37.92°C	[17]
	Gasoline 95% evaporated	150.7°C	[17]
	Kerosene 95% evaporated	229.9°C	[17]
	LG 95% evaporated	290.7°C	[17]
	HG cloud point	0.6°C	[17]
	Tray efficiency trays 1-14*	60%	
	Tray efficiency trays 15-52*	100%	
	Stripping steam flow rate	19.74 ton/d	[17]
Kero Side stripper C102	Nr of stages*	3	
	Stripping steam flow rate	11.40 ton/d	[17]

Process Unit	Parameter	Value	Reference
LG side	Nr of stages*	3	[17]
stripper C103	Stripping steam flow rate	8.49 ton/d	
HG side	Nr of stages*	3	[17]
stripper C105	Stripping steam flow rate	5.53 ton/d	

4.1.1 Data Collection

To develop the model, a detailed understanding of the process was required. Information about the process setup was collected from process descriptions, equipment specifications, and process control manuals. Process and Instrumentation Diagrams (P&IDs) were analyzed to identify how unit operations are interconnected via process streams, which was used to define system boundaries and establish the mass balance. The P&IDs also provided information on instrument locations and availability. Instrument tags were used to retrieve operational data from the digital process monitoring system. In addition, construction drawings were reviewed to obtain detailed design information, such as vessel dimensions, for example the volume of the pre-flash vessel. Information on the connections and streams between equipment is shared in Chapter 3, along with detailed configuration setup of the distillation column. Typical operating temperature, pressure and flow rates are also shared in Table 4.1. Vessel sizing equipment is not shared. Properties of the individual crude oils used are not shared, however the properties of the final blend used in the base case is shared in Table 4.2.

4.1.2 Model Validation using Base Case

A real working condition from a time with stable operation and representative conditions of the CDU was used to validate the model. In Figure 4.1 below, the selected base case is marked with the white square. The time for the base case was April 7th at 4 am. The process parameters which were monitored for stability when choosing the base case were the furnace outlet temperature (purple trend line), total liquid crude feed rate (blue trend line), and total fuel gas consumption (orange trend line). In the graph, the different variable have individual scales on the y-axis. At the selected time for the base case, there was little variation in the parameters of interest, which was considered a requirement for using it as base case.

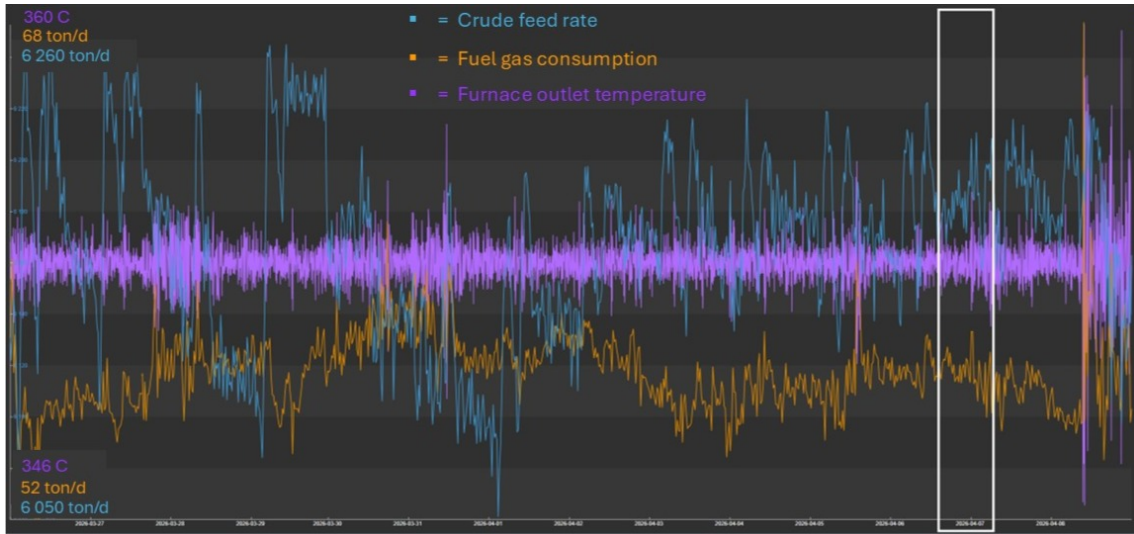


Figure 4.1: Process parameter variation around base case

4.1.3 Crude Feed Composition

In the refinery, a blend of multiple different crude oils is typically fed to the CDU. The individual crude oils were manually added to Hysys, and then blended in a mixer component as shown in Figure 4.2. To add the crude oils to Hysys, information from the crude supplier regarding density, UOP K-factor, viscosity, light ends composition and yield distribution was used as input data. A true boiling point (TBP) distillation curve was established by entering the TBP in steps of 5 vol-%. The Peng-Robinson Equation of State fluid package was selected as it is suitable for gas, oil and petrochemical applications [18].

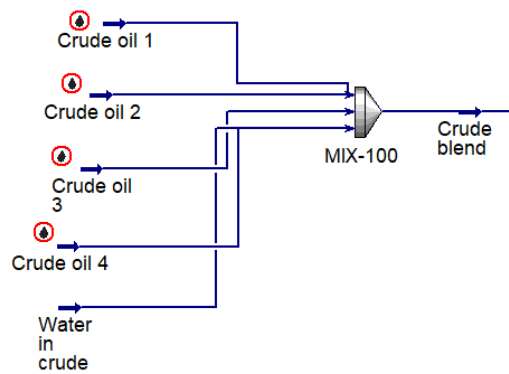


Figure 4.2: Crude oil blending in Aspen Hysys

The bulk properties, light components split and TBP range for the crude blend used in the base case is shown in Table 4.2.

Table 4.2: Parameters for crude blend used in base case

Parameter	Value	Unit
Density	833.9	(kg/m ³)
Watson K	11.66	-
Kinematic viscosity @20°C	5.083	cSt
Methane	0.01	wt. %
Ethane	0.01	wt. %
Propane	0.27	wt. %
i-Butane	0.21	wt. %
n-Butane	0.83	wt. %
i-Pentane	0.78	wt. %
n-Pentane	0.109	wt. %
TBP 0%	-28.28	°C
TBP 5%	50.78	°C
TBP 10%	85.97	°C
TBP 30%	174.5	°C
TBP 50%	269.6	°C
TBP 70%	364.6	°C
TBP 90%	496.6	°C
TBP 95%	578.1	°C
TBP 100%	707.3	°C

4.1.4 Pre-flash

The pre-flash vessel was modeled using a "Component splitter" unit in Hysys. From live process data, the portion of the incoming crude feed that was flashed off in the pre-flash could be determined. To mirror that portion, the TBP temperature was used as a cut point in Hysys. The initial cut point for the vapor stream was set just below the TBP for the lightest component in the crude blend, in this case that was at -162°C for methane. The initial TBP cut point for the liquid crude feed was tuned iteratively until the right portions were achieved - typically around 50°C to achieve 3-4% flashing. The set up of the pre-flash in Aspen Hysys is shown in Figure 4.3.

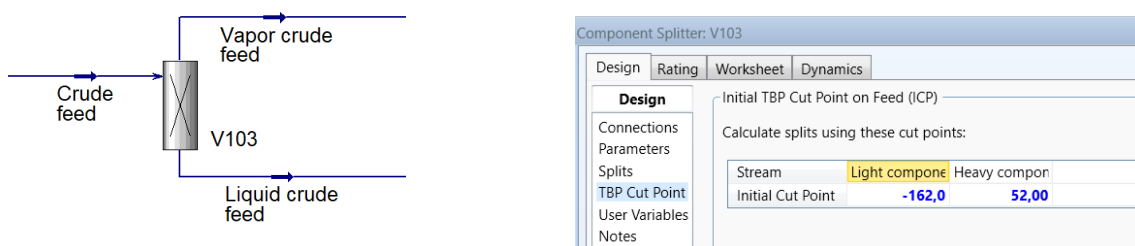
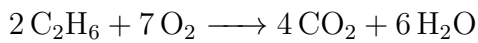
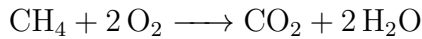


Figure 4.3: Pre-flash in Aspen Hysys using TBP cut points to steer partition of vapor and liquid.

4.1.5 Furnace

The furnace was modeled using a "Fired heater" unit in Hysys. Figure 4.4 shows the furnace with the crude oil inlet and outlet streams, a fuel inlet stream, and air feed stream and a fuel gas combustion product stream. The composition of the fuel stream was a mixture of 80 wt.% methane and 20 wt.% ethane. This was done so that the mixture would have a representative carbon content similar to the refinery fuel gas. The carbon content of the refinery fuel gas is measured by an instrument and available as a live tag. The combustion reactions of methane and ethane were specified in the properties environment.



Both combustion reactions were set as conversion reactions. The conversion of the reactants was determined by Equation 4.1:

$$\text{Conversion} = C0 + C1 \cdot T + C2 \cdot T^2 \quad (4.1)$$

For both combustion reactions, the parameters were set to $C0 = 100$, $C1 = 0$ and $C2 = 0$, so that the conversion was total and the reaction was independent of temperature. The furnace efficiency was set to 65% percent and the Excess air percent was set to 10%.

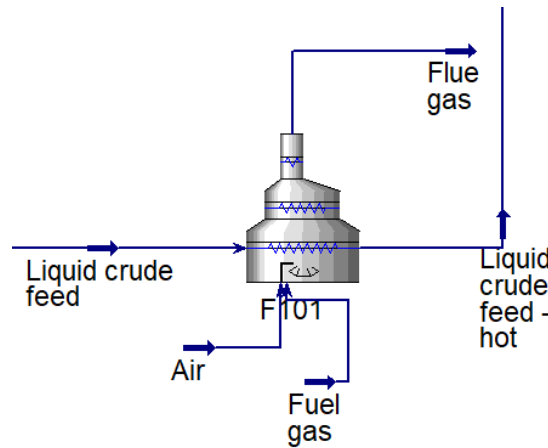


Figure 4.4: Furnace set up in Aspen Hysys

4.1.6 Atmospheric Distillation Column

The atmospheric distillation column model was developed in an iterative manner, starting with a very simplified set up and increasing column complexity by hand. This was done to ensure convergence of the model when increasing complexity. A "Refluxed absorber" unit was used in Aspen Hysys, which has a partial condenser but no reboiler. The first column model, pictured in Figure 4.5, featured only the liquid crude feed, main steam feed, and product streams at the top and bottom of

the column, i.e gasoline, water and gases withdrawn at the top and residue bottom product.

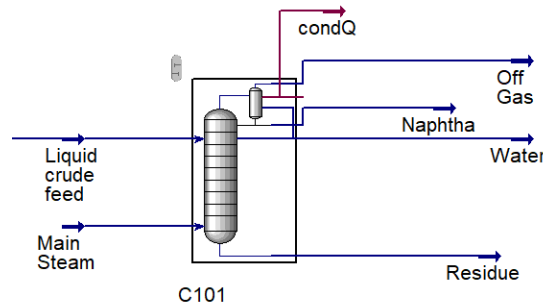


Figure 4.5: First version of distillation column in Hysys with only top and bottom products

To bring the degrees of freedom down to zero, the parameter of the product streams which was specified was the flow rates. Hysys then calculated the distillation curve of the products. The set up of specified and calculated parameters for the different versions of the distillation columns is summarized in Table 4.3.

Table 4.3: Specified and calculated parameters for the column in the different column versions

	Specified product parameter	Calculated product parameter
Column version 1 & 2	Gasoline flow rate	Gasoline TBP 95%
	Kerosene flow rate	Kerosene flow rate D1160 95%
	Light gasoil flow rate	Light gasoil flow rate D1160 95%
	Heavy gasoil flow rate	Heavy gasoil D1160 95%
Column version 3	Gasoline TBP 95%	Gasoline flow rate
	Kerosene D1160 95%	Kerosene flow rate
	Light gasoil D1160 95%	Light gasoil flow rate
	Heavy gasoil D1160 95%	Heavy gasoil flow rate

The second column model, pictured in Figure 4.6, included the remaining product fractions kerosene, light gasoil and heavy gasoil as side draws.

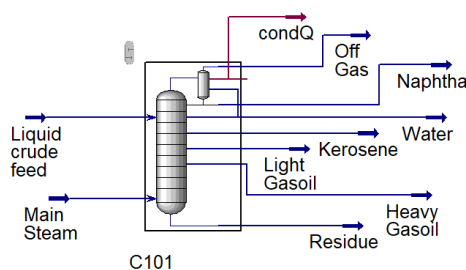


Figure 4.6: Second version of distillation column in Hysys with side draws added

The third and final column model, pictured in Figure 4.7, included side strippers for kerosene, light gasoil and heavy gasoil instead of side draws as previously used. Each side stripper had its own steam feed. The vapor crude feed from the pre-flash V103 was also connected to the column. Once the column with side strippers was converged, the solution constraint set up was changed so that distillation curve data of the outgoing fractions gasoline, kerosene, light gasoil and heavy gasoil was specified. For gasoline, the 95% TBP temperature was specified. For kerosene and light gasoil, the ASTM D1160 95% boiling point temperature was specified. For heavy gasoil, the cloud point was specified. The column then calculated the product flow rates. The tray efficiency was assumed to be 100%. The number of trays and the location of column feed and draw streams were based on the actual distillation tower at St1 refinery [14], see Figure 3.7.

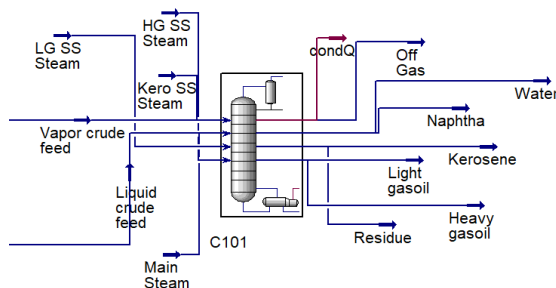


Figure 4.7: Final version of distillation column in Hysys with side strippers added

4.2 Test Runs

Multiple test runs were conducted at the refinery during which the temperature of the crude oil exiting the furnace was decreased stepwise. Figure 4.8 visualizes the working method of the test runs. In each round, the temperature of the outgoing crude feed from the furnace was decreased by 3°C , starting from normal operating conditions. The CDU was then run for 5 hours at the new temperature to allow the process to stabilize, before samples were taken of the outgoing products, see Figure 3.1 for sample point location. The samples were analyzed for boiling point range,

density and sulfur content. The product flow rates from the refinery during the test runs were compared to the model output flow rates.

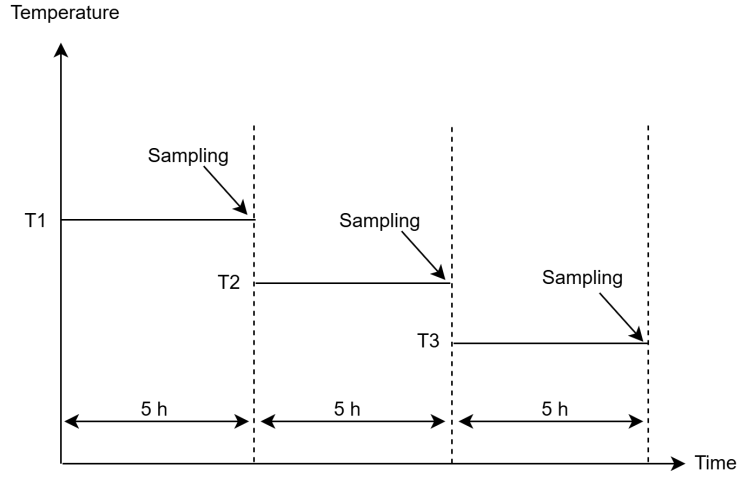


Figure 4.8: Set up of test runs

Table 4.4: Furnace outlet temperature during test runs

	Temperature 1	Temperature 2	Temperature 3
Test run 1	351°C	348°C	345°C
Test run 2	349°C	346°C	343°C

4.3 Performance Indicators

The following section defines the performance indicators of this study and how they were determined. Fuel consumption and steam consumption are considered primary performance indicators, while CO₂-emissions and economic potential are considered secondary performance indicators as they are linearly proportionate to the primary performance indicators.

4.3.1 Product Separation

Product separation was judged by the boiling point ranges. Separation was considered to be complete if the boiling point ranges for the different products did not overlap [13].

4.3.2 Model Accuracy

For each product flow rate, an absolute error and a relative error has been calculated according to Equation 4.2 and Equation 4.3.

$$E_{absolute} = |x_{simulation} - x_{measured}| \quad (4.2)$$

$$E_{relative} = \frac{E_{absolute}}{x_{measured}} \cdot 100\% \quad (4.3)$$

4.3.3 Fuel Consumption in Furnace

The fuel gas consumption to the furnace was monitored by flow meters, one to section A and one to section B of the furnace F101. The readings from the flow meters were used to evaluate the impact of the test runs. The fuel gas flow rate was divided by the crude feed rate to obtain the specific fuel gas flow according to Equation 4.4.

$$\dot{m}_{fuel\ gas\ F101, specific} = \frac{\dot{m}_{fuel\ gas, F101}}{\dot{m}_{crude\ feed}} \quad (4.4)$$

The specific fuel gas flow at each furnace outlet temperature was evaluated for the minimum value, the average value and the maximum value at each temperature.

4.3.4 Steam Consumption

The steam consumption in the distillation column and the side strippers was monitored by flow meters - one flow meter for steam to each side stripper and one for the steam to the main column. The readings from the flow meter were used to monitor the impact of the test runs. The steam flow rate was divided by the crude feed rate to obtain the specific steam consumption according to Equation 4.5. To simplify the evaluation of steam consumption, two scenarios for the production of stripping steam were considered: a worst case scenario where HP steam was let down to MP or LP stripping steam, and a best case scenario where MP or LP steam was produced in a waste heat recovery steam boiler. In the case of HP steam let down, it was assumed that 14.85 ton HP steam were produced per ton of fuel gas. The fuel gas needed to produce HP steam was then added to the fuel gas to F101 when considering the total fuel gas demand as shown in Equation 4.6. In the case of MP or LP steam generation in a waste heat recovery boiler, the stripping steam was assumed to be available in excess and thus not cause any emissions in the production process.

$$\dot{m}_{steam, specific} = \frac{\dot{m}_{steam, C101} + \dot{m}_{steam, C102} + \dot{m}_{steam, C103} + \dot{m}_{steam, C105}}{\dot{m}_{crude\ feed}} \quad (4.5)$$

The specific steam flow at each furnace outlet temperature was evaluated for the minimum value, the average value and the maximum value at each temperature.

$$\dot{m}_{total\ fuel\ gas, specific} = \dot{m}_{fuel\ gas\ F101, specific} + \frac{\dot{m}_{steam, specific}}{14.85} \quad (4.6)$$

4.3.5 CO₂-emissions

When calculating CO₂-emissions from the fuel gas it was assumed that the combustion reaction is complete, and the oxidation factor is thus unity. The emitted CO₂ was calculated as

$$\dot{m}_{CO_2, specific} = \dot{m}_{fuel\ gas, specific} \cdot \frac{x_C}{100} \cdot \frac{M_{CO_2}}{M_C} \cdot OF \quad (4.7)$$

where $\dot{m}_{CO_2}$ is the emission of CO₂ in ton/h, $\dot{m}_{fuel\ gas}$ is the consumption of fuel gas in ton/h, x_C is the carbon content in the fuel in mass %, and OF is the oxidation factor. The carbon content in the fuel gas was continuously measured by an instrument. It was assumed that the fuel consists only of carbon and hydrogen. The parameters used for calculating CO₂-emissions are presented in Table 4.5.

Table 4.5: Parameters for CO₂-emissions calculation

Parameter	Symbol	Value	Unit
Carbon content	x_C	78.4	wt. %
Oxidation factor	OF	1	-
Molar mass, CO ₂	M_{CO_2}	44.01	g/mol
Molar mass, C	M_C	12.01	g/mol

4.3.6 Economic Evaluation

The variable cost, OPEX, for operating the furnace and the distillation column with side strippers was evaluated for different furnace outlet temperature for a proposed yearly crude throughput of 6,000 ton/d according to Equation 4.8.

$$OPEX = \dot{m}_{crudefeed} \cdot (\dot{m}_{fuel, specific} \cdot C_{fuel} + \dot{m}_{CO_2, specific} \cdot C_{EU\ ETS}) \quad (4.8)$$

The considered costs were fuel cost, C_{fuel} , and EU ETS allowance cost, $C_{EU\ ETS}$. As fuel gas is not a commercial product, LPG price was used as an estimation to evaluate the impact of reduced fuel demand. A typical price for LPG was assumed to be 520 USD/ton. The cost of CO₂ was assumed to be 76.4 EUR/ton CO₂, which was the average price for the period 11th May 2025 - 11th May 2026 [19]. Euro and USD were converted to SEK using the average exchange rate of 10.88 SEK/EUR and 9.33 SEK/USD for the same period according to Sweden's central bank. A summary of the costs linked to fuel consumption and CO₂-emission are presented in Table 4.6.

Table 4.6: Parameters for variable operating cost calculation

Parameter	Value	Unit
EU-ETS price	831	SEK/ton CO ₂ -eq.
Fuel price	4,851	SEK/ton fuel

4. Method

A sensitivity analysis of changes to the fuel cost and the EU ETS cost was performed. A change of -20%, -10%, +10% and +20% for each of the cost parameters was considered.

5

Results & Discussion

5.1 Model Validation

The flow rates predicted by the model are compared to the measured values for the base case in Figure 5.1.

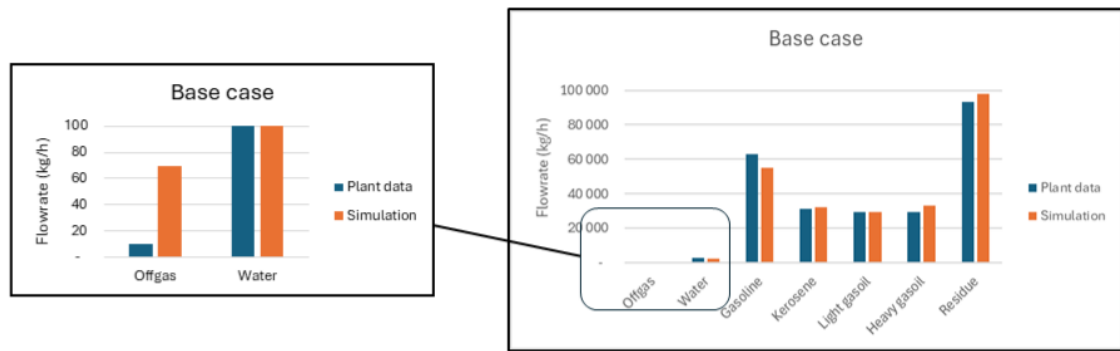


Figure 5.1: Comparison of product flow rates between the measured data and simulated data for base case

The errors between the measured data and the simulated data are compiled in table 5.1.

Table 5.1: Error analysis of model in base case

Product	Absolute error, E_{abs}	Relative error, E_{rel}
Gases	59 kg/h	623%
Water	100 kg/h	4%
Gasoline	8,031 kg/h	13%
Kerosene	1,151 kg/h	4%
Light gasoil	67 kg/h	0%
Heavy gasoil	3,709 kg/h	13%
Residue	5,010 kg/h	5%

The measured gas stream corresponds to 0.003% of the total product streams, thus the small difference in mass flow between the measured and simulated value (69 kg/h) translates to a large difference in relative error (623%). In addition to the gas stream, the largest relative errors are 13% for gasoline and heavy gasoil. One

possible source of error is the characterization information of the crude oils. These are provided from the crude supplier once at the start of purchasing a new crude, but then that field can be drilled for many years without updating the characterization. The specific cargoes used in the refinery during the case study could thus differ in composition from the available data. If there is a discrepancy between the input data into the model and the actual composition of the crude oil, then the output from the simulation will struggle to be 100% accurate.

The model was applied for the conditions during the test runs. Figure 5.2 shows the result for T2 and T3 from test run 2. The discrepancies between measured data and model are similar during the test run as for the base case.

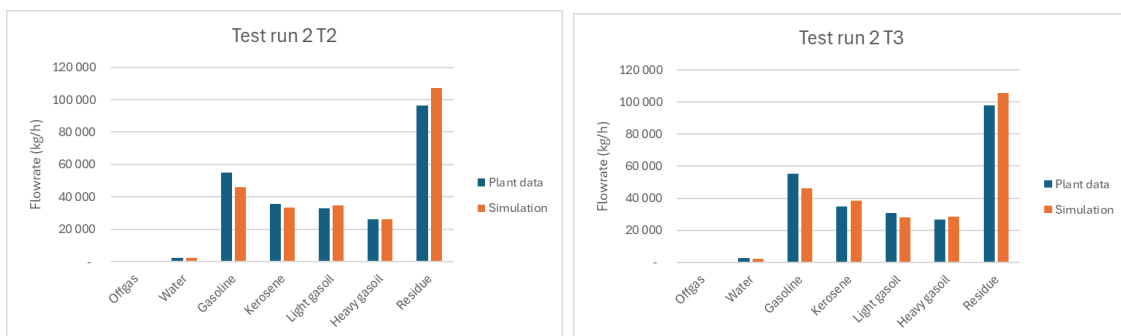


Figure 5.2: Comparison of product flow rates between the measured data and simulated data for T2 and T3 from test run 2

5.2 Source of Error during Test Run 1

A general comment is to be made about the results from test run 1. During the test run, the primary performance indicators experienced a large variation, which was unexpected. Variation in the primary performance indicators was closely related to variation in the crude feed rate. This variation in crude feed rate was unintentional. Upon investigation, it could be seen that of the four valves used to control the crude feed rate into the different loops in the furnace, two valves were operating in fully open mode (see Figure 3.6 for furnace layout). This was a consequence of the Advanced Process Controller trying to regulate to achieve more even outlet temperatures between the four loops in the furnace. In each section, one of the loops was too hot, prompting the APC to increase the flow rate. The setpoint for the desired flow rate was higher than what could be achieved by opening the valves to the maximum. Once the furnace outlet temperature was decreased during the test run, the decrease in temperature resulted in a lower portion of the crude feed evaporating in the furnace tubes. Less vapor decreased the volume, allowing for an unintentional increase in feed, as the valves were still completely open and driven by the higher setpoint. Figure 5.3 shows the setpoint (purple line) being higher than the crude feed rate (orange line) for loop A during test run 1. The valve is completely open, shown by the green values (105% open, or max open). This meant that this part of the process was not within an adjustable range. For comparison of

how the same display looked for the C-loop which was adjustable, with setpoint and flow rate aligning well, see Figure 5.4. Due to this phenomenon, the results from test run 1 didn't capture the response of the fuel gas demand to the implemented change. The results from test run 1 are available in Appendix A1.

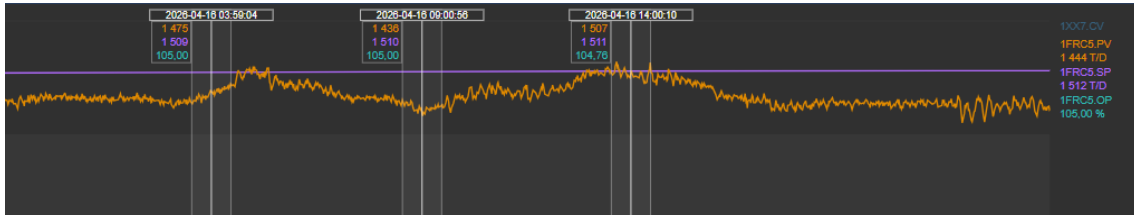


Figure 5.3: Setpoint vs flow rate in loop C, F101B

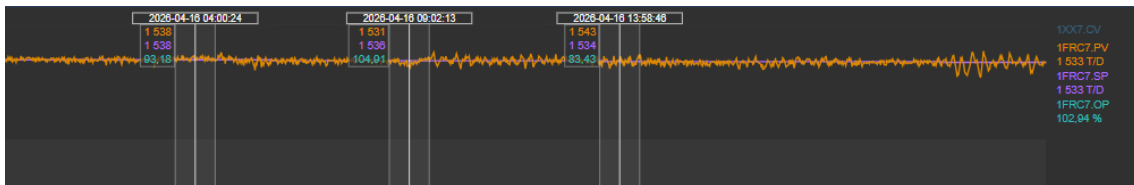


Figure 5.4: Setpoint vs flow rate in loop A, F101B

5.3 Product Separation

The boiling point analysis of the product samples from test run 2 are presented in Figure 5.5. The figure shows "gaps" between the 95% temperature for gasoline and the 5% temperature for kerosene, indicating clear separation of the two products. The 95% temperature for kerosene is equal to the 5% temperature for light gasoil, which also indicates clear separation. Light gasoil 95% temperature is "overlapping" with the 5% temperature for heavy gasoil, which indicates incomplete separation between these two products.

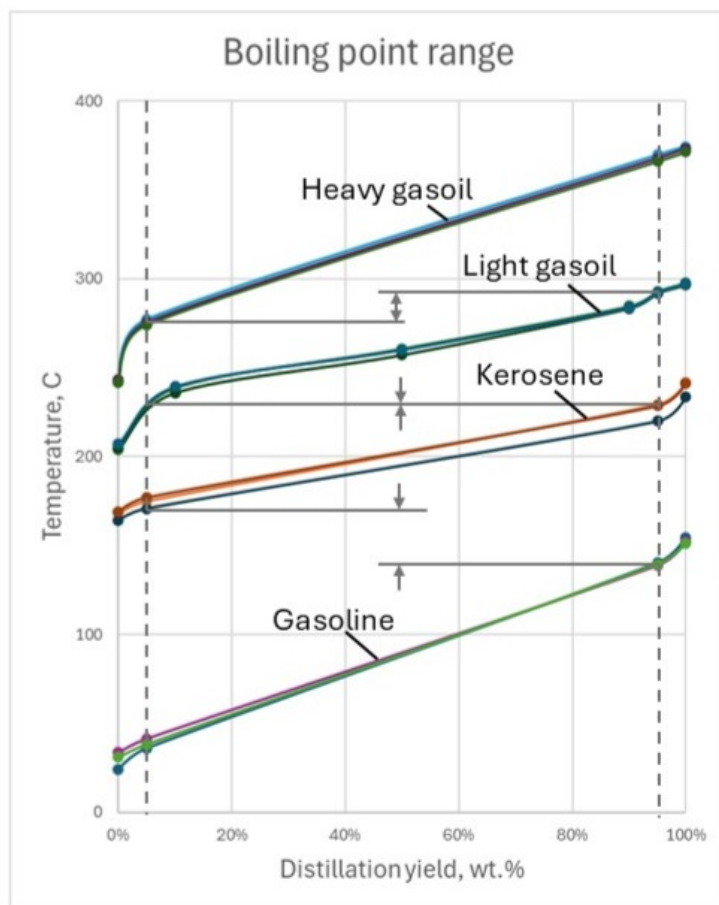


Figure 5.5: Boiling point range for products during test run 2

All furnace outlet temperatures resulted in similar boiling point curves for gasoline, and met the requirement of Final Boiling Point (FBP) < 180°C described in Section 3.8. Light gasoil met the requirement for 95% boiling point temperatures close to 290°C for all furnace outlet temperatures. For heavy gasoil, Figure 5.6 shows the cloud point at the first furnace outlet temperature T1 was 1.4°C higher than the target at 3°C. The following cloud points at T2 and T3 were both between -2°C to -3°C, which was complying with the requirements. That showed that the process was able to produce approved heavy gasoil even at the lowest tested temperatures. Although the heavy gasoil has a high cloud point at T1, this alone does not mean that the furnace outlet temperature is inappropriate, especially considering that all other analysis were in line with the targets at T1. The quality of the outgoing fractions is not jeopardized by a decrease of 10°C in furnace outlet temperatures from the base case at 353°C down to T3 at 343°C.

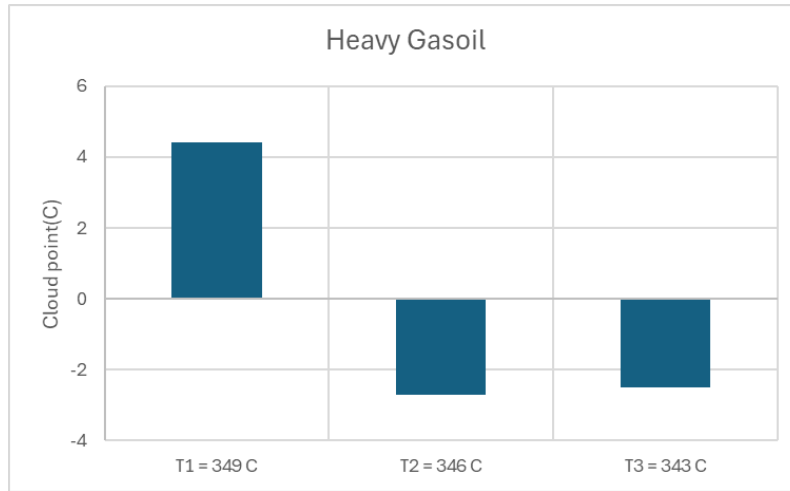


Figure 5.6: Cloud point of heavy gasoil test run 2

5.4 Fuel Consumption & Steam Consumption

The specific fuel gas consumption to the furnace during test run 2, calculated according to Equation 4.4, is presented in Figure A2. The result showed that lower furnace outlet temperature required less fuel gas per ton of crude feed.

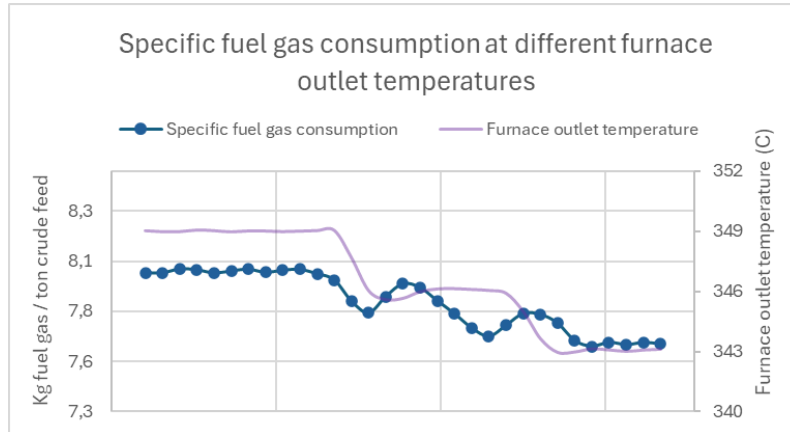


Figure 5.7: Specific fuel gas consumption test run 2

The specific steam consumption in the distillation column and the side strippers, calculated according to Equation 4.5, during test run 2 is presented in Figure A2. The result showed an increase in steam demand as the furnace outlet temperature was decreased, which means that more fuel gas is needed to produce HP steam in the case of using HP steam according to Equation 4.6.

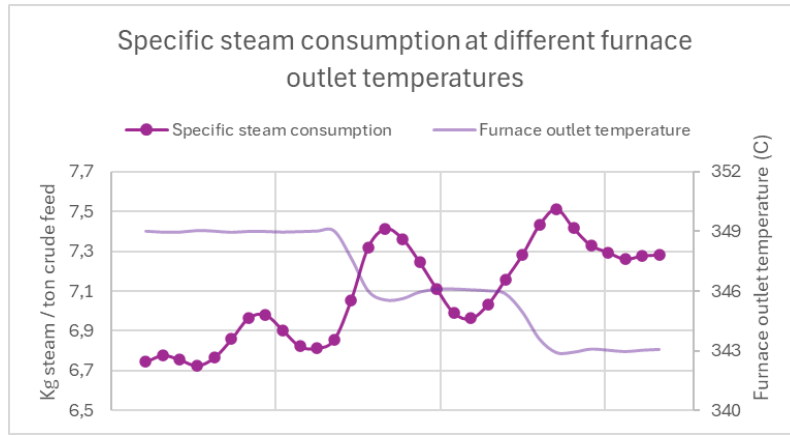


Figure 5.8: Specific steam consumption test run 2

Both the specific fuel gas consumption in the furnace and the stripping steam consumption showed a trend which overall correlated to the temperature profile, but both experienced some fluctuations, especially at the intermediate temperature T2 at 346°C. The changes in fuel gas consumption compared to the base case are presented in Table 5.2. In the table, the two scenarios for steam generation are considered and the minimum, average and maximum value at each furnace outlet temperature are included as described in Section 4.3.3. The lowest furnace outlet temperature T3 (343°C) achieved the biggest reduction in fuel gas demand compared to the base case. The reduction was largest for the scenario of excess steam availability.

Table 5.2: Reduction of fuel gas during test run 2 compared to base case

Furnace outlet temperature	$\Delta \dot{m}_{fuel\ gas}$ (Best case)			$\Delta \dot{m}_{fuel\ gas}$ (Worst case)		
	Min	Avg	Max	Min	Avg	Max
349°C	-12.8%	-12.4%	-12.2%	-12.6%	-12.1%	-11.8%
346°C	-15.9%	-14.3%	-13.0%	-15.4%	-13.7%	-12.3%
343°C	-16.5%	-15.8%	-15.2%	-15.7%	-15.0%	-14.3%

For a standard crude oil throughput of 6,000 ton/d the absolute yearly reduction of fuel gas (assuming HP steam and based on average values of operating conditions) amounts to 3,158 tons of fuel gas per year, which corresponds to approximately 2.2% of the refinery's total fuel gas consumption [14].

5.5 Emissions

The specific CO₂-emission calculated according to Equation 4.7 is presented in Figure 5.9. The calculation contains a linear relation between the fuel gas consumption and the CO₂-emissions, so the trends are similar to those for the fuel gas consumption itself. When the process is within the adjustable range as during test run 2,

lower furnace outlet temperature decreases specific CO₂-emissions (calculated according to Equation 4.7), as is shown in Figure 5.9.

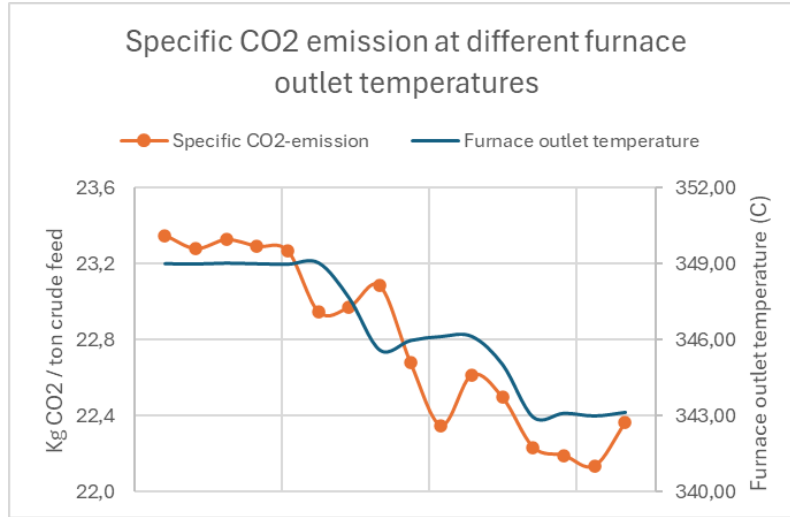


Figure 5.9: Specific CO₂-emissions test run 2

For a standard crude oil throughput of 6,000 ton/d the absolute yearly reduction of CO₂-emissions (assuming HP steam and based on average values of operating conditions) amounts to 9,072 tons of CO₂-emissions per year, which corresponds to approximately 1.4% of the refinery's total CO₂-emissions [14].

5.6 Economic Evaluation

The reduced demand for fuel gas from Table 5.2 and the reduced emissions connected to the use of the fuel gas shown in Figure 5.9 are evaluated from an economic perspective according to the assumptions presented in Section 4.3.6, using a fuel price of 4,851 SEK/ton fuel and a EU-ETS price of 831 SEK/ton CO₂-equivalents. The result for average fuel gas demand assuming HP steam at furnace outlet temperature T₃=343°C from test run 2 is shown in Figure 5.10 with the different cost contributions visible. For comparison, the cost for the base case is shown next to the bar for reduced furnace outlet temperature. The base case with furnace outlet temperature of 353°C has a yearly cost of 153 MSEK/year in total. The total yearly cost for furnace outlet temperature 343°C is 130 MSEK/year.

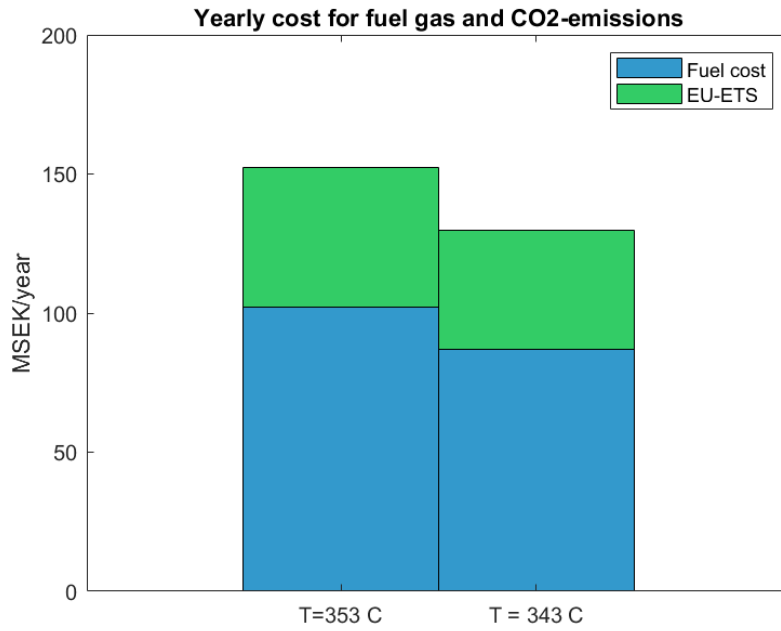


Figure 5.10: Yearly cost for fuel to furnace & emissions

A sensitivity analysis of changes to the fuel cost and the EU ETS CO₂ allowance cost is presented in Table 5.3. The result shows that changes to the fuel cost have larger impact on the total yearly variable cost than the CO₂ allowance cost, which is due to fuel gas being the larger contribute to variable operating cost to begin with.

Table 5.3: Sensitivity analysis of changes to fuel cost and CO₂ allowance cost

ΔC_{fuel}	OPEX @ 353°C	OPEX @ 343°C
-20%	132 MSEK/yr	112 MSEK/yr
-10%	142 MSEK/yr	121 MSEK/yr
+10%	163 MSEK/yr	138 MSEK/yr
+20%	173 MSEK/yr	147 MSEK/yr
$\Delta C_{EU ETS}$	OPEX @ 353°C	OPEX @ 343°C
-20%	142 MSEK/yr	121 MSEK/yr
-10%	147 MSEK/yr	125 MSEK/yr
+10%	158 MSEK/yr	134 MSEK/yr
+20%	163 MSEK/yr	138 MSEK/yr

5.7 Outlook

In the future, crude oil furnaces can potentially be electrified and the fuel gas can then instead be sold for other purposes. In this case, the furnace duty will be supplied by electricity instead of fuel gas. The furnace outlet temperature of 343°C corresponds to a furnace duty of 20.15 MW according to the Aspen HYSYS model.

The yearly furnace duty is then 176,540 MWh. If the electricity price is assumed to be 0.834 kr/kWh (Jan-April 2026 average for Elområde 3 in Sweden [20]), the yearly electricity cost will be 147 MSEK/year, which is less than the total variable cost for the base case but more than for furnace outlet temperature of 343°C. Valorizing the fuel gas is research topic in and of itself, but it can be expected to bring some additional income if sold.

6

Conclusion

This master thesis evaluates the impact of reduced furnace outlet temperature in a crude distillation unit at St1 Refinery on product separation, fuel demand and CO₂-emissions. The pre-flash, furnace and atmospheric distillation column have been considered and modeled in Aspen HYSYS. The model can be used to predict yield trends as impact of varied furnace outlet temperature. The test runs confirmed the possibility to successfully implement the reduced furnace outlet temperature. A requirement for reducing the furnace outlet temperature is to operate the furnace with process control adjustment margin, otherwise unwanted effects can occur. When this requirement was fulfilled, the product separation and quality was not jeopardized by reduced furnace outlet temperature. Reducing the furnace outlet temperature by 10°C, from 353°C to 343°C, resulted in lowering the fuel gas demand and CO₂-emissions by 15%. This corresponded to economic savings of 23 MSEK/year in fuel cost and CO₂-emission cost.

References

- [1] Anders Svensson. *Koppartrans*. Online. Aug. 2014. URL: <https://gamlagoteborg.se/2014/08/22/koppartrans/>.
- [2] D.Börjesson. “Personal communication with St1”. 2026.
- [3] E. Worrell et al. “Energy consumption by industrial processes in the European Union”. In: *Energy* 19.11 (1994), pp. 1113–1129. ISSN: 0360-5442. DOI: [https://doi.org/10.1016/0360-5442\(94\)90068-X](https://doi.org/10.1016/0360-5442(94)90068-X). URL: <https://www.sciencedirect.com/science/article/pii/036054429490068X>.
- [4] European Commission. *About the EU ETS*. Online. URL: https://climate.ec.europa.eu/eu-action/carbon-markets/about-eu-ets_en.
- [5] U.S Energy Information Administration. *Crude oils have different quality characteristics*. Online. June 2012. URL: <https://www.eia.gov/todayinenergy/detail.php?id=7110>.
- [6] *Aspen Hysys Properties and Methods Technical Reference*. Aspen Technology. 2013. URL: https://esupport.aspentech.com/apex/S_Article?id=000054059.
- [7] David S. J. Jones Steven A. Treese Peter R. Pujadó. *Handbook of Petroleum Processing*. Second edition. Springer Reference, 2015.
- [8] Kunru Yang et al. “Improving energy saving of crude oil distillation units with optimal operations”. In: *Journal of Cleaner Production* 263 (2020), p. 121340. ISSN: 0959-6526. DOI: <https://doi.org/10.1016/j.jclepro.2020.121340>. URL: <https://www.sciencedirect.com/science/article/pii/S0959652620313871>.
- [9] Juma Haydary and Pavlík Tomáš. “Steady-state and dynamic simulation of crude oil distillation using ASPEN Plus and ASPEN dynamics”. In: *Petroleum and Coal* 51 (Jan. 2009).
- [10] *Aspen HYSYS Unit Operations Reference Guide*. Aspen Technology. 2019. URL: https://aspentechsupport.blob.core.windows.net/knowledge/003392458_HYSYSUnitOpsV11-Ref.pdf?rscd=inline%3Bfilename%3D%22HYSYSUnitOpsV11-Ref.pdf%22&se=2026-05-25T11%3A53Z&sig=3pZeST5WYk6vCbIOUw%2Br%2BKFgVHV806snHvV8qtWwamE%3D&sp=r&sr=b&st=2026-05-25T11%3A43Z&sv=2019-12-12.
- [11] r. A. Shank and T. R. McCartney. “CHARACTERIZATION OF CRUDE OIL FOULING: DEFINING THE COKE SPECTRUM”. In: *Heat Exchanger Fouling and Cleaning*. 2019. URL: https://heatexchanger-fouling.com/wp-content/uploads/2021/09/04_Shank.McCartney.pdf.
- [12] F. Coletti, B.D. Crittenden, and S. Macchietto. “Chapter Two - Basic Science of the Fouling Process”. In: *Crude Oil Fouling*. Ed. by Francesco Coletti and

- Geoffrey F. Hewitt. Boston: Gulf Professional Publishing, 2015, pp. 23–50. ISBN: 978-0-12-801256-7. DOI: <https://doi.org/10.1016/B978-0-12-801256-7.00002-6>. URL: <https://www.sciencedirect.com/science/article/pii/B9780128012567000026>.
- [13] Gunter Alfke, Walther W. Irion, and Otto S. Neuwirth. “Oil Refining”. In: *Ullmann’s Encyclopedia of Industrial Chemistry*. John Wiley Sons, Ltd, 2007. ISBN: 9783527306732. DOI: https://doi.org/10.1002/14356007.a18_051.pub2. eprint: https://onlinelibrary.wiley.com/doi/pdf/10.1002/14356007.a18_051.pub2. URL: https://onlinelibrary.wiley.com/doi/abs/10.1002/14356007.a18_051.pub2.
- [14] St1 Refinery. “Internal documentation”. 2026.
- [15] Massimiliano Errico, Giuseppe Tola, and Michele Mascia. “Energy saving in a crude distillation unit by a preflash implementation”. In: *Applied Thermal Engineering* 29.8 (2009), pp. 1642–1647. ISSN: 1359-4311. DOI: <https://doi.org/10.1016/j.applthermaleng.2008.07.011>. URL: <https://www.sciencedirect.com/science/article/pii/S1359431108003177>.
- [16] R.Olander. “Personal communication with St1”. 2026.
- [17] St1 Refinery. “Archive data”. 2026.
- [18] *Aspen Hysys Simulation Basis Guide*. Aspen Technology. 2011. URL: https://aspentechsupport.blob.core.windows.net/knowledge/003465174_AspenHYSYSSimBasisV7_3-Ref.pdf?rscd=inline%3Bfilename%3D%22AspenHYSYSSimBasisV7_3-Ref.pdf%22&se=2026-05-11T11%3A59Z&sig=bpAhAomQ0eYAZoic4I1p0BnCwDwJakSTEqwCI%2BBK9rQ%3D&sp=r&sr=b&st=2026-05-11T11%3A49Z&sv=2019-12-12.
- [19] Trading Economics. *EU Carbon Permits*. Online. 2026. URL: <https://tradingeconomics.com/commodity/carbon>.
- [20] Konsumenternas energimarknadsbyrå. *Månadspriser på elbörsen*. Online. URL: <https://www.energimarknadsbyran.se/el/dina-elavtal-och-kostnader/elhandelsavtalet/elpriser-statistik/manadspriser-pa-elborsen/>.

A

Appendix 1

A1 Results from Test Run 1

Product Analysis

The results from test run 1 are shown in Figure A1 for product quality, in Figure A2 for fuel as consumption, in Figure A3 for steam consumption, and in Figure ?? for CO₂-emissions. The results do not show a correlation to the furnace outlet temperature due to the lack of process control discussed in section 5.2.

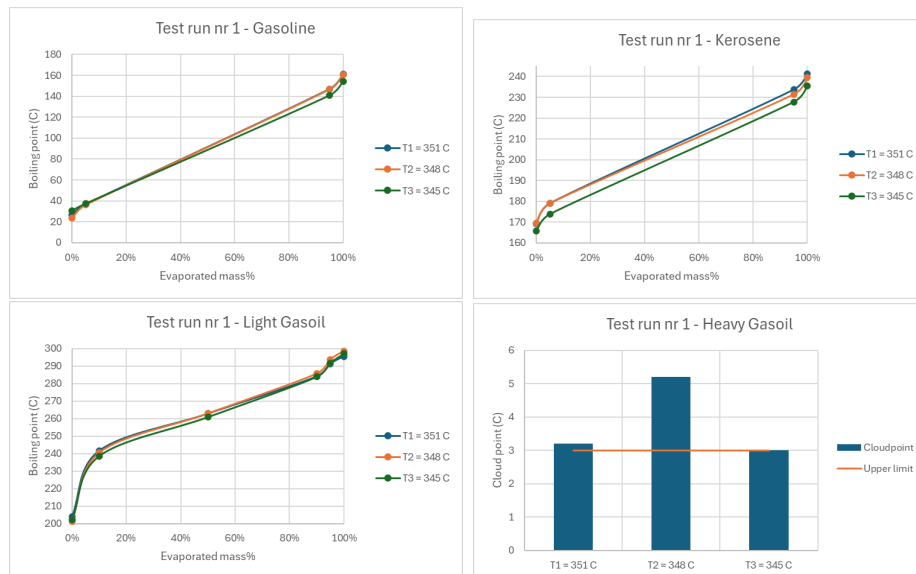


Figure A1: Product quality analysis test run nr 1

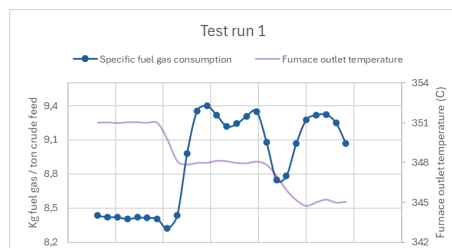


Figure A2: Specific fuel gas consumption

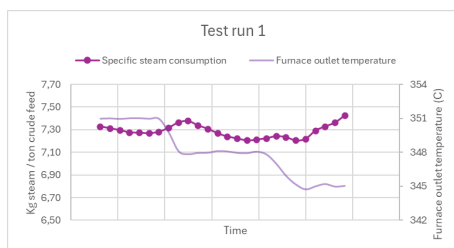


Figure A3: Specific steam consumption

DEPARTMENT OF ENVIRONMENTAL AND ENERGY SCIENCES
CHALMERS UNIVERSITY OF TECHNOLOGY
Gothenburg, Sweden
www.chalmers.se



CHALMERS
UNIVERSITY OF TECHNOLOGY



CHALMERS
UNIVERSITY OF TECHNOLOGY