

**PROCESS OPTIMISATION FOR OIL AND GAS PLATFORM WITH PROCESS
INTEGRATION TECHNIQUES**

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ABSTRACT

Heating utility system is commonly used in offshore platform to provide heating requirement for heavy crude oil of high viscosity from the reservoir. Heating the oil reduces its viscosity and improves oil/water separation which eventually leads to reduced size and cost of the separator. In conventional design, heating medium is distributed to the heat exchangers in parallel configuration to provide the required heating. All heat exchangers will receive the heating medium at supply temperature in this configuration. This parallel design however contributes to the overdesign of heating utility system. Besides, the phase separation system which separates the heated crude oil into oil, water and gas is designed independently. No prior study has been reported to investigate the interactions of the two systems. The independent design approach may lead to increase overall cost of both systems. Hence, it is the subject of this thesis to address the above research gap.

In the first scope of the thesis, *process integration* approach based on *pinch analysis* technique was used to design optimum heating utility system of offshore platform; the latter may be considered as a special case of the established *heat exchange network* (HEN) synthesis problem. A revised graphical targeting technique is proposed, which allows the identification of optimum heating medium requirement that fulfils both minimum flowrate and total cost solutions. The latter includes capital and operating costs trade-off between HEN and the waste heat recovery unit (WHRU). The proposed approach is based on the iteration of minimum approach temperature ($\Delta T_{\min}$) for selected values, which then identifies the optimum $\Delta T_{\min}$ that produces the lowest cost. The configuration of the heating utility system is designed using a revised *nearest neighbour algorithm* (NNA) proposed in this thesis.

In the subsequent scope, pinch analysis technique was implemented on an optimisation framework known as *automated targeting model* (ATM). The latter treats heating medium inlet and outlet temperatures as variables in order to determine an optimum solution. The ATM overcomes the tedious manual exercises performed in earlier approach, where calculations were repeated for each

different minimum approach temperature ($\Delta T_{\min}$) before an optimum solution can be determined. The ATM also caters for capital and operating costs trade-off for the heating utility system.

In the final scope of the thesis, a novel hybrid approach is proposed to solve the simultaneous optimisation problem for heating utility and phase separation systems in the oil and gas platform. The proposed approach makes use of the pinch-based ATM to find the optimum trade-off between heating utility system and phase separation processes. For the ATM, a novel algorithm is introduced to automate the process for temperature sorting and to locate the heating medium users in their associated temperature intervals. This is a new extension to the existing ATM, which requires user intervention to carry out these tasks. The revised ATM is capable in catering temperature changes of the heaters located upstream of the crude oil separators. The separators are sized independently for optimum cost at different outlet temperatures of the heaters; the latter leads to different oil properties which affect the sizing of the separators. This optimised separator cost is then used for simultaneous optimisation of heating utility and phase separation systems.

In the last chapter, the thesis is concluded. Future extensions on the research works are also proposed in this chapters.

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LIST OF ABBREVIATIONS

ABC	Artificial bee colony
ATM	Automated targeting model
ENNA	Enhanced nearest neighbour algorithm
ERA	Energy recovery diagram
FEED	Front end engineering design
FOA	Falcon optimisation algorithm
FPSO	Floating production storage off loading
GA	Genetic algorithm
GCC	Grand composite curve
HEN	Heat exchanger network
HIWN	Heat integrated water network
HMCC	Heating medium composite curve
HRL	Heat recovery loop
HRPD	Heat recovery pinch diagram
LP	Linear programming
LQR	Low quality region
LR	Local recycle
MILP	Mixed integer linear programming
MINLP	Mixed integer non-linear programming
MPTA	Modified problem table algorithm
NLP	Non-linear programming
NNA	Nearest neighbour algorithm
PFD	Process flow diagram
PSO	Particle swarm optimisation
PTB	Per thousand barrels of crude oil
HQR	High quality region
RCN	Resource conservation networks

SA	Simulated annealing
S&W	Sediment and water
STHE	Shell and tube heat exchanger
SWS	Stage wise superstructure
TAC	Total annualised cost
TEG	Thermoelectric generator
TSHI	Total site heat integration
WHRU	Waste heat recovery unit

LIST OF SYMBOLS

U	Overall heat transfer coefficient
s	Tube wall thickness
h	Film heat transfer coefficient
A	Area
AF	Annualised factor
M	Constant depending on equipment type
q	Stream enthalpy
T	Temperature
Dia	Diameter
P	Power
d	Head loss at equipment and differential height required
hl	Head loss
Hgt	Height
CP	Specific heat capacity
F	Flowrate
C	Cost
N	Number of units or shells
a	Installation cost
b	Area cost
W	Duration
G	Electricity tariff
i	Fractional interest rate per year
SR	Source
SK	Sink
Q	Enthalpy
dQ	Heat load
rdQ	Reverse heat load

ST	Heating medium supply temperature
v	Velocity
wt	Weight
Re	Reynolds
CL	Loss coefficient
cf	Fanning factor
L	Length
J	Pipe costing coefficient
WHRU	Waste heat recovery unit
O	Order of superiority
Tint	Temperature intervals
X	Total temperature levels are larger than the other temperatures
Z	Interval level
K	Separator sizing coefficient
Vol	Volume
rt	Residence time
R	Horizontal cylinder partial volume coefficient (liquid filled fraction of the vessel- dimensionless)
ht	Depth from liquid level to bottom of vessel divided by vessel inside diameter (dimensionless)
CD	Drag coefficient
TP	Vessel thickness
E	Fractional weld efficiency
CA	Corrosion allowance
PS	Pressure
DW	dilution water
AC	bbl brine in S21 per bbl of clean crude
BC	bbl brine outlet to storage tank per bbl of clean crude

VF	Volume fraction
Sl _t	Salt
Subscript	
in	Inlet
inn	Inner
o	Outer
out	outlet
oin	Outer based on core tube surface area
tb	Tube
opt	Optimum
P&F	Piping and fittings
g	gas
W	Weighted
C	Core tube
g	Gravity
HM	Heating Medium
HMR	Heating Medium Return
st	Shaft
U	Utility
PP	Pump
Min	Minimum
Max	Maximum
LGM	Log mean
t	Intervals
<i>i</i>	Source
<i>j</i>	Sink
m	Heating medium user
h	Hot

c	Cold
P	Pinch
bfly	Butterfly valve
chv	Check valve
gtv	Gate valve
gv	Globe valve
Ent	Entrance
<i>n</i>	Available outlet temperatures options
k	Mirror temperature
z	Interval level
SB	Souders-Brown
Liq	Liquid
wtr	Water
eff	Effective
tot	Total
od	Oil droplet
actg	Actual gas
wd	Water droplet
int	Internal
vss	Vessel
st	Steel
ww	Wash water

Superscript

<i>n</i>	Number of years
B	Base
<i>M</i>	Constant depending on equipment type

Greek Symbols

λ	Heat conductivity
η	Efficiency
Δ	Difference
ρ	Density
μ	Viscosity

Binaries variables

ω_t	For selecting HMCC and heating medium temperature at interval, t that produces $\Delta T_{\min}$ value
φ_1	For selecting Fanning factor
φ_2	For selecting Fanning factor
φ_3	For selecting Fanning factor
φ_{gv1}	For selecting loss coefficient globe valve
φ_{gv2}	For selecting loss coefficient globe valve
φ_{gtv1}	For selecting loss coefficient gate valve
φ_{gtv2}	For selecting loss coefficient gate valve
φ_{chv1}	For selecting loss coefficient check valve
φ_{chv2}	For selecting loss coefficient check valve
φ_{bend1}	For selecting loss coefficient bend
φ_{bend2}	For selecting loss coefficient bend
$\psi_{i,k}$	For selecting larger temperature
$\zeta_{i,t}$	To relate order of superiority to binary ψ
$\delta_{m,z}$	For determining whether an outlet temperature exist in an interval
$\tau_{m,z}$	For determining whether an inlet temperature exist in an interval

CHAPTER 1

INTRODUCTION

1.1. Problem Background

In the past few years, the oil and gas industry went through one of its difficult periods in the past decade. The price of the crude oil lost 81% of its peak price, USD 161/barrel in June 2008 to USD 30/barrel in Jan 2016 (Crude Oil Price History Chart, 2018). This triggered many oil producing companies to reduce their operation management budget and also to postpone new exploration projects to the future. In response to the low oil price, member countries of the Organisation of the Petroleum Exporting Countries (OPEC) agreed to reduce their total production output to stabilise the crude price since 2017. Since then, the crude oil has seen some recovery in price, with the current price stables at approximately USD 62-75/barrel since Jan 2018. Beside this drastic move, a longer term approach is required for cost efficient exploration and production of crude oil in order to last the vicissitudes of the market cycles.

Apart from the commercial side, the extraction and production of oil and gas from petroleum fields is generally energy-intensive and known to cause significant environmental impact. Oil and gas processing platform consumes large amount of power ranging from 10 to several hundred Megawatt power, depending on the properties of the oil field, the system design set-up, and the export specifications (Nguyen et al., 2014). The power generated from the combustion of diesel and fuel gas for on-site power releases large quantities of CO₂ to the atmosphere. Besides, flaring and venting practices to remove unwanted gas emits CH₄ to the atmosphere, which is more harmful than CO₂-emissions with regard to global warming (Nguyen et al., 2014). Oil and gas platforms also discharge chemicals such as biocides and methanol used for the treatment of the produced water effluents and cooling water to the sea. These environmental impacts can be managed more efficiently with better design approach to oil and gas platform.

1.2. Design improvement for offshore platform

There are a few design changes that can be carried out in order to improve the energy usage in oil gas platform and to reduce the environmental impact. One of them is the installation of multi-level production manifold which will allow gas extracted at high pressure from the wells to be processed separately at higher pressure without mixing with low pressure crude oil from different wells. According to Nguyen et al. (2016) this configuration of multiple operation levels in the production manifolds may result in significant energy savings provided that the pressure levels and well allocations are selected adequately to minimise the power consumption of the processing plant, while at the same time ensuring high recoveries of light and heavy hydrocarbons in the gas and oil streams, respectively. The downsides of this modification are that more cooling requirement will be needed and increment in system complexity. Note that, this modification is more suited at design stage rather than retrofitted to an existing platform.

Installation of multiphase flow expanders could help in generating additional power to the oil and gas platform. The flow expanders will replace the valves at the production manifold and able to harness energy from the feed streams that are at high pressure and temperature. This modification however has several drawbacks. First, it is only viable to fields with high gas fraction, pressure and temperature. Another drawback is that the performance of the expander may not be consistently efficient over the field lifetime since the production of oil, gas and water varies significantly through time. An expander designed for early production where water extraction is at its minimum may become inefficient towards the end of the life of the fields. Impurities and sand in the crude oil may also hamper the installation of expander, as designer will need to consider the impact of erosion on the equipment. Additional cost is required for installing sand treatment system or removal system for other impurities.

Avoiding the recirculation of gas around the compressors may lead to better energy efficiency in oil and gas platform since additional power demand at the compressor and cooling requirement can be eliminated. Recirculating gas help to prevent surge at the compressor and ensures the compressor is operated at high efficiency when gas flow rate upstream of the compressor is reduced. Anti-surge

recirculation can be avoided by designing the compression train not only capable operating at high efficiency at peak condition but also at acceptable efficiency during part load condition. Besides, installing smaller parallel compression trains to delay the start of off-design operation may in limiting the anti-surge recirculation. This modification will require case to case evaluation since there is no two platforms that has the same compression demand.

Apart from the above mentioned modifications, the application of heat integration in offshore platform can also help in minimising the energy usage. Nguyen et al. (2016) studied the use of central utility system to recover energy from cooling utility system into heating utility system. However, this has little promises, since most heating demands take place at temperature higher than the temperature of the cooling water. While the opposite scenario is observed for heating medium, where most cooling demands requires lower temperature than the temperature of the heating medium. In this case, the cooling demands take place at temperature lower than the temperature of the heating medium. There is however no detailed study on heating utility system itself. In conventional design, heating medium is fed to all users at its supply temperature in a parallel arrangement. This configuration may not lead to optimum design since not all users will require heating medium at the highest temperature. Besides, the amount of heat provided at the heat exchanger has a direct impact not only on the cost of the heating utility system, but also on the phase separation system downstream of the heat exchangers. To date, no systematic approach has been proposed to determine the optimum design of heating utility and phase separation systems. Research conducted by de Oilivera and Van Hombeeck (1997) has shown that petroleum heating and separation steps are the most inefficient exergy users in the oil and gas platform. This thesis aims to address this design problem with the use of process integration methods, with the aim to find an optimum cost solution.

1.3. Process Description

Figures 1.1 and 1.2 show the heating utility and phase separation systems for an oil and gas platform, respectively. The main function of the heating utility system is to generate and provide heat to the various oil (heavy/light/crude) heaters, which is used to break the emulsion that is formed in the crude oil. As shown in Figure 1.2, the heaters are located upstream of the production/electrostatic

separators. Other users of the heating medium include the test separator heater (used during crude oil testing), maintenance heaters (such as diesel flushing/pigging heaters), closed drain drum heater and high pressure flare knockout drum heater (which are used during the facility maintenance). Note that these users are only in service intermittently, and hence are excluded from the analysis. The waste heat recovery unit (WHRU, Figure 1.1) uses flue gas from gas turbine compressor and/or generator to provide heating to the heating medium in rising its temperature. In the conventional parallel design of the heating utility systems, all heaters (which are the main element of the *heat exchanger network* – HEN) receive the heating medium at the supply temperature. The heating medium then leaves the HEN to the heating medium pump for recirculation. The latter will raise the pressure of the heating medium above its saturation temperature, so to account for pressure losses throughout the system. Another process unit in the heating utility system is the expansion tank which serves to provide buffer volume in the event of abnormal operation such as expansion of inventory in the system. For protection against overheating, a heating medium dump cooler is installed in the system (see Figure 1.1). Oil, water and mixture of glycol and water are commonly used as the heating medium. Oil is used for high temperature heating while the latter two are used for low temperature heating. Oil is less suitable from safety standpoint, however, due to its higher operating temperature range, it requires smaller heat exchanger as compared to when water or glycol mixture are used (McKetta, 1990).

Separation of oil, gases and processed water in oil and gas platform is normally carried out at the central processing unit. The reason is that, it is more economic to collect crude oil from different reservoirs to be processed centrally, rather than having the separation done at individual facilities. Figure 1.2 shows two different types of crude oil (heavy and light) entering their dedicated slug catchers. Heavy crude oil is more viscous in nature and has an American Petroleum Institute (API) gravity value below 30°; while those with equal or greater than 30°API are referred to as light crude oil (Canadian Legal Information Institute, 1996). In Figure 1.2, the slug catcher is used to collect and store liquids that have settled in pipelines during maintenance procedure known as pipelines pigging.

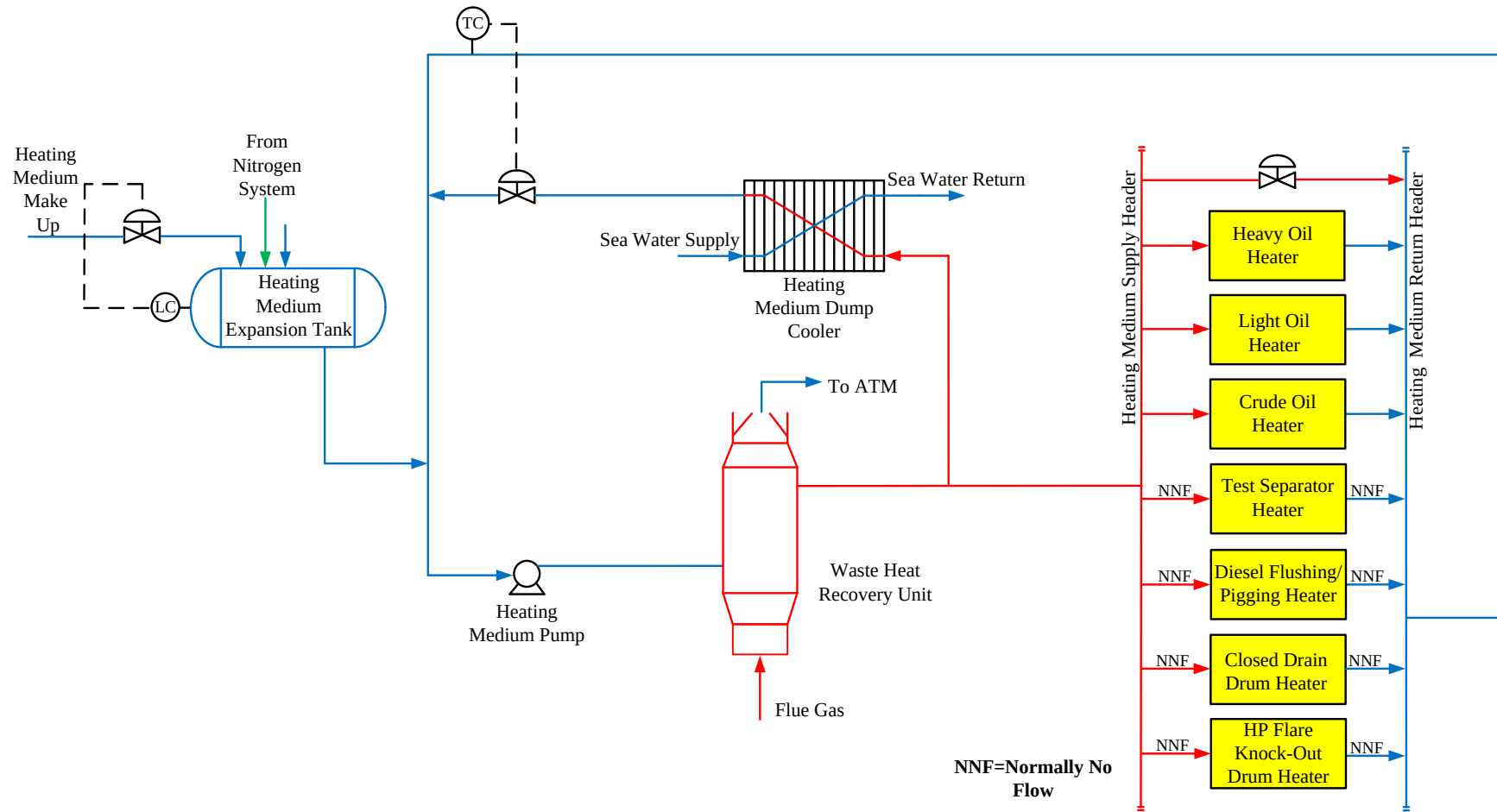


Figure 1.1. PFD diagram for heating utility system for oil and gas platform

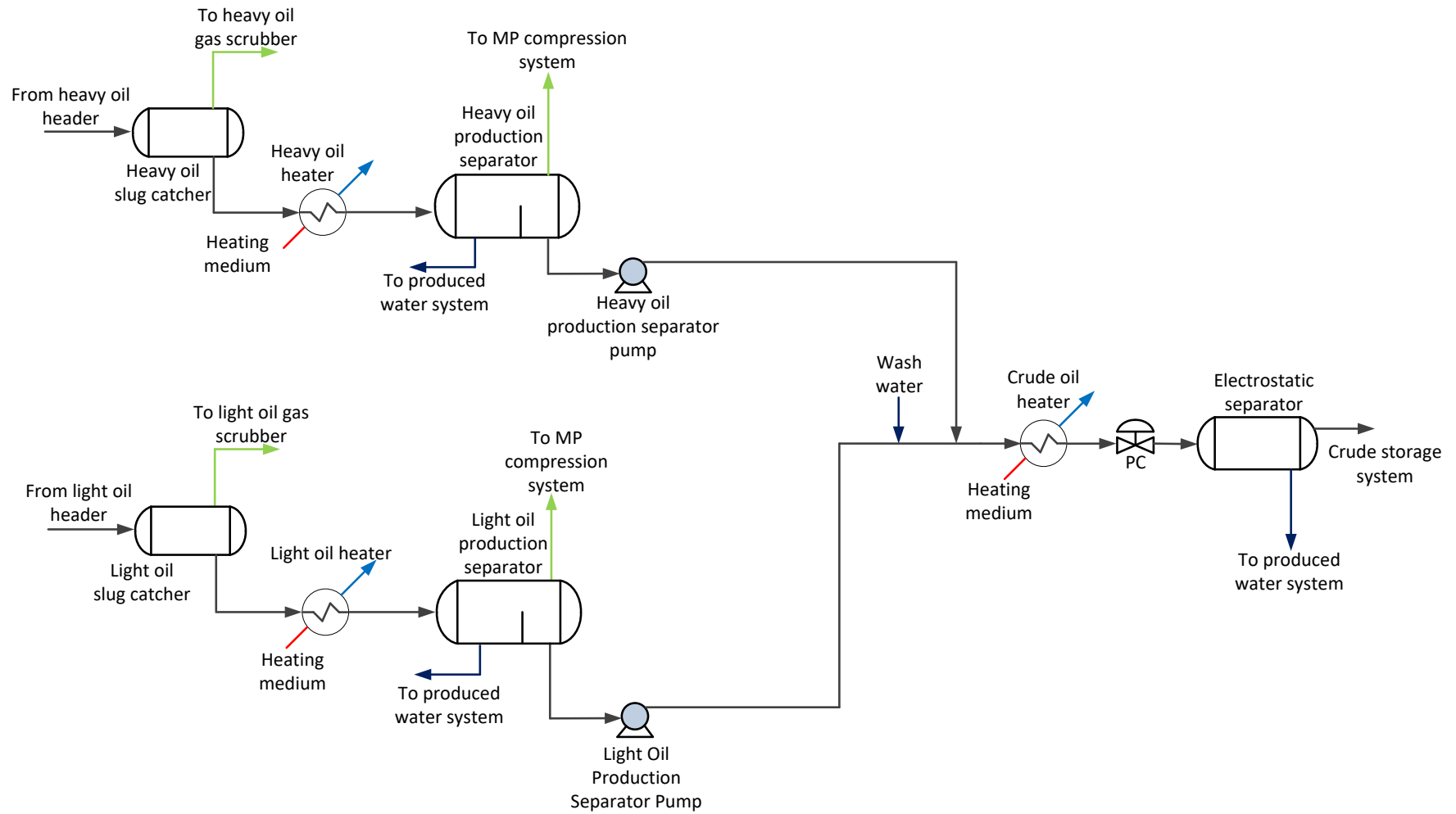


Figure 1.2. PFD diagram for phase separation system for oil and gas platform

During the process, devices referred to “pigs” are inserted into pipelines to perform cleaning and inspection activities. Besides, they also serve to separate most gases from the liquid phase. The separated gas is sent to the gas scrubber where its condensate and water content are removed. The separated liquid from the slug catcher is sent to their respective heaters prior to entering the light and heavy oil production separators.

Properties of the crude oil, e.g. density, viscosity, volumetric flowrate will change when the oil is heated, hence affecting the size of the separators. The latter is used for the separation of carry over gas, oil and water. The separated gas is sent to the medium pressure compression system while the processed water is sent to water treatment system, where processed water will be treated for reuse in the water injection systems. The separated light and heavy crude oil are then pumped to a common header where they are mixed. Salt content in the crude oil is a concern in field processing, as brine solution is corrosive for pipelines and oil vessels. Besides, the mixed oil still contains some sediment and water content which needs to be reduced to the permissible contents, i.e. 0.1 – 3.0 wt% (Manning and Thompson, 1995), in order to meet the sales specification. Dehydration and desalting processes are carried out in the electrostatic separator in order to meet these requirements. Wash water with low salt content is added upstream of the heater to dilute the salt concentration. At the electrostatic separator, electric current is used to further break down the emulsion in the crude oil. The produced water separated from the crude oil is sent to the water treatment, while the crude oil (which now meets the required specification) is sent to crude oil storage system, before being transported to onshore processing facilities.

1.4. Problem Statement

The heating utility system has always been designed conventionally with parallel arrangement where all heating medium users (heat exchangers) receive heating medium at the supply temperature. Literature by Iranian Petroleum Standard (2003) as well as by Ennis (2016) support this claim. This approach may not lead to optimum design of heating utility system, as not all users require heating medium at the highest temperature. The use of process integration may help in reducing the overall cost of the system by reusing some of the heating medium to another user with lower temperature

demand. At the moment, the heat exchanger model studied in field of heat integration commonly deals with process-to-process heat exchangers, assuming that external utility (e.g. stream) is available at its supply conditions (e.g. with fixed inlet and outlet temperatures). While there are models on the utility system, they are not extended far enough to consider variables such as inlet and outlet heating medium temperatures, pipe and fittings losses, pumping and regeneration cost all at the same time. Besides, by only optimising a single system without consideration of its interaction with other systems may only lead to suboptimal result. The heating utility system is closely tied to the phase separation system. The amount of the heat supplied at the heat exchanger affects the size of the separator and its cost. This relationship is complex to study due to the series arrangement of the crude oil heater, where the upstream outlet temperature of light and heavy oil heaters effect the inlet temperature of crude oil heater. Due to this arrangement, the chemical properties of the crude oil inlet stream have to be calculated each time the upstream heaters outlet temperature is changed. Pinch-based approaches provide a great deal of insight for process integration even when the scale of problem is complex. Graphical approach is the simplest form of the approach, capable of providing a reasonable solution to be used a yardstick for more complex approaches. Alternatively, pinch-based mathematical model avoids iteration and is able to include constraints that model the industrial problem more realistically.

1.5. Research Objective

The objective of this research is to produce a systematic approach to design heating utility and phase separation system that produces the lowest total annualised cost (TAC). The main objective can be further elaborated as follow:

1. To develop a heat integration model specifically for heating utility system that considers the impact of variable heating medium temperatures on the overall cost of the system.
2. To develop a fully-automated targeting algorithm for HEN design that achieves minimum cost solution.
3. To perform simultaneous optimisation of heating utility and phase separation systems with variable enthalpy load at the heat exchangers.

1.6. Thesis Outline

Chapter 1 introduces the background of the problem faced by the oil and gas industry and potential modifications that can be done to improve the offshore platform design. A detailed process description of heating utility and phase separation systems was also provided in this chapter along with the objectives of this thesis.

Chapter 2 presents a critical review on various methods for design improvement of heating utility system. In the first two sections, design and optimisation of heat transfer unit, heat exchangers and WHRU were considered. Next, literatures on the use of thermoelectric generator (TEG) system to recover more waste heat were presented. Beside the use of TEG, installation of steam bottoming cycle and organic Rankine cycle on onshore platform were reviewed. Heat integration approaches that were established for the optimum synthesis of HEN are also reviewed in detail in this chapter. This is followed by a review of the systematic approaches developed in the past for flowsheet optimisation which considers other systems together with HEN. At the end of this chapter, research gaps and opportunities for improvement are identified as the area of concern for this research thesis.

Several research scopes are proposed in Chapter 3 based on the research gaps highlighted in the preceding chapter. Next, systematic research methodology to address these research scopes is discussed in detail.

In Chapter 4, a revised graphical targeting technique based on Pinch Analysis is proposed for the heating utility system of offshore platform. The model explores capital and operating costs trade-off among HEN, WHRU and pump. The heating medium inlet temperature is fixed, while the outlet temperature is set as variable in determining the optimum cost solution. A revised *Nearest Neighbour Algorithm* (NNA) is proposed for HEN design. Literature example and an industrial case study are used to elucidate the proposed approaches.

In Chapter 5, the graphical model presented in Chapter 4 is incorporated into an optimisation framework known as *automated targeting model* (ATM). Both heating medium inlet and outlet temperatures are set as variables in the model. Piping cost together with pipe and fitting losses are

also added as additional costing variables in the model. Hypothetical examples and industrial case study are used to elucidate the newly extended technique.

In Chapter 6, the ATM model is improved for the use in simultaneous optimisation of heating utility and phase separation systems. In order to accommodate changing enthalpy, a novel algorithm is proposed to automatically sort the heating medium user temperatures in descending order and to locate them in the correct temperature intervals. The model considers costing for both separator and heating utility systems when determining the optimum solution. An industrial case study is used to elucidate the application of the newly proposed approach.

In Chapter 7, concluding remarks on the works presented in this thesis are provided to the readers. Future research opportunities on improving the current work are also identified.

CHAPTER 2

LITERATURE REVIEW

2.1 Heat exchanger optimisation

From Figure 1.1, it can be noticed that there are three continuous heaters and four intermittent heaters that are used to exchange heat between heating utility and its users (heat exchangers). Optimisation may be carried out on these heaters to optimise its size as well the unit cost. Chauduri et al. (1997) investigated the use of simulated annealing (SA) algorithm for the optimisation of area and linearised cost of shell-and-tube heat exchanger (STHE). The results showed that considerable savings was achieved (i.e. 84%) for area objective function and 74% for cost objective function, when compared with the base case. Meanwhile, Selbas et al. (2006) used a different stochastic approach known as genetic algorithm (GA) for the optimisation of STHE. They considered variables such as outside diameter of the tubes, flow arrangement of the tubes, number of tube passages, outside shell diameter and baffles spacing, leading to a total of 47,040 possible STHE design combinations. They concluded that based on the optimal designs obtained from GA provide significant improvement as compared to the traditional designs, with solutions obtained significantly faster than other methods. Particle swarm optimisation (PSO) has also been used for the optimisation of STHE by Patel and Rao (2010). They varied the shell internal diameter, outer tube diameter and baffle spacing in their optimisation model, with two tube layouts viz. triangle and square are also considered for optimisation. They carried four different cases with different duty requirement. In all reported cases, PSO outperformed GA. Other stochastic approach that has been used for STHE is artificial bee colony (ABC) by Sahin et al. (2011). They also considered similar variables as Patel and Rao (2010) and based on their result, ABC also outperformed GA.

Hadidi and Naziri (2013) applied biogeography-based algorithm (BBO) for the cost optimisation of STHE. In their model, apart from the capital cost of STHE, they also included discounted annual energy expenditures cost related to pumping fluid into STHE while varying tube length, tube outer diameter, pitch size, and baffle spacing. Based on their result, total saving of 56.1% was achieved when compared to an original design from literature, with large saving (96%)

contributed by the reduction in operating cost. Apart from original design, their result was also compared with other stochastic approaches i.e. GA, PSO and ABC with BBO outperforming all of them. In recent work by Segundo et al., (2019) falcon optimisation algorithm (FOA) was used for optimisation of STHE and plate-fin heat exchanger. For performance evaluation, they compared FOA against other stochastic algorithm i.e. GA, PSO, bee algorithm and firefly algorithm based on twelve single-objective benchmark test functions. Apart from one test function (Styblinski-Tang), FOA produced the best result when compared to other algorithms. For STFE cost optimisation (capital and operation), FOA also outperformed other algorithms when variables such as shell internal diameter, tubes outside diameter, and baffles spacing were considered in the model. Based on these literatures, the heat exchanger used in heating utility system can be improved in cost and value using some of these stochastic algorithms.

2.2 Waste heat recovering unit

A variety of heat exchangers (e.g. shell and tube, finned tube, tubular, brazed plate, etc.) can be used as waste heat recovery unit (WHRU, Fluid Dynamics Integrity, 2018). Each heat exchanger has its own design variables. For instance, Awais and Bhuiyan (2018) summarised the design variations for finned tube heat exchanger (see Figure 2.1), which include geometry of tubes, material and parameters e.g. fin pitch, tube pitch. These design variations can be explored using the stochastic approaches explained earlier for cost effective design of WHRU. New type of heat exchanger has also been studied for the use as WHRU. In the work of Tian et al. (2017), gravity pipe heat exchanger working principle was adopted for the use of waste heat recovery. Gravity pipe heat exchanger works with an intermediate medium (e.g. water) which transfer heat from exhaust gas to the cold stream. For the case of water, when it is heated by the hot stream, it evaporates and moves up to the upper section of the tube. It then transfers heat to the cold stream, which then condenses into liquid phase and flow back to the tube bottom via gravity, where it is reheated by the hot stream. Similar principle was applied for waste heat recovery with exhaust gas flowing in smooth tubes (Tian et al., 2017). For the latter, the cold stream (air) is passed against spiralled finned tubes containing the intermediate fluid. Based on an actual model installed in China, Tian et al. (2017) reported the natural gas savings of

15% was achieved from heating the air for combustion. This design principle can be extended for WHRU in the oil and gas platform; however, studies are required to determine the benefits for replacing the existing heat exchangers.

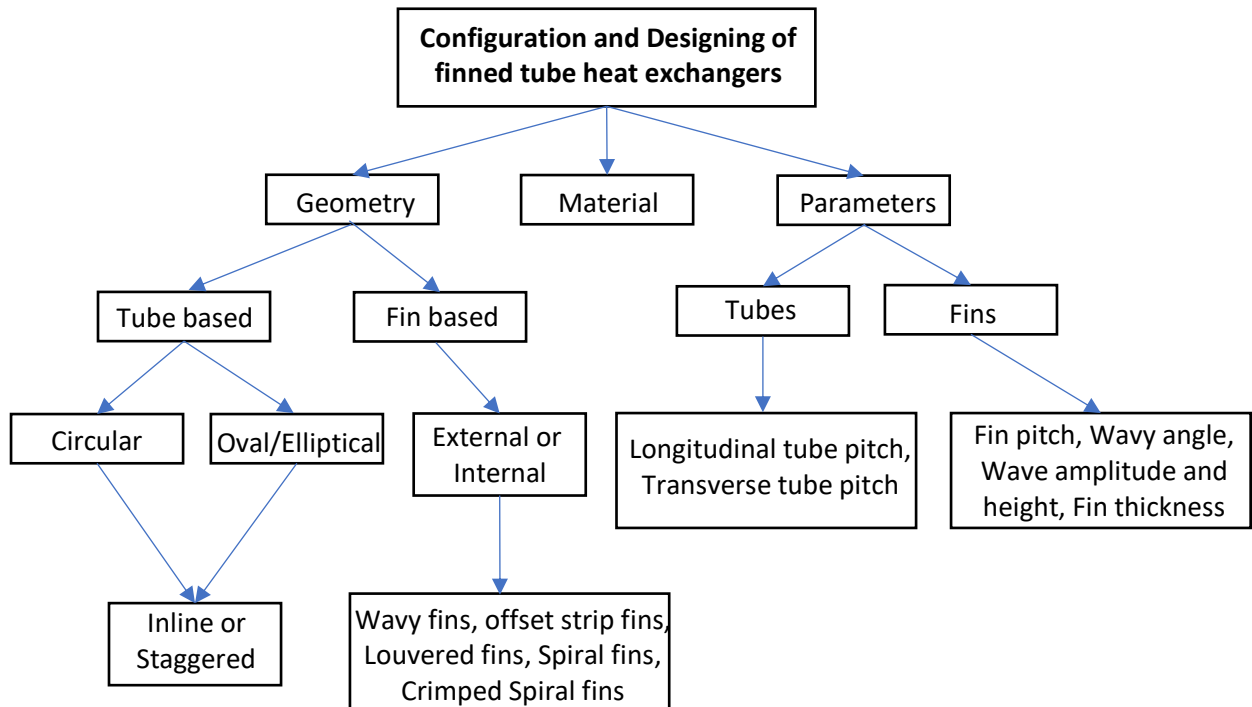


Figure 2.1: Configuration and design of finned tube heat exchangers

The use of thermoelectric generator (TEG) has been studied for electricity generation from waste heat. TEG are small tile shaped devices that operates as a heat engine, which converts heat into electricity using the Seebeck effect. Orr and Akbazadeh (2017) explored the prospects of waste heat recovery and power generation using TEG. They determined that 5.9 MW electricity can be possibly generated from a 500 MW of gas turbine power plant, which can power up 2950 homes that requires 2 kW each. This value was based on a conservative 0.59% conversion of waste heat into electricity. One of the challenges in using TEG is the low power generation efficiency. In order to overcome this, different design of TEG has been studied. Dai et al. (2011) carried out experimental study using liquid metal-based TEG system to harvest waste heat, which showed an improved efficiency of 2%. Remeli et al. (2015) explored a new method of recovering waste heat and electricity by integrating heat pipes into TEG. Their proposed method produced power output efficiency of 2.02%. In a recent work by Wang et al. (2016), a slightly higher efficiency of 2.05% was reported using metal foam filled TEG.

The optimum efficiency was obtained using metal foam with 5 pores per insert. This continuous improvement in power production efficiency of TEG helps to increase the benefit of adopting such unit against the incurred cost.

Installation of TEG as part of combined system with steam turbines cycle has been investigated by Yazawa et al. (2013). Their theoretical model studied the optimum point of operation that maximises power output and efficiency for the combined systems. The TEG considered was made of nano-structured material which is capable of producing high power output. Based on the result obtained, installing TEG with steam turbine cycle produces economic benefit which is attributed to effective reuse of a large amount of collateral heat losses that would otherwise occur within a standalone steam turbines cycle. In a recent work by Lan et al. (2018), a smaller scale vehicle waste heat recovery was studied via numerical and physical models. Their TEG system was made from commercial thermoelectric modules and counter flow heat exchanger which was installed to a heavy-duty diesel engine. Based on their tests, simulation of steady-state operating points as well as the response to typical engine cycle test showed good agreement with experimental data. Besides, they also concluded that the thermal contact conductance and thermal conductivity of hot side heat exchanger have major influences on the overall power produced. Physical model experiments on smaller heat waste system helps to accelerate actual application of larger scale TEG (e.g. waste heat recovery in oil and gas platform) in the future.

2.3 Installation of steam bottoming cycle or organic Rankine cycle system

Another possible option for more waste heat recovery is the installation of steam bottoming cycles, which can be used to generate electricity (Nguyen et al., 2016). Beside inhouse use of the generated electricity, power export can also be carried out if the platforms are connected to onshore grid or other facilities. Nguyen et al. (2016) carried out theoretical studies on the possible benefits of installing steam bottoming cycle on some platforms in the North Sea, with different gas turbines capacity and heating load requirements. Based on the result, they concluded that installing steam bottoming cycle results in a greater power capacity (if required), or in a lower fuel gas consumption and smaller CO₂ emissions (Nguyen et al., 2016). However, the authors highlighted that the addition

of the system to oil and gas platform will result in greater space and weight requirements, unless it replaces one of the existing gas turbines. Besides, Nguyen et al. (2016) also suggested that introducing steam bottoming cycle is a complex design task, as many layouts can be suggested, depending on the energy requirements of the platform and on the plant layouts.

Apart from the flue gas, low temperature waste heat is also available from gas after each compression stage before it is cooled. Due to the gas being at low temperature, steam bottoming cycle is not a possible option. Instead, organic Rankine cycle using fluids such as propane, carbon dioxide, or ethane-propane mixture can be used to generate electricity. Note however that according to Nguyen et al. (2016), the integration of organic Rankine cycles has never been proven in an offshore platform and heat recovery from the gas cooling steps is particularly challenging.

2.4 Process Integration

Apart from optimisation of heat exchangers, WHRU and installation of TEG and Rankine cycle systems, another important approach that can improve energy efficiency of the heating utility system is process integration. Process integration may be defined as *a holistic approach to process design, retrofitting and operation which emphasises the unity of the process* (El-Halwagi, 2006). Hohmann (1971) first introduced a graphical process integration technique for the synthesis of HEN in the late 1970s. This concept was later improved by Linnhoff and Flower (1978) with the introduction of algebraic targeting technique to determine the minimum utility requirement. Since then, many process integration techniques including pinch analysis and mathematical optimisation have been developed over the past few decades to find optimum solution for overall process system design, covering both energy and mass recovery systems.

2.4.1 Graphical approach in HEN

In the work first proposed by Hohmann (1971), heat integration was carried out by matching the hot process streams (*sources*) to the cold process streams (*sinks*) in a temperature vs enthalpy diagram.

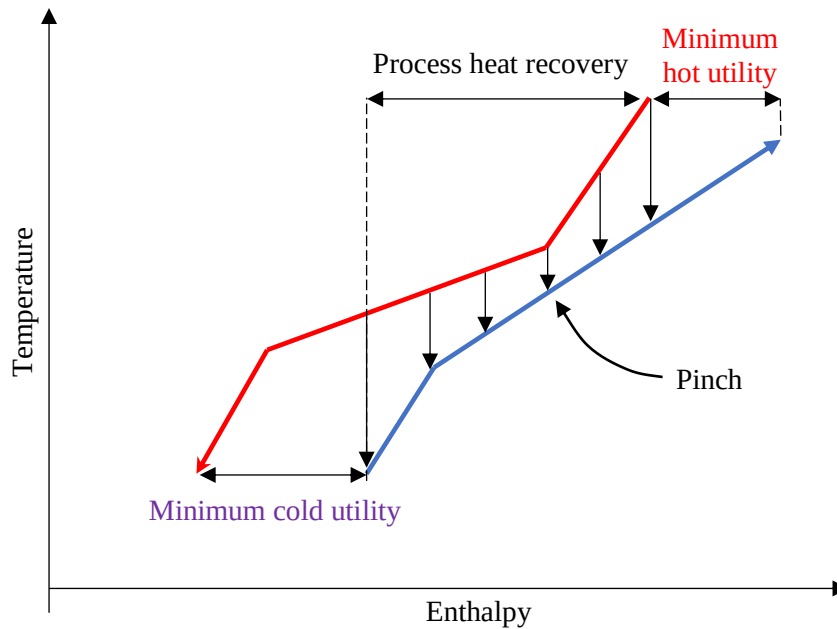


Figure 2.2. Composite Curves for minimum utility targeting (Hohmann, 1971)

The figure that is generated is known as the heat recovery pinch diagram (HRPD), which consists of hot and cold composite curves, which represent the sources and the sinks respectively. These curves can be shifted on the enthalpy axis to determine the minimum hot and cold utility target. The point of closest approach between the hot and cold composite curves is known as the pinch point (Figure 2.2) where minimum hot and cold utility requirements can be identified.

Beside the use of HRPD for minimum utility targeting, heat transfer area of the heat exchanger network (HEN) can also be determined based on Ahmad and Linnhoff (1990) “supertargeting” approach. The HEN area is targeted by calculating each enthalpy interval using the “Bath formula” as shown in Figure 2.3. The accuracy of area targeted using the simple method proposed by Ahmad and Linnhoff (1990) is within 10% of accuracy when the film heat transfer coefficient for the streams is within an order of magnitude. Note that, the model assumes vertical heat transfer. When the heat transfer coefficient varies significantly from stream to stream, Ahmad and Linnhoff (1990) “supertargeting” does not predict the true minimum network area. In this case nonvertical heat transfer is required to achieve the minimum area and linear programming can be used to calculate the area.

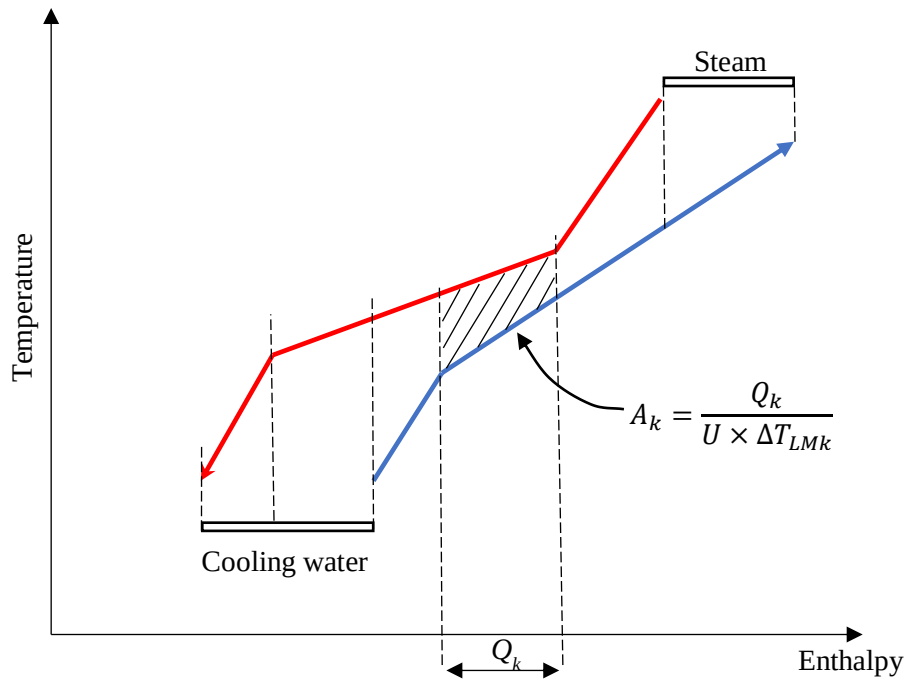


Figure 2.3. Composite curves for minimum utility targeting (Ahmad and Linnhoff, 1990)

2.4.2 Mathematical approach in HEN

Pinch-based graphical approaches provide great insights to the HEN synthesis problem, however they are tedious and suffer with inaccuracy problems. Some early developed mathematical approach such as the *transportation* model by Cerda et al. (1983) and *transshipment model* by Papoulias and Grossmann (1983) aimed to address this problem. In the transshipment model (Figures 2.4 and 2.5), heat is regarded as a commodity that is transported from a set of sources to some intermediate storage/warehouse (temperature intervals), before it is sent to a set of sinks (Gundersen, 1991). This differs from the transportation model (Cerda et al., 1983) where the commodity (heat) is transferred directly from the sources to sinks without any intermediate storage (Gundersen, 1991). These models are linear and can be solved easily. Note that the transshipment model has more compact formulation due to less equations. Based on the minimum utility targets, Papoulias and Grossmann (1983) showed that the fewest units and matches can be predicted with a MILP

transshipment model. A non-linear model can then be used to derive the network structure from the predicted heat recovery and matches.

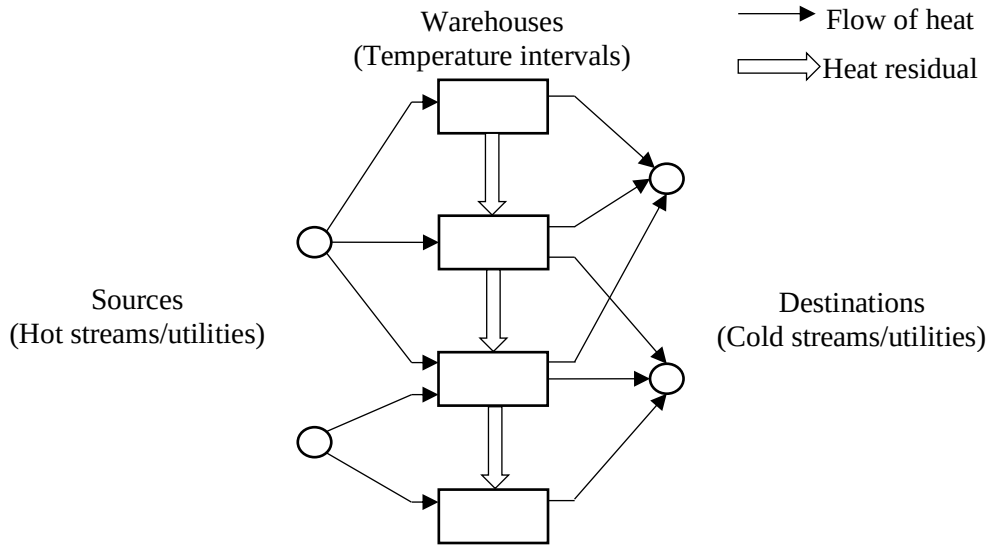


Figure 2.4. The transshipment model (Papoulias and Grossmann, 1983)

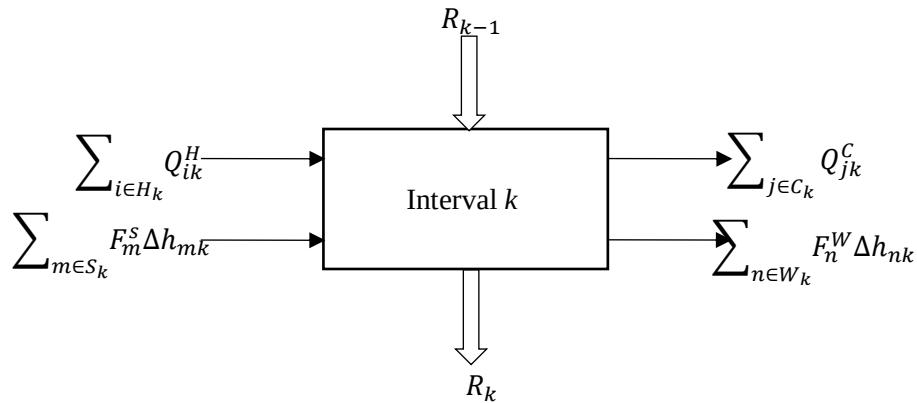


Figure 2.5. Heat flow pattern in each temperature interval (Papoulias and Grossmann, 1983)

In a recent work by Hong et al. (2017), they proposed a new transshipment model based on the original transshipment model (Papoulias and Grossmann, 1983) which is capable to synthesis the HEN. Their model considers the mass flow pattern of hot and cold streams on top of the heat flow pattern considered in the earlier model. With former inclusion, splitting, by-passing and recycling of streams, as well as isothermal and non-isothermal mixing of streams can be considered. The model requires

$\Delta T_{\min}$ value to be assumed in order to arrive at a solution, thus limits its ability to determine the optimum $\Delta T_{\min}$ that produces the lowest cost without iteration. This limitation can be overcome with the *stage-wise superstructure* (SWS; Figure 2.6) proposed by Yee and Grossmann (1990). The SWS minimises heat exchanger (fixed charges and area costs) and utility costs simultaneously, without the need for $\Delta T_{\min}$ iteration. The idea behind the SWS is the sequence of stages in which hot and cold streams are successively split and remixed. The number of stages is a variable, normally chosen from the larger number among the hot and cold streams. While SWS has the limitation for not considering all possible options for stream splitting (e.g. branches with two or more sequential matches), it has two major advantages. Firstly, imposing constraints for no stream splitting is relatively easy with SWS, in which binary variables are used to specify that not more than one match can take place for any given stream in each stage. Secondly, since SWS assumes isothermal mixing, no variables for the flowrates are required. This leads the constraints to become linear which greatly simplifies the solution of the MINLP model.

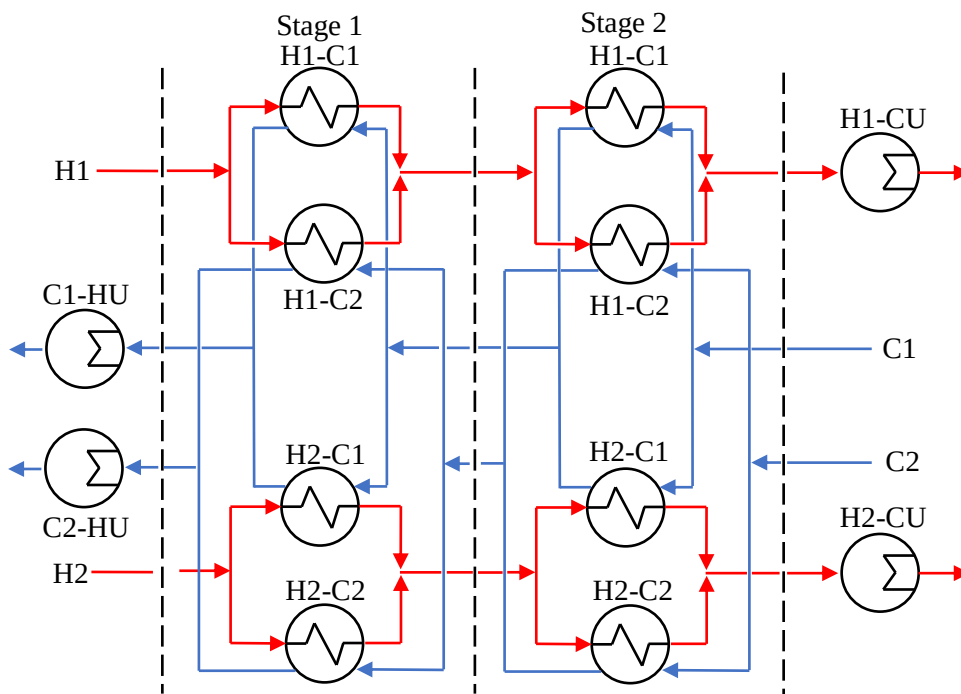


Figure 2.6. SWS superstructure for HEN (Yee and Grossmann, 1990)

Improvements on the SWS model has been carried out by Huang et al. (2012) who considered non-isothermal mixing in order to obtain superior HEN design. In their model, crossflow between the substreams within a stage are allowed, unlike in original SWS model by Yee and Grossmann (1990), where it is assumed at most one heat exchanger for each substream in a stage. They compared other models against theirs using literature examples and results obtain suggest that their proposed model is superior in determining the optimum total annualised cost (TAC) value, however with the expense of higher computational time for the model to converge. A latter work by Huang and Karimi (2014) expanded the SWS model to allows cross-flow, cycling matching, matches in series on sub-streams at each stage, and allowance for multiple utilities to be used at any stage within the superstructure. Due to the complexity of the model, the solution takes an ample of time for convergence and almost always requires an initiation point for some of its variables to solve (Huang and Karimi, 2014).

2.4.3 Automated targeting model (ATM)

ATM is based on pinch analysis technique but is carried out mathematically thus, avoiding the tedious iteration process and guarantees accurate answer unlike the graphical approach. It was originally established for the synthesis of mass exchange network by El-Halwagi and Manousiouthakis (1990). ATM is conceptually similar to the problem table approach first presented by Linnhoff and Flower (1978), where the inlet and outlet conditions are used to set up a cascade of intervals (e.g temperatures intervals for heat integration). The surplus of properties (e.g. heat) is allowed to cascade down to the next intervals. The biggest deficit found at an interval represents the minimum amount of fresh utilities required to be supplied to meet the problem specifications. The technique was later extended for material resource conservation network for in-plant reuse/recycle (Ng et al., 2007a) and regeneration (Ng et al., 2007b). Lee et al. (2009) further extended the use of ATM model for energy resources planning, while Ooi et al. (2014) applied ATM for power generation sector. Beside its successful application in continuous processes, ATM has also been extended by Foo (2010) for batch HEN and water network synthesis. Goh et. al. (2016) have recently use the ATM approach for HEN synthesis with utility system. In their method, they fix the $\Delta T_{\min}$ for process to process heat integration and target the utilities required such as boiler fuel, cooling water and chilled water that produces the

lowest operating cost. Ng et. al. (2017) presented *steam cascade analysis* (SCA) on an ATM framework for cogeneration potential targeting and steam allocation for steam distribution network. The objective of the model was to find the optimum flowrate of steams with the highest pressure that meets the demand of the users while considering the potential cogeneration of power between different processing sites. Both works however, did not consider the trade-off between HEN area and utility cost.

2.4.4 Total site heat integration

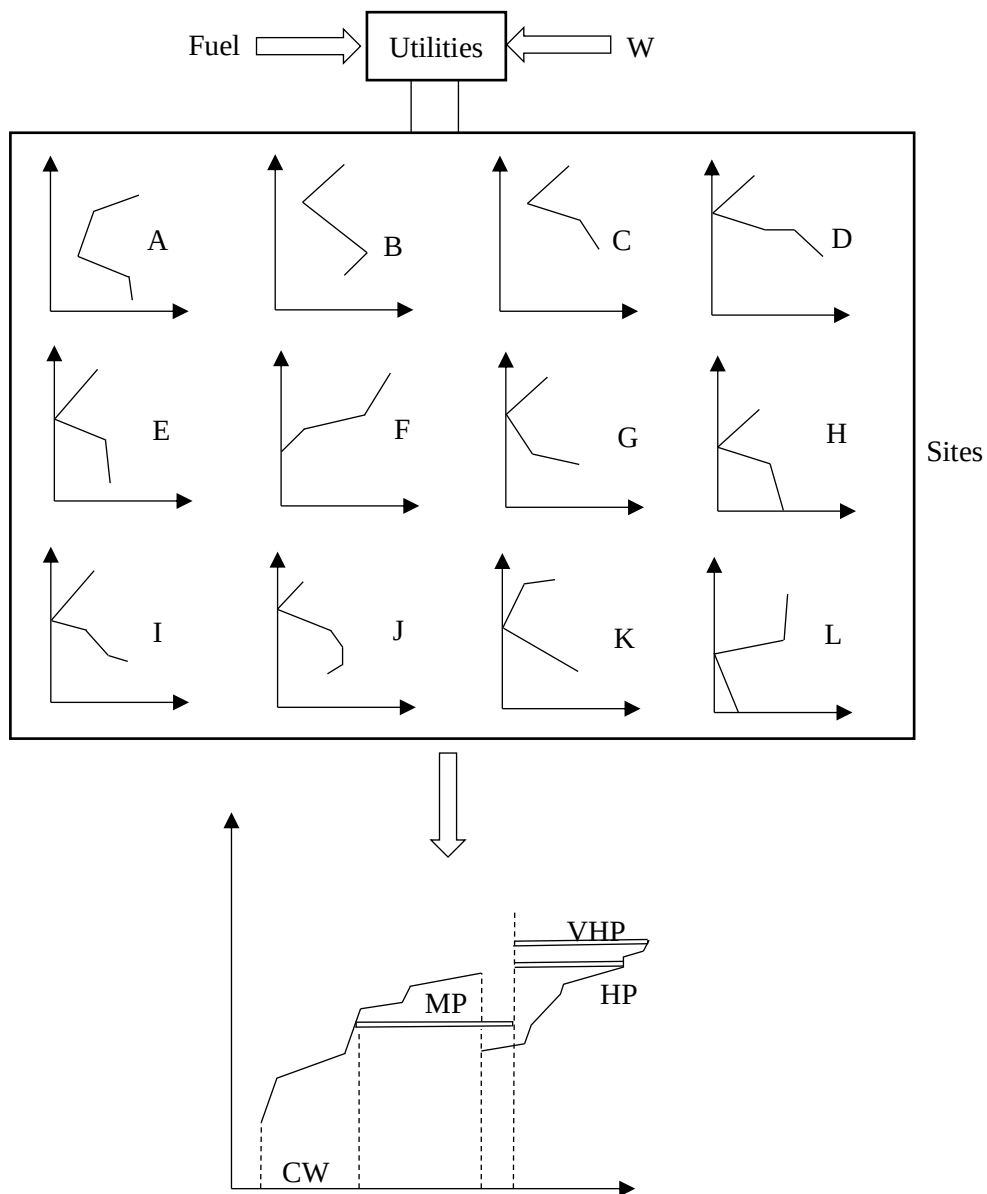


Figure 2.7. Dhole and Linnhoff (1993) TSHI of multiple sites with utility streams

Apart from the conventional HEN problems that deal with process-to-process heat recovery, another approach that is widely studied is *total site heat integration* (TSHI). The latter was first pinned down by Dhole and Linnhoff (1993) whom describes TSHI as *heat integration that incorporates several processes, serviced by and linked through a central utility system* (Figure 2.7). TSHI allows heat integration between several individual processes to recover heat indirectly via a common utility system, which offers additional inter-process heat recovery through consumption and generation of utilities. Research work in this area has been largely based on modification on the conventional process-to-process heat recovery problems that were discussed earlier.

2.4.4.1 TSHI based on graphical approach

Klemes et al. (1997) extended the initial TSHI work on different processes in a plant by considering multiple plants in close vicinity that can be integrated together. They proposed site composite curve which combines the heat demands and sinks of multiple plants in a single composite curve where the overall energy targets (isothermal utilities) for the plants can be carried out simultaneously. According to Varbanov et al. (2012) a single uniform $\Delta T_{\min}$ integrated in a Total Site does not guarantee optimality. In order to overcome this limitation, they propose a sequential graphical approach that takes into account the variation of $\Delta T_{\min}$ according to different processes in chemical plants. Their obtained results however were mixed. At the process level, only a unit (out of four) showed reduction in cooling utilities usage, while both cold and hot utilities usage increased by 30% and 18% respectively at site level, in comparison with the result when average $\Delta T_{\min}$ value is used. Tarighaleslami et al. proposed (2017) a unified TSHI that allows both isothermal and non-isothermal fluids to be targeted together for multiple plants. In their later work (2018), they extended their unified TSHI approach by including the synthesis of heat exchanger network which uses combination of isothermal and non-isothermal fluids. Ramananiah and Khanam (2018) proposed a modified TSHI which caters for an industry specific problem, i.e. sponge iron cluster.

These works on TSHI and *heat recovery loop* (HRL) focuses mainly on the targeting of minimum utility. Their solutions do not guarantee the optimum cost of the entire system as HEN area was not included in the analysis. To overcome this limitation, Boldyryev et al. (2014) proposed a

methodology to calculate the HEN area for Total Site based on the modification of Ahmad and Linnhoff (1990) model for isothermal fluid. They varied the temperatures of the utilities at each enthalpy interval and calculated the corresponding HEN area; doing so enables the identification of utility temperatures that lead to minimum network area. For the design of HRL, Walmsey et al. (2013). proposed a methodology using $\Delta T_{\min}$ approach to calculate the heat exchanger areas. The result from their proposed method was then compared to three programming optimisation approaches. Based on the comparison, they concluded that the $\Delta T_{\min}$ approach is effective. In their work however, the focus was solely on the heat exchanger area and not the overall cost of the system. In their subsequent work, Walmsley et al. (2015) considered a combined inter-plant heat recovery via intermediate fluid and solar systems in a dairy plant using the $\Delta T_{\min}$ approach. These combined systems improve solar utilisation while using the same intermediate fluid infrastructure such as piping, pumping and storage. In their model however, only parts of the system (i.e. the utilities and the solar collector area) were considered, but not the HEN.

2.4.4.2 TSHI based on mathematical approach

Apart from the above-mentioned graphical approaches, mathematical model has also been studied for the application of TSHI. These mathematical approaches largely based on Yee and Grossmann (1990) SWS. Laukkanen et al. (2012) used the modified SWS model to achieve simultaneous synthesis of HEN, where heat transfer can take place between processes directly, or indirectly through intermediate fluid. In their model, the utilities are also allowed to transfer heat with the intermediate fluid which was held at a fixed temperature. The proposed model was robust however the authors noticed some illogical result from their model. This illogical result was with the case where no intermediate fluid was used, in which the result obtained was not the same with the case where process-to-process heat integration was carried out. Note that both cases become equivalent when no intermediate fluid is used. The authors attributed this irregularity to the non-convex nature of the problem. Hipolito-Valencia et al. (2013) also used the SWS model to develop a multi-objective optimisation problem for inter-plant trigeneration systems. The model however only

considers process-to-process heat recovery among plants, but integrates Steam Rankine cycle, Organic Rankine cycle and the absorption refrigeration cycle for inter-plants heat recovery scheme.

Nemet et al. (2015a) proposed two different mathematical approaches that solve utility and HEN area costs simultaneously. In their first approach, they proposed a sequential model where the process-to-process. Heat integration is carried out first at process level, prior to the total site level in order to synthesise a Total Site HEN. In the second approach, heat integration was carried out simultaneously at process and Total Site levels. Although the latter approach produced better results, the authors explained that the simultaneous model might be too complex to solve, particularly for large-scale problems; and hence the sequential approach are more useful in these cases. In their later work (Nemet et al., 2015b), they made use of the SWS model for TSHI by indirect means (intermediate fluids), while taking into account heat losses, pressure drops and the piping distance between processes in their costing model. Chang et al. (2016) proposed an inter-plant heat integration using HRL in which one plant acts as a sink and the other as a source. Their model only considered heat integration via intermediate fluid without any process-to-process heat integration. In their subsequent work, Chang et al. (2017a) proposed an expanded superstructure model that takes into account energy, heat exchanger, piping and pumping costs. They considered both direct heat integration between process-to-process streams (in-plant) and indirect means between intermediate fluid and process streams (inter-plant). In their later work, Chang et al. (2017b) used SWS model to consider inter-plant and in-plant heat integration using process stream without an intermediate fluid. In order to accommodate process integration between different plants, they proposed adding intervals to the SWS model, that is equivalent to the number of plants involved in inter-plant heat integration. These intervals allow every process stream to be transported into any plant through the pipeline networks between the plants. Hong et al. (2019) also studied the same problem but using a new transshipment model they proposed earlier (Hong et al. 2017). Their result showed that the new transshipment model was able to produce much lower TAC than that by Chang et al. (2017b). The superior result obtained by Hong et al. (2019) is attributed to new transshipment model that is able to consider multiple stream splitting, stream by-passing, isothermal and non-isothermal mixing, and

recycling flows, which was not considered in the model by Hong et al. (2017). Note however that all of these works assume constant utility temperatures and ignored the cost of utility regeneration.

2.4.5 Utility system

Apart from TSHI, some research works exclusively on cooling and heating utility systems optimisation have also been reported. These include the work on cooling water system by Kim and Smith (2001), chilled water system by Foo et al. (2014) and hot oil system by Ataei et al. (2014). Kim and Smith (2001) and Ataei et al. (2014) works were based on graphical approach, while Foo et al. (2014) proposed two approaches i.e. graphical and algebraic, either of them can be used to determine minimum cooling and chilled water required by the system. These works however did not consider the trade-off between capital cost of HEN and the utility cost, but instead focused on minimum utility targeting. Bade and Bandyopadhyay (2014) proposed an algebraic approach that is iterative in nature to find the optimum solution which includes the cost of heat exchangers area for thermal fluid system. Their approach however requires detailed design of the HEN in order to determine the overall cost.

2.4.6 Flowsheet optimisation

From the above review, one should note that all the above-mentioned works considered only HEN exclusively without consideration of flowsheet optimisation. The result obtained from such approach may not produce an overall optimum solution, since only part of the problem has been addressed. In view of this, Duran and Grossmann (1986) proposed a simultaneous approach for heat integration and process flowsheet optimisation. In their method, a smooth approximation is used to estimate non-differentiable that was caused by the exclusion of temperature intervals for heat integration. Note that the approach of Duran and Grossmann (1986) has been used by other researchers for simultaneous optimisation of other systems. For instance, Srinivas and El-Halwagi (1994) applied this approach for simultaneous optimisation of mass and HEN. In their work, a sequential approach was adopted where the design targets were first established ahead of actual network design. Baliban et al. (2012) used the smooth approximation approach for simultaneous optimisation of process synthesis along with heat and water integration. Improvements on the smooth approximation approach (Duran and Grossmann, 1986) was carried out by Dowling and Biegler

(2015), where they extended the model by considering multiple heat exchanger zones, which allows process units of different groups to undergo heat integration separately. Grossmann et al. (1998) on the other hand, proposed a disjunctive approach that overcomes difficulties in handling isothermal streams in the model of Duran and Grossmann (1986). Their model is based on tight MINLP formulation with big-M constraints, which can be deduced to MILP problem when dealing with isothermal streams. Navarro-Amoros et al. (2013) also applied disjunctive approach, however their work made use of the transshipment model by Papoulias and Grossmann (1983), which allows temperature intervals to be used for heat integration.

All the above-discussed works discussed for simultaneous optimisation of HEN and process flowsheet involves only the determination of the minimum utilities without consideration the HEN area. Recently, Kong et al. (2017) proposed a simultaneous approach that includes HEN area as part of the variable in heat integration and process optimisation. In their work, they used dynamic enthalpy grid onto which the variable process stream temperatures and enthalpies are mapped (Kong et al., 2017). They choose to substitute the conventional approach in calculating the log mean temperature with Paterson (1984) approximation in order to avoid numerical difficulties in their model (Kong et al., 2017). Note however that this approximation may lead to error close to 5%, as reported by Hong et al. (2017) based on some test cases. Thus, with the use of the Paterson approximation, the error of the targeted values can increase further, since the work of Kong et al. (2017) is based on the model of Linnhoff and Ahmad (1990) where the inaccuracy is up to 10% (for cases with film heat transfer coefficient within an order of magnitude).

2.5 Research gaps

Based on the literature review carried out in the preceding sections, the research gaps are identified as follow:

1. The work on utility system are limited. The main research works on heat integration focus on process-to-process heat recovery, with the remaining heating or cooling duties supplied by hot or cold utility systems. In this conventional heat integration problem, heating utility system always uses parallel design as shown in Figure 1.1.
2. The works on TSHI consider the use of intermediate fluid such as water or oil to transfer heat indirectly. However, these works focus mostly on the heat integration between multiple process plants. Most works considered fixed inlet and outlet temperatures for the utility streams. Besides, there was no cost consideration for reheating the utility fluid to supply temperature before recirculation.
3. works on utility system are largely on the targeting the minimum flowrate for the heating or cooling medium without simultaneous optimisation of heat exchanger area. The only work that considered the trade-off between the area of heat exchanger and utility costs involves the design of the HEN before its cost was determined (Bade and Bandyopadhyay, 2014) which is not effective if large number of iterations are needed.
4. ATM application on HEN problems are limited. Published works focused mostly on utility targeting (such as steam), without consideration on the cost of heat exchanger area in the optimisation function. ATM has good potential since it provides pinch-based insights while providing accuracy in calculation and able to include other constraints in the model.
5. Research study on flowsheet optimisation between HEN and phase separation processes is limited. In the past, works on simultaneous optimisation of heat integration were reported on mass exchange network (Srinivas and El-Halwagi, 1994), water and power (Baliban et al., 2012), and chemical reactors (Duran and Grossmann, 1986 and Kong et al., 2017). The work on separation processes by Navarro-Amoros et al. (2013) and Dowling and Biegler (2015)

focused on distillation units, which differs from phase separation units considered in this work.

6. Models that considered temperature and enthalpy load as variables at heat exchanger only involves minimum utility targeting. One work that considered heat exchanger area as part of the optimisation function uses Patterson approximation to calculate the log mean temperature (Kong et al., 2017). This approach however may lead to larger error, since the “supertargeting” model that is used for utility and heat exchanger area targeting has built-in inaccuracy, since the HEN area is approximated using the composite curves.
7. The overall heat transfer coefficients used in calculating the heat exchangers and WHRU area are determined based on assumed values of film heat transfer coefficient for hot and cold streams. The sensitivity of choosing these values are not normally assessed and this may lead to large discrepancies between the result obtained from heat integration when actual sizing of heat exchangers is carried out.
8. In conventional heat integration problem, the heat exchangers and WHRU design are not optimised. As discussed in literature review, there are number of design options that can be considered for heat exchangers, ranging from outside diameter of the tubes, flow arrangement of the tubes, number of tube passages, etc. that can impact the cost of heat exchanger. Simultaneous optimisation of heat exchangers with heat integration on heating utility system may lead to lower cost of the system than just applying heat integration alone.
9. Inclusion of other system such as TEG, steam bottoming cycle or organic Rankine cycle to the heating utility system has not been studied in term of actual cost benefit. Structural cost will also need to be included in the overall cost model in order to realise the actual benefit, if any.

CHAPTER 3

SCOPES AND METHODOLOGY OF RESEARCH

3.1. Research Scopes

Items one to six identified as research gaps in Chapter 2 forms the basis for the scopes of this thesis. The first four research gaps were important to be addressed as it focusses on creating a specific heat integration model for heating utility system; the latter can be expanded to fulfil the other research gaps. On the other hand, items five and six of the research gaps were prioritised instead of the rest because it connects the study of heating utility system to another process system in oil and gas platform. With consideration of multiple systems interaction, a bigger cost benefit was hoped to be achieved. Thus, the main scopes of this research to address the selected research gaps are summarised as below:

1. Graphical-based approach for heating utility system that considers the trade-off between utility and capital cost of equipment.

Process heat integration using the “supertargeting” framework of Linnhoff and Ahmad (1990) is extended to determine an alternative design to the conventional parallel design of the heating utility system. With reuse of the heating medium, the cost of the overall system is expected to reduce. This is examined by considering the overall cost of the system which includes heat exchanger, waste heat recovery unit (WHRU), pump and utility. Besides, the impact of outlet temperature of the heating utility medium on the overall cost of the system is to be analysed while obtaining the optimum cost design.

2. Automated-based approach for targeting heating utility system.

Additional variables i.e. inlet temperature, piping and fitting losses, and piping cost are considered in order to obtain design improvement from the graphical approach examined in the first scope. The automated targeting model (ATM) model that is adopted for this study allows more constraints to be included which in return capable on delivering a better result while representing the heating utility system more accurately. Besides, it avoids the tedious

iteration process of minimum approach temperature ($\Delta T_{\min}$) used in graphical approach in determining the optimum heating utility system design.

3. A systematic methodology for simultaneous optimisation of heating utility and phase separation systems.

In previous two scopes, only the heating utility system was subjected to design optimisation. Optimising a system independently without consideration on how its design changes can impact other system may lead to suboptimal result. With this view, both heating utility and phase separation systems are solved simultaneously to obtain the lowest overall total annualised cost (TAC). The phase separation system is closely tied to the heating utility system since heating is provided to reduce the size of the separators. The separators are sized for minimum cost independently for different heaters outlet temperatures and the result are examined. The costs of separators are then used for simultaneous optimisation with the heating utility system.

3.2 Research Methodology

The proposed research scopes are explored using heat integration methods. First, an extensive literature review is carried out on all research works that were established for heat exchanger network (HEN). Next, the two systems i.e. heating utility and phase separation systems are studied and the application of existing heat integration methods are examined. From the gathered information, new model for targeting heating utility system is developed. The cost of conventional parallel design of heating utility system is used as the benchmark to judge the performance of the new model. The model is then expanded further with the inclusion of phase separation system as part of the overall optimisation problem. More details on the methodology based on the scope identified earlier is provided below and Figure 3.1.

1. Graphical based approach for heating utility system that considers the trade-off between utility and capital cost of equipment.

A pinch-based graphical approach is proposed to solve the heating utility system for minimum TAC. The “supertargeting” concept that is used in the graphical approach allows

the targeting of HEN area to be carried prior to the detailed design of HEN. The model is solved by first calculating the TAC for different heat recovery minimum approach temperatures that are prechosen. Then, the optimum minimum approach temperature is determined by the lowest TAC among all calculated values. This identified optimum minimum approach temperature value is then used to design the HEN using the revised nearest neighbour algorithm (NNA).

2. ATM-based approach for targeting heating utility system.

ATM framework is used to formulate the graphical approach mathematically in order to avoid the iteration of minimum approach temperature and making it as a continuous variable. Since the ATM model considers both the inlet and outlet temperatures of the heating medium as variables, the mathematical computation becomes more complex. In order to overcome this, a new approach is shown to calculate the minimum and maximum heating medium flowrates to ease the computational difficulty. To closely resembles the offshore facility, the process piping is sized according to offshore specification in the model.

3. A systematic methodology for simultaneous optimisation of heating utility and phase separation systems.

Due to linear arrangement of the crude oil heater (Figure 1.2), solving simultaneous optimisation model for heating utility and phase separation systems becomes very complex. The complexity arises because any changes in the upstream heaters (i.e. light oil and heavy oil heaters) effects the inlet condition of the oil into crude oil heater. Each change in the oil condition requires a new set of chemical properties data to be generated from the simulation software. Besides, the number of possible temperatures combination between all heaters also increases the complexity significantly. In order to overcome these problems, a systematic methodology is proposed to reduce the number of computations needed to determine the optimum solution. Hence, a new novel algorithm is proposed to the ATM in order to sort its temperature intervals automatically. This is required since heating medium user outlet temperatures and enthalpy load is no longer fixed parameters but have a become discrete variables. For the separators, for each different outlet temperature, an optimum size that

produces the lowest cost is determined. This information is then paired with the corresponding outlet temperatures of the heaters when determining the optimum solution for both systems. The model is solved for different combination of outlet temperatures and the optimum solution is determined from the lowest TAC value.

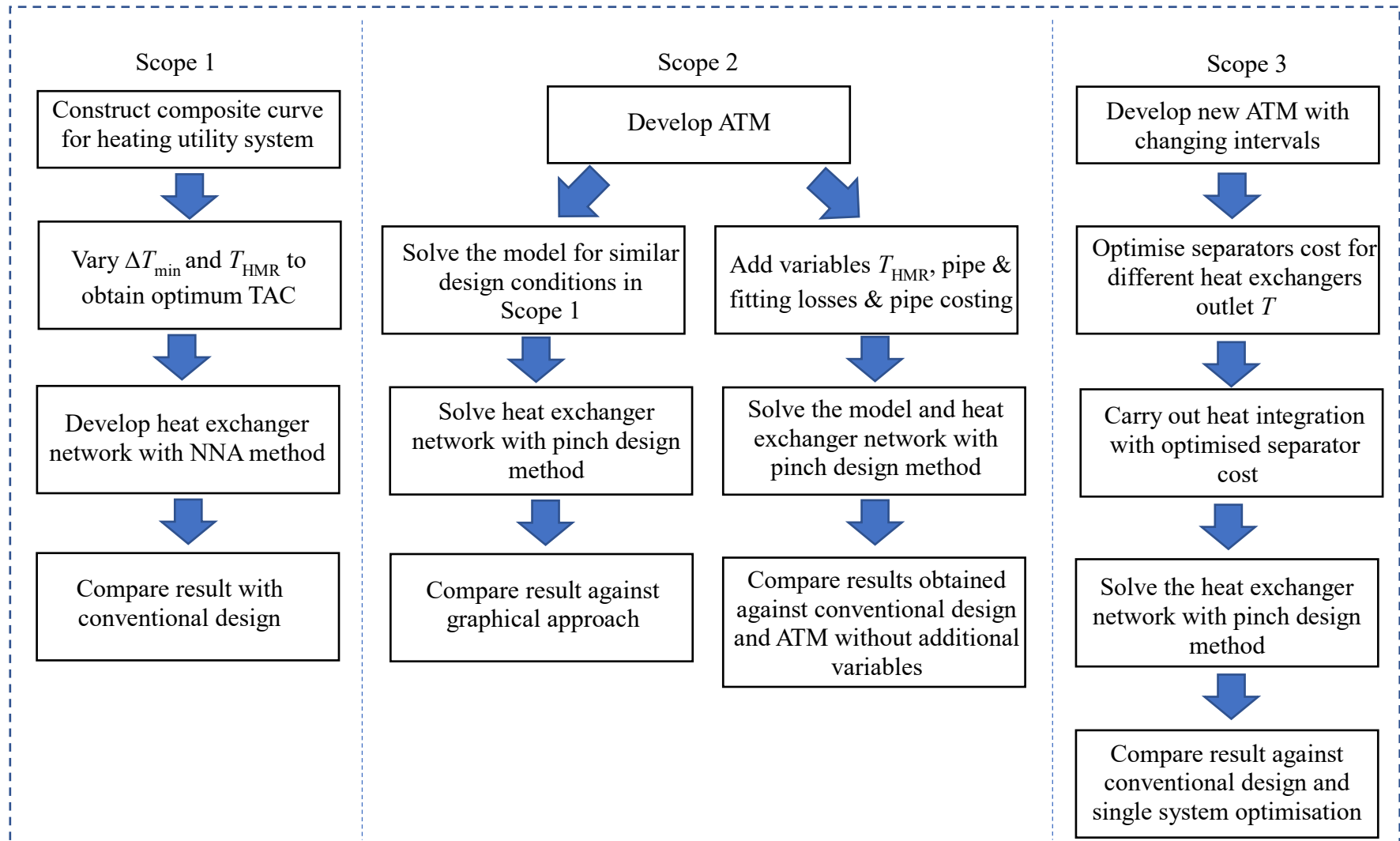


Figure 3.1. Research methodology

CHAPTER 4

GRAPHICAL APPROACH FOR TARGETING AND DESIGN OF HEATING UTILITY SYSTEM

4.1 Introduction

In the following sections, the formal problem statement for the heating utility system for a normal offshore platform is introduced. Next, the overall methodology for targeting and design for the heating utility system is outlined. This includes the revised version of the conventional heat recovery pinch diagram (HRPD, Linnhoff et al., 1982), which is used for targeting the minimum heating medium flowrate and HEN area (i.e. “Supertargeting”). Note that the conventional heat recovery pinch diagram (HRPD, Linnhoff et al., 1982) is not directly applicable in this case due to the absent of process hot streams where energy may be recovered (through process-to-process heat exchanges). A heating medium supply line is instead utilised to determine the minimum flowrate of heating medium. Next, the revised enhanced nearest neighbour algorithm (ENNA) procedure is proposed for the design of the heating utility system. A literature example from Bade and Bandyopadhyay (2014) and an industrial case study were used to illustrate the proposed procedure.

4.2 Heating utility system and problem statement

The formal problem statement for heating utility system design can be stated as follows. Given is a set of heating medium user $m \in M$, each with a fixed heat load Q_m to be provided by the heating medium (water or hot oil), and with specified maximum process inlet ($T_{in, m}$) and outlet ($T_{out, m}$) temperatures. Waste heat recovery unit (WHRU) reheats the heating medium at the return temperature (T_{HMR}) using flue gas from gas turbine generator and/or gas turbine compressor at specified maximum inlet ($T_{in, g}$) and outlet ($T_{out, g}$) temperatures. The reheated heating medium at circulation temperature (T_{HM}) is to be re-circulated among the heating medium users (heat exchangers) to achieve the minimum use of heating medium. The main objective is to synthesise an optimum heating medium network, where heat load requirements of all heating medium users are to be fulfilled. The optimum heating medium network may feature the minimum heating medium flowrate for a given ΔT_{min} , or minimum total cost in which the optimum ΔT_{min} is to be identified.

4.3 Overall targeting procedure

Figure 4.1 shows the overall procedure for targeting and design of the heating utility system. For a specified minimum approach temperature ($\Delta T_{\min}$), the minimum flowrate of the heating medium will be determined, followed by the minimum heat transfer area of the heat exchanger network (HEN), WHRU and pump power. To perform targeting for the heating medium, a revised HRPD is needed. The calculated flowrate, area targets and pump power are then converted into cost targets. Once the optimum $\Delta T_{\min}$ is determined (i.e. which leads to minimum total annualised cost (TAC)), the heating utility system is designed. The latter task is completed using the revised ENNA procedure. Both the revised HRPD and nearest neighbour algorithm (NNA) are presented in the following subsections.

4.3.1. Graphical technique for heating medium and HEN area targeting

In this section, the conventional HRPD (Linnhoff et al., 1982) is revised as the heating medium composite curve (HMCC) to determine the minimum flowrate of heating medium (hot oil or water) and for targeting the HEN area for a given $\Delta T_{\min}$. As mentioned earlier, the conventional HRPD (Linnhoff et al., 1982) cannot be utilised directly in this case, due to the absent of process hot streams. Hence, a supply line is used to determine the flowrate of the heating medium.

The steps for constructing a HMCC are given as follow:

1. Plot the limiting profiles of all heating medium users on a temperature versus enthalpy diagram, according to their maximum inlet and outlet temperatures. Within each temperature interval, the individual segments of the heating medium users are merged by adding their heat loads. This forms the HMCC (see Figure 4.2). This is conceptually similar to the cold composite curve in the conventional HRPD (Linnhoff et al., 1982).

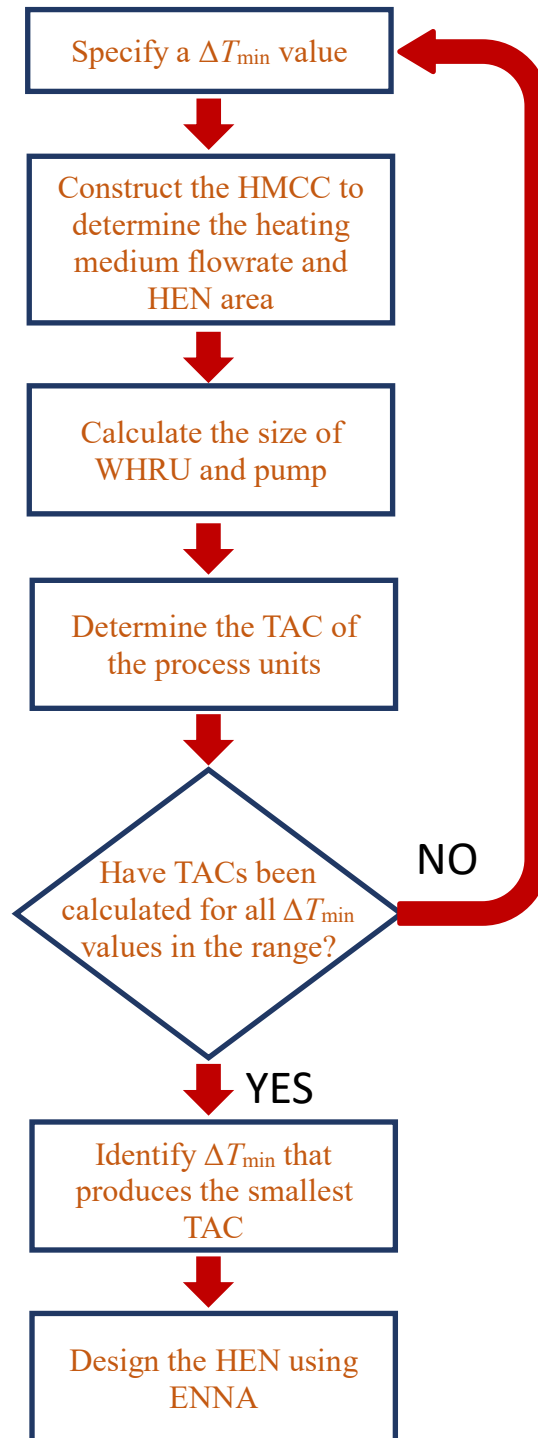


Figure 4.1. Overall procedure for targeting and design of heating utility system

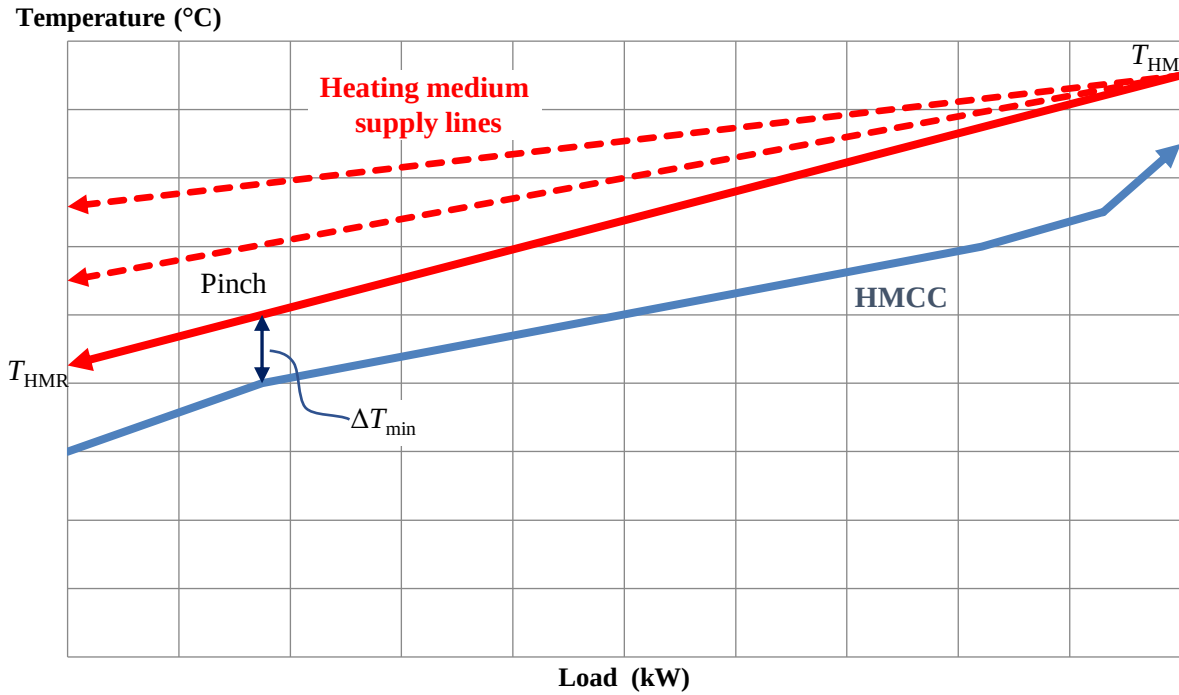


Figure 4.2. The HMCC and heating medium supply lines for targeting

2. The heating medium supply line is then drawn from its supply temperature to provide the heat duty for the entire system. The supply line is required to be segregated from the HMCC with the specified $\Delta T_{\min}$. Note that, the point at which the heating medium supply line is separated from HMCC by the specified $\Delta T_{\min}$ is known as the pinch of the heating utility system. The flowrate of the heating medium is then determined by the inverse slope of the supply line. This step is conceptually similar to flowrate targeting in previous works on cooling water minimisation (Kim and Smith, 2001) and chilled water network (Foo, 2009).
3. To determine the HEN area target (i.e. “Supertargeting”), the HMCC is partitioned vertically at each of its turning points to form enthalpy interval; the latter represents the separate regions of heat transfer (see Figure 4.3). Assuming that vertical heat transfer occurs for each region, the heat transfer area for the HEN (A_{HEN}) is calculated using individual stream film heat transfer coefficient, given by Equation (4.1) (Smith, 2005):

$$A_{\text{HEN}} = \sum_t^{\text{Intervals } T} \frac{1}{\Delta T_{\text{LGM}t}} \left[\sum_i^{\text{Hot streams } I} \frac{q_{i,t}}{h_i} + \sum_j^{\text{Cold streams } J} \frac{q_{j,t}}{h_j} \right] \quad (4.1)$$

where $q_{i,t}$ = stream duty on hot stream i in enthalpy interval t

$q_{j,t}$ = stream duty on cold stream j in enthalpy interval t

h_i = film transfer coefficients for hot streams i / inner heat transfer coefficient (including wall and fouling resistances)

h_j = film transfer coefficients for cold stream j (including wall and fouling resistances)

I = total number of hot streams in enthalpy interval t

J = total number of cold streams in enthalpy interval t

T = total number of enthalpy intervals

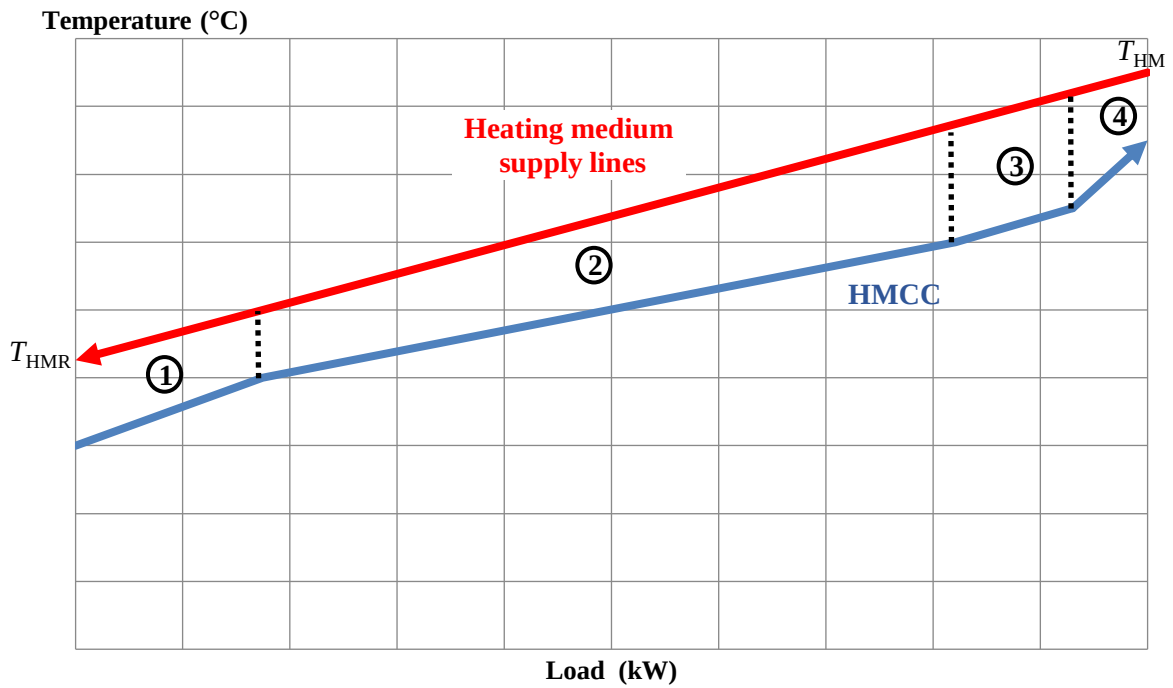


Figure 4.3. Area targeting with HMCC

4.3.2. HEN design with revised NNA

In this section, the NNA originally proposed by Prakash and Shenoy (2005) for resource conservation networks (RCN) design, as well as ENNA by Shenoy and Shenoy (2013) for cooling water network design is modified here for the synthesis of the heating utility network to achieve the various network targets. The revised NNA excludes the step in the ENNA which prioritises the local recycle ahead of a source that is just lower temperature than the sink. However, the removal of local recycle wherever it exists after all the sources and sinks are paired according to the nearest neighbour is still maintained. This removal of local recycle does not exist in NNA (2005), one can also view the revised NNA as an inclusion of this addition step to NNA.

Note however that the data of the heating medium users that is used to plot the HMCC actually corresponds to the *cold streams* (streams to be heated) of the heat exchangers. Data for the *hot streams* (the heating medium itself) are to be identified prior to the use of the revised NNA. This is done by incorporating the minimum approach temperature with the assumption of having the heat exchangers in counter-current configuration (see details in Appendix A).

After the identification of the data of the heating medium in all heat exchangers, these heating medium users are characterised as a pair of sink (where heating medium is utilised) and source (where heating medium is produced). In other words, sink j corresponds to the inlet of the heating medium of user m , with flowrate requirement of F_{SKj} and temperature requirement of T_{SKj} . Conversely, the outlet of heating medium of user m corresponds to source i , with flowrate of F_{SRi} and temperature of T_{SRi} . The steps for using the revised NNA are given as follows.

1. All sources and sinks are arranged in descending order of temperature levels in a *matching matrix*, in column and row. The pinch temperature identified in the targeting stage divides the network into regions of higher (with temperature higher than the pinch) and lower quality (with temperature lower than the pinch). Sources are only permitted to be allocated to sinks in the same region. In other words, allocation of sources and sinks in different regions are prohibited. Cross-pinch regions in the matching matrix (where the allocation of sources to sinks is prohibited) are greyed out. Note that source at the pinch temperature it is an

exceptional case, as it belongs to both the lower and upper pinch region and shall not be greyed out. Start the design from the sink of highest temperature.

2. Sources are allocated to sinks of the same temperature, if found. In the scenario where the source flowrate is less than that required by the sink, the entire source is allocated, and the remaining flowrate is made up from a pair of neighbouring sources of the sink (refer Step 3).
3. Two candidate sources SR_i and SR_{i+1} are identified to satisfy the heat capacity flowrate and enthalpy constraints of sink SK_j . Note that SR_i and SR_{i+1} are the nearest available “neighbours” to sink SK_j , with temperature levels just higher and just lower than that of the sink, i.e. $T_{SR_i} > T_{SK_j} > T_{SR_{i+1}}$. The allocated flowrates of these sources are determined using Eqs. (4.2) and (4.3).

$$F_{SR_i, SK_j} + F_{SR_{i+1}, SK_j} = F_{SK_j} \quad (4.2)$$

$$F_{SR_i, SK_j} T_{SR_i} + F_{SR_{i+1}, SK_j} T_{SR_{i+1}} = F_{SK_j} T_{SK_j} \quad (4.3)$$

where F_{SR_i, SK_j} is the allocated capacity flowrate from SR_i to SK_j .

There exist two different scenarios for this step. In the scenario where one of the source candidate has insufficient flowrate, $F_{SR_i} < F_{SR_i, SK_j}$ or $F_{SR_{i+1}} < F_{SR_{i+1}, SK_j}$, the source candidate (of insufficient flowrate) is completely utilised for the sink. A new candidate source is to be identified to replace the insufficient source (with just lower or just higher temperatures than the sink). On the other hand, for cases where both the candidate sources (SR_i and SR_{i+1}) have insufficient flowrates to fulfill the requirement, i.e. $F_{SR_i} < F_{SR_i, SK_j}$ and $F_{SR_{i+1}} < F_{SR_{i+1}, SK_j}$, both the higher and lower temperature source is utilised fully for the sink. Then, a new pair of candidate sources (with just lower and just higher temperatures than the sink) is next identified. Note that for both scenarios, Equation (4.2) and (4.3) needs to be updated with the new flowrate and enthalpy requirements. This is done by deducting the flowrate and enthalpy provided by the earlier source to the sink.

4. Proceed to Step 2 for the next sink of lower temperature. Once all sinks are completed, proceed to Step 5.

5. All *local recycle* (LR) matches between the source and sink that belong to the same heating medium user are identified and eliminated to decrease the flowrates that pass through them. Flowrates and temperatures of the LR's sinks and sources are adjusted accordingly. Doing this leads to reduced size of heating medium users (Shenoy and Shenoy, 2013).

The revised NNA procedure is summarised in a flow diagram given in Figure 4.4. Note that the newly added step of the revised NNA is marked with a thicker border line in Figure 4.4. The detailed step in using the procedure will be illustrated with the industrial case study.

In the following sections, a literature example from Bade and Bandyopadhyay (2014) and an industrial case study are used to illustrate the HMCC procedure.

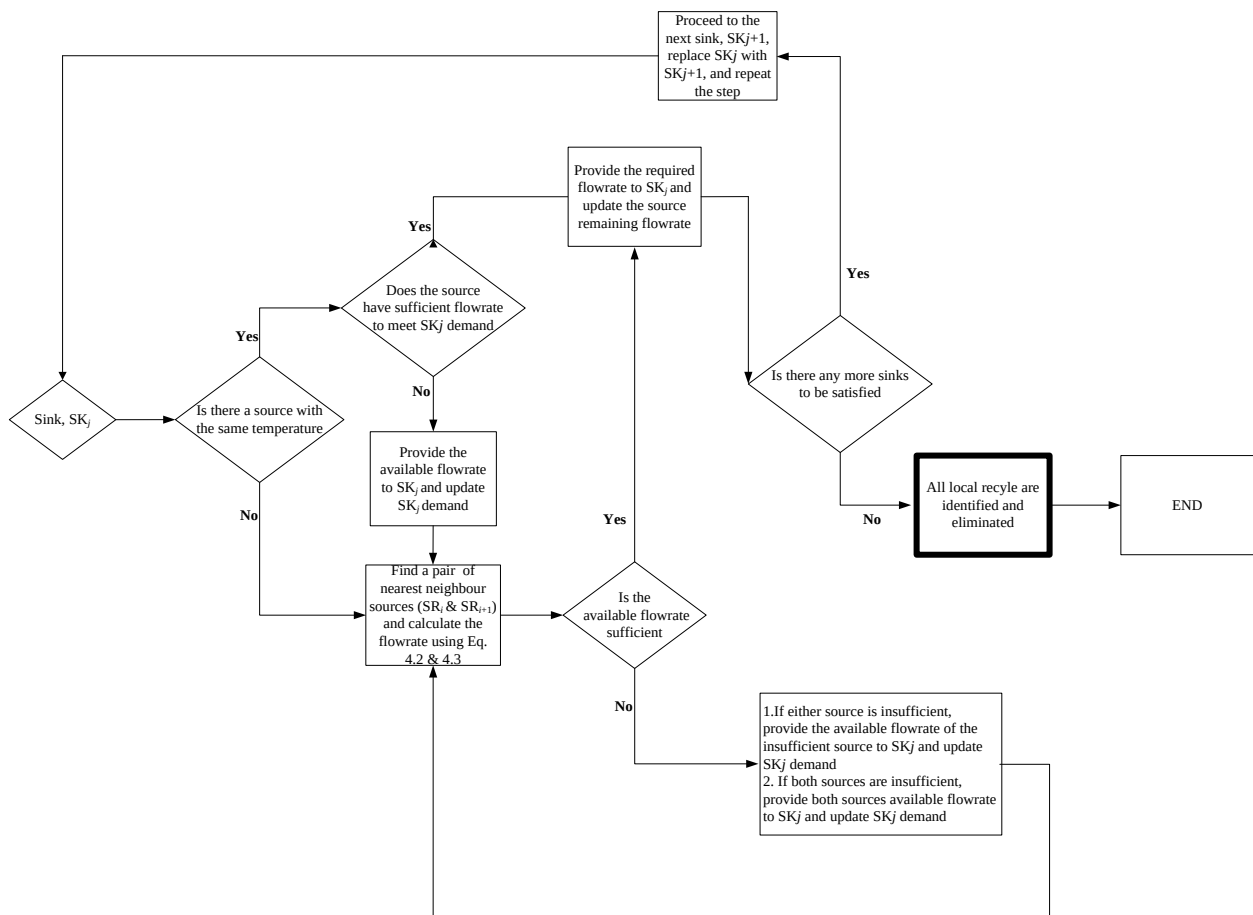


Figure 4.4. Flow diagram for revised NNA procedure (box with thicker border indicates newly added step)

4.4 Literature example

A literature example from Bade and Bandyopadhyay (2014) is used for illustration here. For this case, the heating medium of 370 °C is supplied to three reboilers in parallel to meet their heating requirements; with the data shown in Table 4.1. Even though this system is not used on the offshore platform, it closely resembles the heating utility system of the offshore platform. For HEN area targeting, the film heat transfers coefficients of the hot stream and cold streams, h_i and h_j are both taken as 0.8 kW/m² °C.

Table 4.1.

Process stream data for literature example (Bade and Bandyopadhyay, 2014)

Unit, M	Inlet ($T_{in, m}$) and outlet temperature ($T_{out, m}$) (°C)	Load, Q_m (kW)
1	200	1,800
2	175	1,600
3	150	2,400
Total		5,800

The HMCC is next plotted for ΔT_{min} ranging between 5 and 25 °C, in order to determine the optimum flowrate for heating utility and area targets for the HEN. Following steps 1-3 of the graphical targeting procedure, the HMCC is constructed in Figure 4.5 for ΔT_{min} of 10 °C. Note that since the heating medium is used for reboiler operations in this case, the inlet and outlet temperatures of the process stream remain constant. As such, the HMCC is represented by horizontal segments at the respective temperature intervals (see Figure 4.5). Note also that the horizontal distance of the HMCC represents the total heat requirements for the three heating medium users, i.e. 5800 kW.

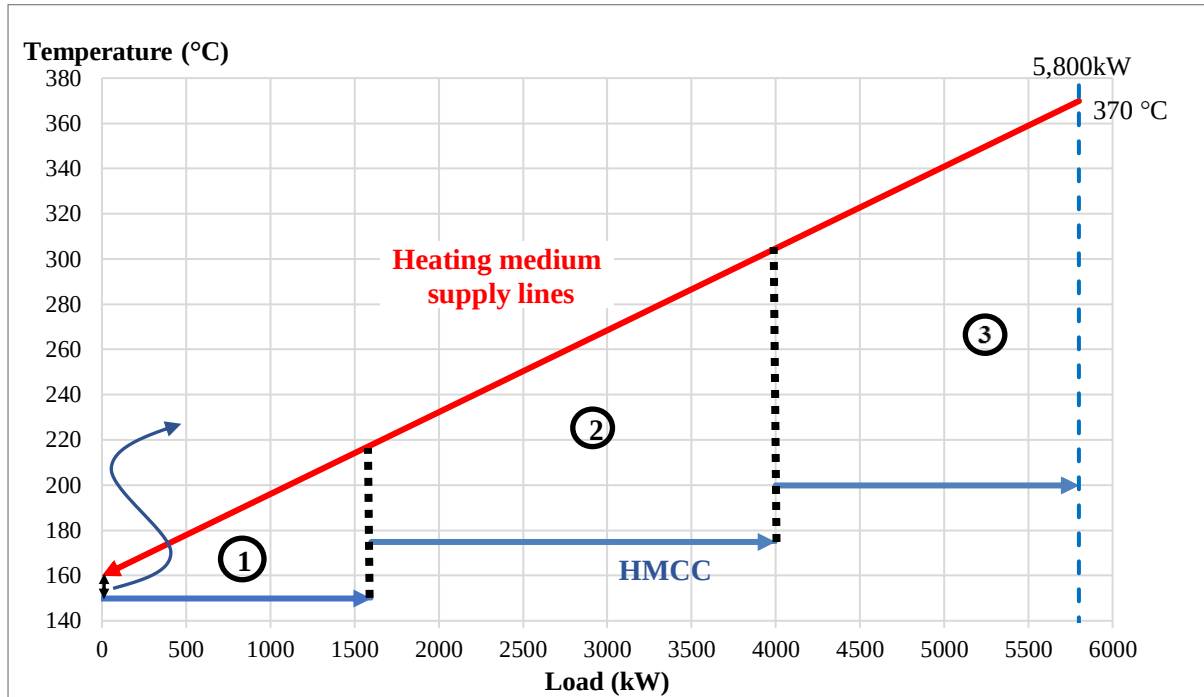


Figure 4.5. The HMCC for literature example ($\Delta T_{\min}=10\text{ }^{\circ}\text{C}$)

For $\Delta T_{\min}$ of $10\text{ }^{\circ}\text{C}$, the inverse slope of the supply line gives the heating medium capacity flowrate, which is $27.62\text{ kW}/^{\circ}\text{C}$ ($= 5,800\text{ kW}/(370 - 160\text{ }^{\circ}\text{C})$). Another important information from Figure 4.5 is the pinch temperatures. The latter can be viewed as the overall bottleneck of the problem, where the heating medium flowrate can be minimised. For this case study, the hot and cold pinch temperatures correspond to $170\text{ }^{\circ}\text{C}$ (heating medium) and $160\text{ }^{\circ}\text{C}$ (reboilers). Note that hot Pinch Temperature also corresponds to the heating medium return temperature (T_{HMR}), i.e. $160\text{ }^{\circ}\text{C}$ (see Figure 4.5). The total area of the HEN for the three regions in Figure 4.5 is determined as 242 m^2 using Equation (1). The steps are next repeated for $\Delta T_{\min}$ of 5, 15, 20 and $25\text{ }^{\circ}\text{C}$, with the result tabulated in Table 4.2.

Table 4.2.

Result for flowrate and HEN area targeting for literature example ($\Delta T_{\min}$ ranges between 5-25 °C)

$\Delta T_{\min}$	Heat capacity flowrate (kW/°C)	Heating medium return temperature T_{HMR} (°C)	HEN area from supertargeting (m ²)
5	27.0	155	285.6
10	27.6	160	242.1
15	28.3	165	217.1
20	29.0	170	199.6
25	29.7	175	186.2

The targeting results in Table 4.2 are comparable with those reported by Bade and Bandyopadhyay (2014). The heat capacity flowrate and heating medium temperature at various $\Delta T_{\min}$ values are found to be identical. Note however that in the original work by Bade and Bandyopadhyay (2014), the hot stream (i.e. heating medium) temperature is used. This means that additional steps are required in order to generate heating medium stream data from the process stream data, and this is to be conducted for all $\Delta T_{\min}$ values. Furthermore, the graphical method of Bade and Bandyopadhyay (2014) is based on the Grand Composite Curve (GCC). The GCC is useful for targeting the heating medium flowrate, but does not allow the area targeting (i.e. supertargeting) to be performed. In other words, the HEN is to be designed for each $\Delta T_{\min}$ value, before the trade-off analysis can be performed. This is certainly more cumbersome and tedious. Unlike the original work (Bade and Bandyopadhyay, 2014), the HMCC proposed in this work utilises process stream data directly, which allows area targeting to be carried out for every given $\Delta T_{\min}$ value, hence avoids the needs of detailed design of HEN. This also allows the trade-off analysis to be carried out at the targeting stage. The HEN for the literature example is shown in Figure 4.6.

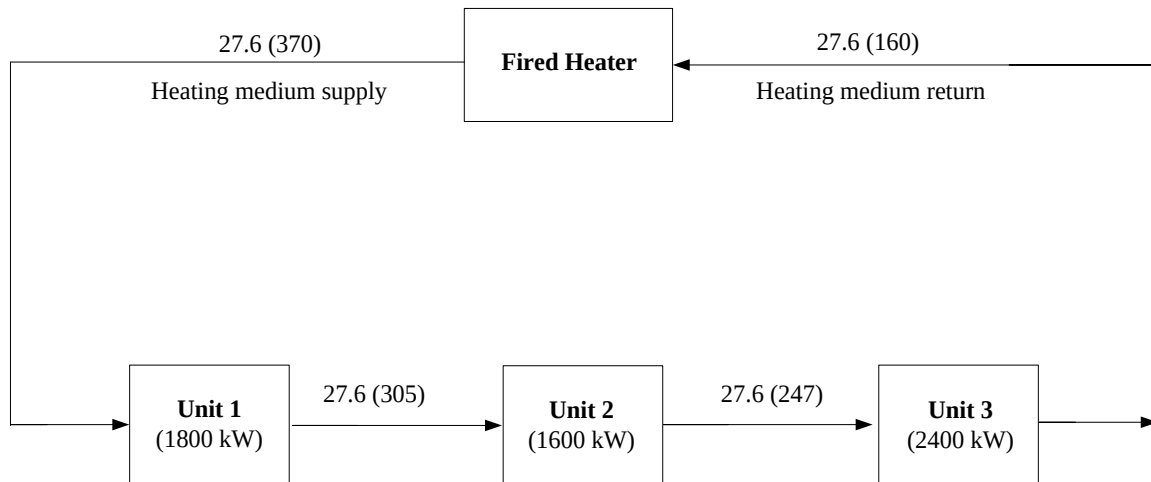


Figure 4.6. Network design for literature example (heating medium flowrate in kW/°C; temperature in °C, given in parenthesis)

4.5 Industrial case study

In this section, an industrial case study is illustrated to demonstrate the use of the revised graphical targeting technique to design a cost-optimum heating utility system. An offshore floating production storage off loading (FPSO) platform is to be installed at Angola (Africa). Its process flow diagram (PFD) is similar to that in Figure 1.1, with its data given in Table 4.3. In order to design a cost-optimum heating utility system, the capital cost of equipment (HEN, waste heat recovery unit and pump) and operating cost of the system (mainly the shaft power of pump) are considered. Note that the heating medium expansion tank and dump cooler are also excluded from cost optimisation, as the dump cooler is only used for abnormal operation and the cost for the heating medium expansion tank is assumed negligible. This FPSO uses water as heating medium, which is available at 130 °C (T_{HM} , 8.5 Barg). The “Supertargeting” approach is adopted here to synthesise a cost optimum heating utility system, with ΔT_{min} for the HEN to be determined as part of the problem and may vary between 10 and 50 °C. Note that apart from the HEN, the ΔT_{min} also affects the WHRU design. The latter is included in the analysis.

4.5.1. Flowrate and area targeting for HEN and WHRU

Taking $\Delta T_{\min} = 20^\circ\text{C}$ as an example, the steps outlined in the earlier section is used to construct the HMCC, as shown in Figure 4.7. The HMCC is partitioned at each of its turning point, resulted into 4 regions of heat transfer. The heat transfer area for the HEN (A_{HEN}) is then calculated using individual stream film heat transfer coefficient, given by Equation (4.1). The film heat transfers coefficient of the hot stream, h_i is taken as $2 \text{ kWm}^2 / ^\circ\text{C}$, while those for the cold streams are given in Table 4.3.

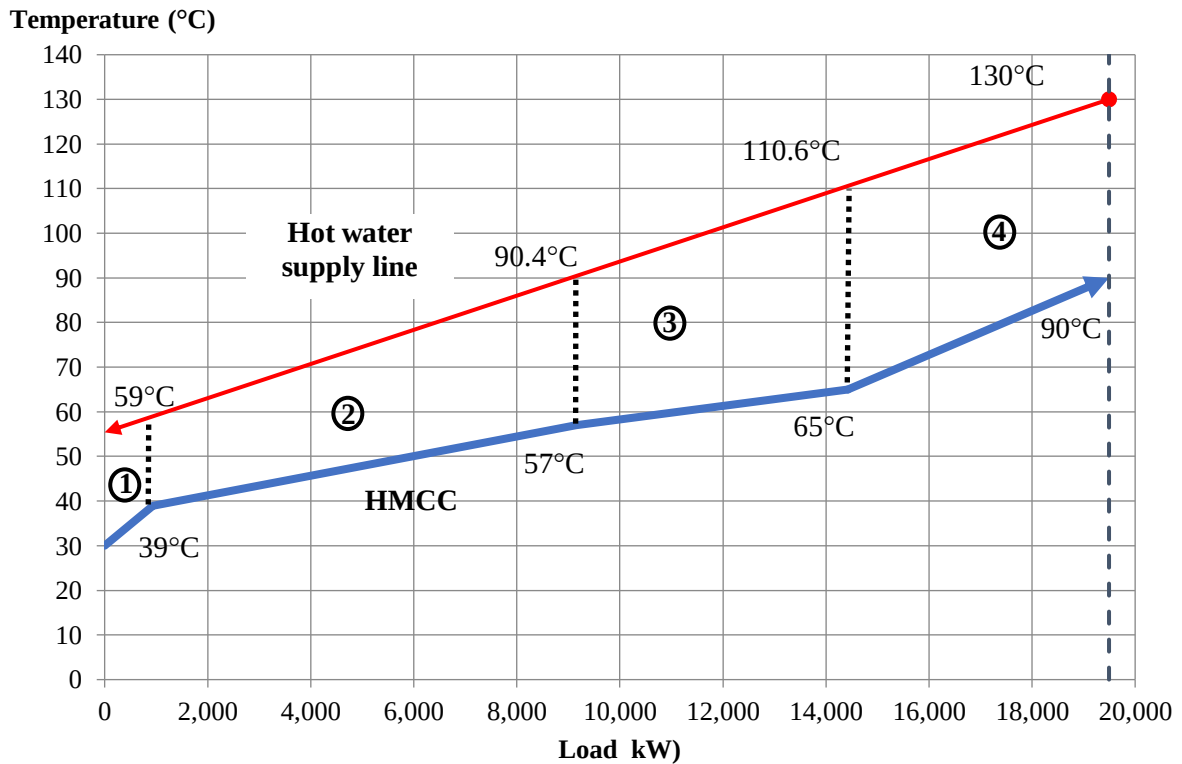


Figure 4.7. HMCC for industrial case study ($\Delta T_{\min} = 20^\circ\text{C}$)

Table 4.3.

Process stream data for industrial case study

Inlet Temperature, $T_{in, m} (^{\circ}\text{C})$	Outlet Temperature, $T_{out, m} (^{\circ}\text{C})$	Cold streams film heat transfer coefficient, $h_i (\text{kWm}^2/^{\circ}\text{C})$	Load, $Q_m (\text{kW})$	Heat capacity flowrate, $F_m (\text{kW}/^{\circ}\text{C})$
57	90	1.5	6,689	202.70
39	65	1	9,174	352.85
30	65	2	3,630	103.71
			19,493	659.26

The next step is to determine the heat transfer area of the WHRU. Common WHRU utilised at offshore platform is the finned tube-type heat exchanger, since it enhances heat transfer for fluids with lower film heat transfer coefficients. The use of fins attached to the external tube surface of the heat exchanger provides additional surface area for heat transfer. The steps for calculating the heat transfer area for the finned tube heat exchanger (WHRU) is shown in Appendix B.

The impacts of different $\Delta T_{\min}$ on the inlet temperature for the WHRU as well as its area are shown in Figure 4.8. As shown, the varying of $\Delta T_{\min}$ of the HEN directly affects the inlet temperature of the heating medium (water for this case), which consequently affects the area required for WHRU in similar relationship. The calculated heat transfer areas for the heat exchangers in the HEN and WHRU for a range of $\Delta T_{\min}$ are shown in Table 4.4.

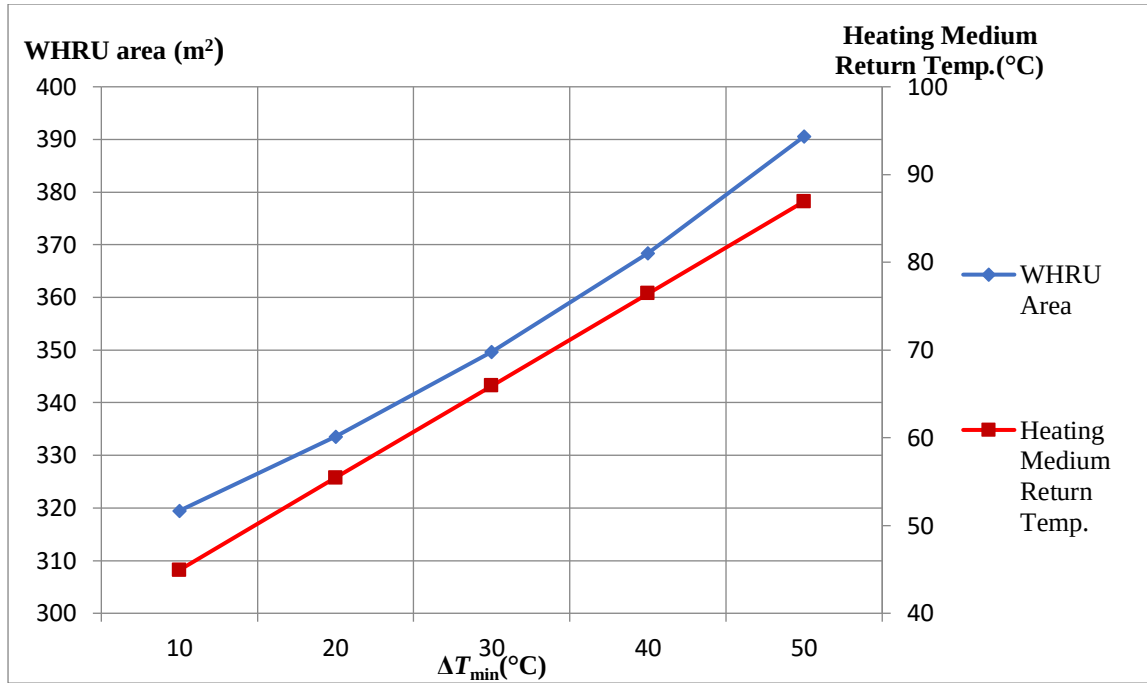


Figure 4.8. Effects of varying $\Delta T_{\min}$ on heating medium inlet temperature and area required for WHRU

Note that the flue gas from turbine leaves the WHRU at 150 °C, which is well above the dew point of the gas (71 °C). As such, the possibility of gas condensation (which will lead to corrosion in WHRU and network piping) is eliminated. From Table 4.4, it can be seen that the total area required for HEN reduces as $\Delta T_{\min}$ increases. This is due to the increment in log mean temperature as $\Delta T_{\min}$ increases. On the other hand, the opposite is observed for the WHRU, where increasing $\Delta T_{\min}$ has an inverse effect on its heat transfer area. This is because as $\Delta T_{\min}$ increases, the heating medium inlet temperature (to the WHRU) increases as well reducing the driving force temperature at the WHRU. The trade-off between the HEN and WHRU can be determined by calculating the cost these units, which will be discussed later.

Table 4.4.Heat transfer area for process heat exchangers and WHRU at different $\Delta T_{\min}$.

$\Delta T_{\min}$	Region in HEN (Figure 12)					WHRU				
	1 (m ²)	2 (m ²)	3 (m ²)	4 (m ²)	Total Area (m ²)	Heating medium inlet temp., T_{HMR} (°C)	Heating medium outlet temp., T_{HM} (°C)	Fluegas inlet temp., $T_{\text{in, g}}$ (°C)	Flue gas outlet temp., $T_{\text{out, g}}$ (°C)	Area (m ²)
10	76	654	200	142	1072	44.9	130	433	150	319
20	41	436	177	138	792	55.4				334
30	28	332	160	135	655	65.9				350
40	22	270	146	131	567	76.4				368
50	17	228	134	127	506	86.9				391

4.5.2. Pump Sizing

Apart from the heat transfer area of the HEN and WHRU, the pump plays a major role on both capital and operating costs of the heating medium network. The pump power is calculated based on the required head and the heating medium flowrate as shown in Appendix C. The the pipe and fittings losses ($hl_{p\&f}$) are assumed as 0.5 bar. While pressure drop across the heat exchanger units are assumed as 1.0 bar and the elevation head required is assumed as 3.5 bar ($d = 3.5 + 1.0$). The fluid density (ρ) and heat capacity is taken as of are taken as 984 kg/m³ and 4.326 kJ/kg°C respectively. The result for pump shaft power requirements for different $\Delta T_{\min}$ is summarised in Table 4.5.

From Table 4.5, it can be seen that increasing $\Delta T_{\min}$ has an inverse effect on the flowrate and pump power. This is because when $\Delta T_{\min}$ increases, the inlet temperature (T_{HMR}) of the heating medium to WHRU also increases. At fixed heat load, heating medium with smaller temperature change between T_{HM} and T_{HMR} will require more flowrate to meet the energy demand.

From the result presented earlier, it can be summarised that the increase of $\Delta T_{\min}$ reduces process heat exchanger area, but increases the area of WHRU, power required for pump and heating medium flowrate. By carrying out economic analysis on these variables, the optimum $\Delta T_{\min}$ which produces the overall lowest total cost can be determined.

Table 4.5.Pump flowrate and power required at different $\Delta T_{\min}$

$\Delta T_{\min}$	Heat capacity flowrate for hot water* (kW/°C)	Hot water flowrate (m ³ /hr)	Pump shaft power (kW)
10	252.04	213.13	49.3
20	287.54	243.15	56.3
30	334.68	283.01	65.5
40	400.30	338.50	78.4
50	497.74	421.06	97.5

*Note: with 10% additional margin

4.5.3 Capital and operating cost

Capital cost correlations for heat exchangers (CC_{HEN}), WHRU (C_{WHRU}) and pump (C_{PP}) as well the operating expenditures calculation are given in appendix D. Note that the capital cost correlation for WHRU is based on that for air cooled heat exchanger. The operating cost is calculated based on the assumption that the pump will be used for 8760 h. The tariff for electricity is taken as USD 0.12/kWh.

The calculated total overall and the breakdown cost for the given range of $\Delta T_{\min}$ are shown in Table 4.6. As shown, the smallest TAC is determined as \$204,357/y, corresponds to the optimum $\Delta T_{\min}$ of 20 °C.

Table 4.6.Total overall and the breakdown cost for variation of $\Delta T_{\min}$

$\Delta T_{\min}$	Annualised Capital Cost			Total annualised capital cost (\$/y)	Total utility cost (\$/y)	TAC (\$/y)
	Heat exchangers (\$/y)	WHRU (\$/y)	Pump (\$/y)			
10	114,702	41,831	9,321	165,854	51,862	217,715
20	91,703	43,466	10,022	145,191	59,166	204,357
30	80,317	45,329	10,894	136,540	68,865	205,406
40	73,135	47,483	12,022	132,640	82,368	215,008
50	68,059	50,023	13,555	131,636	102,458	234,094

4.5.4. HEN design with revised NNA

From the insight obtained from “Supertargeting”, an optimum HEN can be designed using the revised NNA. In order to apply the revised NNA, hot streams temperature profiles of the heating medium are to be developed, following the method outlined in Appendix A. Table 4.7 shows the resulting heating medium temperature profiles for the industrial case study.

Table 4.7.

Heating medium hot streams data profile for industrial case study

Unit, m	Inlet Temperature, $T_{\text{HM,in}, m} (^{\circ}\text{C})$	Outlet Temperature, $T_{\text{HM,out}, m} (^{\circ}\text{C})$	Load, Q_m (kW)	Heat capacity flowrate, F_m (kW/ $^{\circ}\text{C}$)
1	130	97	6,689	202.70
2	105	79	9,174	352.85
3	105	70	3,630	103.71
Total			19,493	659.26

As mentioned earlier, each heat exchanger is characterised as a pair of sink and source. The inlet streams for units 1, 2 and 3 are designated as sink SK1, SK2 and SK3; while the outlet streams of these units are designated as sources SR1, SR2 and SR3.

Following the revised NNA procedure, the matching matrix for the industrial case study is shown in Figure 4.9, where all sources (including the heating medium supply, i.e. HM) are listed in the first three columns, while all sinks (including heating medium return, i.e. HMR) are listed in the first three rows. Note that the hot water return temperature has been identified during the targeting stage as 55.4 $^{\circ}\text{C}$ (see Figure 4.7), which is included in the second entry of the last column in Figure 4.9.

		F_{SKj} (kW/°C)	202.7	352.85	103.71	261.40
		T_{SKj} (°C)	110	105	105	55.4
F_{SRi} (kW/°C)	T_{SRi} (°C)	SKj SRi	SK3	SK2	SK1	HMR
261.40	130	HM	126.21	97.21	37.98	
202.70	77	SR3	76.49	126.21		
352.85	59	SR2		129.42	65.73	157.69
103.71	50	SR1				103.71

(a)

		F_{SKj} (kW/°C)	126.21	223.42	103.71	261.4
		T_{SKj} (°C)	130	100.06	85	55.4
F_{SRi} (kW/°C)	T_{SRi} (°C)	SKj SRi	SK3	SK2	SK1	HMR
363.91	130	HM	126.21	97.21	37.98	
126.21	77	SR3	76.49 (LR)	126.21		
233.42	59	SR2		129.42 (LR)	65.73	157.69
103.71	50	SR1				103.71

(b)

Figure 4.9. Heating utility network design for industrial case study: (a) matching matrix; (b) removal of LR match

Note also that the optimum Pinch Temperature (59 °C, see Figure 4.7 segregates the heating medium utility system into higher (HQR) and lower quality regions (LQR). Sources in the HQR (with temperature higher than 59 °C) are prohibited to be allocated to sinks of the LQR (with temperature lower than 59 °C); the same principle applies to sources in the LQR and sinks in the HQR. This implies that the heating medium located in the HQR has to be used completely by the available sinks

in the same region and cannot be returned to heating medium heater which belongs to the LQR. Note however that source at the pinch temperature of 59 °C is an exceptional, as it belongs to both regions.

The HQR that consists of all sinks (SK1-SK3) as well as sources of hot water supply, SR1, and SR2 is first examined. We shall start the design with SK1, following Step 1 of the revised NNA procedure (see Figure 4.9). For sink SK1, since no source of the same temperature is available, step 3 of the procedure is applied where HM of just higher (130 °C) and SR1 of just lower temperature (77 °C) are identified as the candidate sources to be fed to SK1. Their allocated flowrates are determined using Eqs. (4.2) and (4.3). The same step is repeated for SK2 since it does not have a source of the same temperature too. SK2 is paired with HM of just higher (130 °C) and SR1 of just lower temperature (77 °C) and their allocated flowrates are determined using Eqs. (4.2) and (4.3). Note however that the SR1 does not have sufficient flowrate (126.21 kW/°C). SR1 is hence fully allocated to SK2. The flowrate and heat requirement of the latter is updated by deducting SR1 flowrate and heat supplied from the initial requirement of SK2. A new source candidate, i.e. SR2 of just lower temperature (59 °C) is paired with HM of just higher temperature (130 °C) to provide the flowrate and enthalpy requirements of SK2. The allocated flowrates are calculated again using Equations (4.2) and (4.3). For the final sink SK3, Step 3 is followed where HM (130 °C) and SR2 (59 °C) are paired. Equations (4.2) and (4.3) determine that the HM (water) is used completely. Note that this follows the earlier discussion that the HM cannot be sent back to the heater (i.e. the sink HMR). Finally, the unutilized SR2 (157.69 kW/°C) and SR3 (103.71 kW/°C) are sent to the heater for reheating (HMR)

Following Step 4 of the revised NNA, all LR matches of the heating utility system are eliminated. Doing this reduces the source flowrates that pass through the heat exchangers, which leads to smaller and cheaper units. As shown in Figure 4.9, the LR match of SR1- SK1 and SR2-SK2 are eliminated. Following the elimination of the LR matches, the temperature level of SK1 and SK2 and the corresponding flowrates of SR1, SK1, SR2 and SK2 are adjusted (indicated by the bold font). The final configuration of the heating utility network for the hypothetical example is shown in Figure 4.10. The total area of heat exchange for this configuration is determined as 965 m², which is approximately 22% larger from that identified from “Supertargeting” (i.e. 793 m², see Table 4.4). For comparison

with the conventional parallel design, where heating medium is fed to all heat exchangers at 130 °C and returns at 90 °C, the TAC is determined as \$253,227 (see detailed cost elements in Table 4.8). The mixed series/parallel design in Figure 4.10 entails a reduced TAC of \$211,949/y (see detailed cost elements in Table 4.8), which is 16% lower as compared to the conventional design. This is a sizeable saving since offshore platforms are designed for operation between 20 and 30 years. Besides, the new proposed design can be easily implemented subject to operation and safety evaluation.

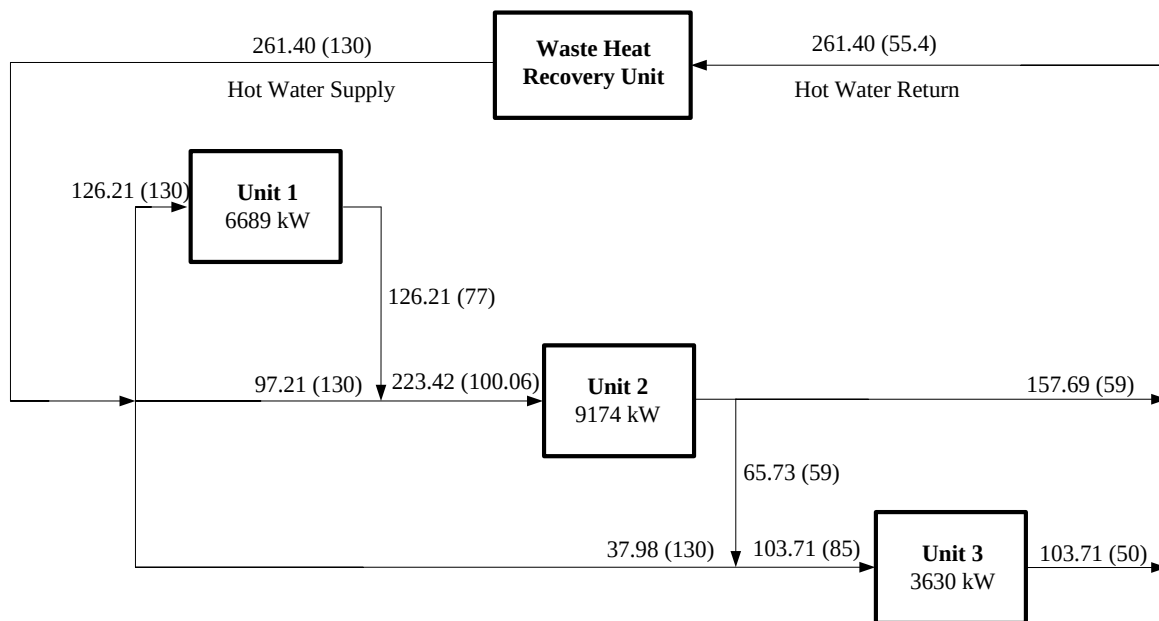


Figure 4.10. Final configuration of heating utility network for industrial case study (heating medium flowrate in kW/°C; temperature in °C, given in parenthesis)

Table 4.8.

Comparison of cost between optimum and conventional design

Description	Optimum design (this work)	Conventional design
HEN (\$/y)	99,295	59,654
WHRU (\$/y)	43,466	52,323
Pump (\$/y)	10,022	15,192
Total Capital Cost(\$/y)	152,783	127,169
Total Utility Cost (\$/y)	59,166	126,058
TAC (\$/y)	211,949	253,227

4.6. Summary

This chapter presented some extended approaches in designing heating utility system for offshore platform. Process integration approach based on pinch technique is adopted here. With the use of the revised graphical targeting method, the minimum heating medium flowrate and HEN area of the heating utility system can be targeted ahead of detailed design. This also allows the targeting for minimum TAC to be performed. This approach is particularly useful during preliminary design where heat exchanger data is not available. The advantage of the revised graphical method is that, it allows the targeting of heat exchanger area to be performed for HEN and WHRU, resulted in an optimum cost-optimum solution, without having to design the HEN.

A revised NNA technique may be used to design the optimum heating utility system network to achieve the identified targets. The proposed methods were applied on a literature example and an industrial case study. The solutions obtained for the literature example is found to be identical. The industrial case study, on the other hand showed a reduction of 17% for its TAC as compared to the conventional design.

CHAPTER 5

AUTOMEATED TARGETINTG TECHNIQUE FOR HEATING UTILITY SYSTEM

5.1 Introduction

The chapter is organised as follows. In the following section, the problem statement is given. Next, the underlying principles of the automated targeting model (ATM) model is explained for the application on heating utility system. Next, the formulation of ATM framework is provided for simultaneous optimisation of utility and heat exchanger network (HEN). The model is extended to include the regeneration of heating medium in the optimisation model and a case study involving an oil and gas production facility is used to illustrate the procedures in obtaining the optimum overall cost of heating utility system.

5.2 Heating utility system and problem statement

The formal problem for heating medium system design can be stated as follows. Given is a set of heating medium user (utility heat exchangers) $m \in M$, each with a fixed heat load Q_m to be provided by a single heating medium (water or hot oil). The process inlet ($T_{in, m}$) and outlet ($T_{out, m}$) temperatures and the heat capacity flowrates of m users are specified. The heating medium is reheated at the heater before being re-circulated among the heat exchangers, with variable supply (T_{HM}) and return temperatures (T_{HMR}).

The main objective is to synthesise an optimum heating medium network with the lowest total annualised cost. This is done by two steps, following the conventional Pinch Analysis technique. In the first step, the extended ATM model is used to determine the performance targets of the network by locating the optimum minimum approach temperature (ΔT_{min}), supply and return temperature as well as the optimum flowrate of the heating medium that lead to the lowest total annualised cost. In the second steps, the performance targets obtained are used to design the heating medium network using the *pinch design method* (Linnhoff and Hindmarsh, 1983). The optimised network may be different from conventional design shown in Figure 1.1.

5.3 Underlying principles of ATM framework

In this chapter, the ATM technique (Ng et al., 2007a) is extended for the targeting of minimum heating medium system cost. Note that in the earlier chapter, the heating medium inlet temperature is fixed, while the outlet temperature is set as a variable. With the ATM technique proposed in this section, both inlet and outlet temperatures are now set as variables and subjected for the optimisation study. Besides, another important improvement from the graphical approach is that the ATM is implemented on an optimisation framework eliminating the iterative procedure for finding the optimum solution. It is capable of locating the exact $\Delta T_{\min}$ that leads to the minimum total annualised cost (TAC).

A new framework is presented in this work (see latter section) to determine the heating medium flowrate and the overall heat transfer area needed for the heat exchangers. It is best to illustrate the concept for the newly proposed framework by revisiting the heating medium composite curve (HMCC) as shown in Figure 5.1. In Figure 5.1(a), the heat transfer area required for the entire HEN is estimated by adding the individual area required in each load interval; the latter is identified by locating the turning points of the HMCC. Each turning point of the HMCC represents a temperature level in the framework (see latter discussion), which can be calculated from the gradient of the supply line and the enthalpy level. Note that, the inverse of the slope also provides information on the required heating medium flowrate. By changing the gradient of the heating medium supply line, different sets of results for heat transfer area, heating medium flowrate and return temperature can be obtained, as shown in Figure 5.1(b). Hence, in a system of a single heating medium and multiple heat exchangers, changing the $\Delta T_{\min}$ value changes the gradient of the supply line, and hence the required heat transfer area of the heat exchangers in the network. This underlying relationship between the supply line and the HMCC is used to target the heat transfer area for the range of $\Delta T_{\min}$ values that was specified for the problem.

In the graphical approach the heating medium supply temperature (T_{HM}) is fixed, hence allowing the supply line to be rotated at the pivot point to find the optimum solution. In this work, the

T_{HM} will be set as a variable which result to numerous possible orientations for the supply line, as shown in Figure 5.2.

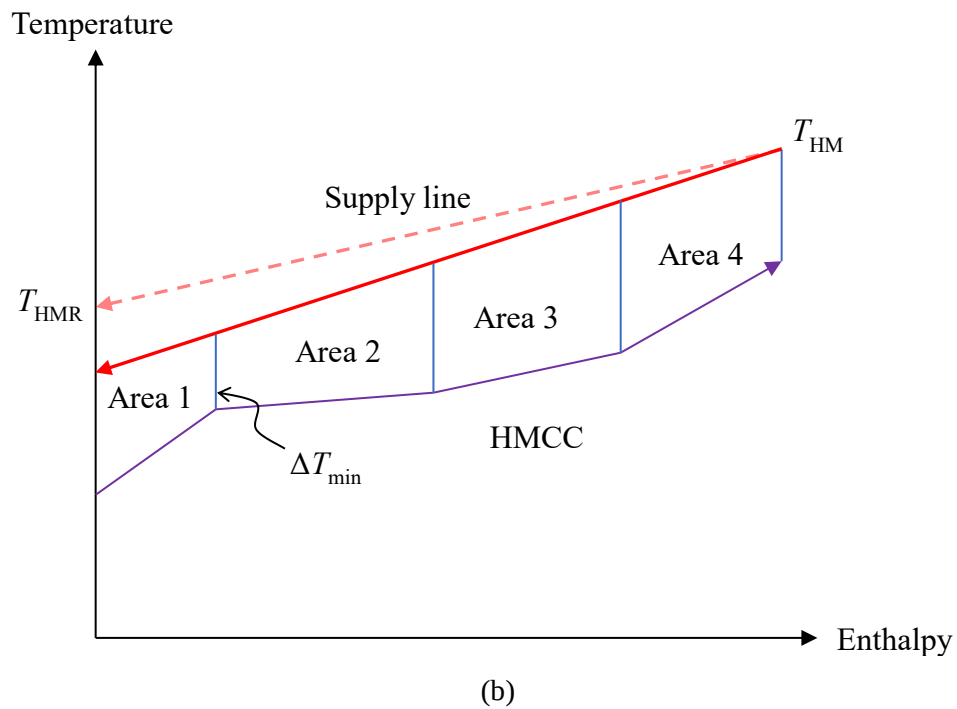
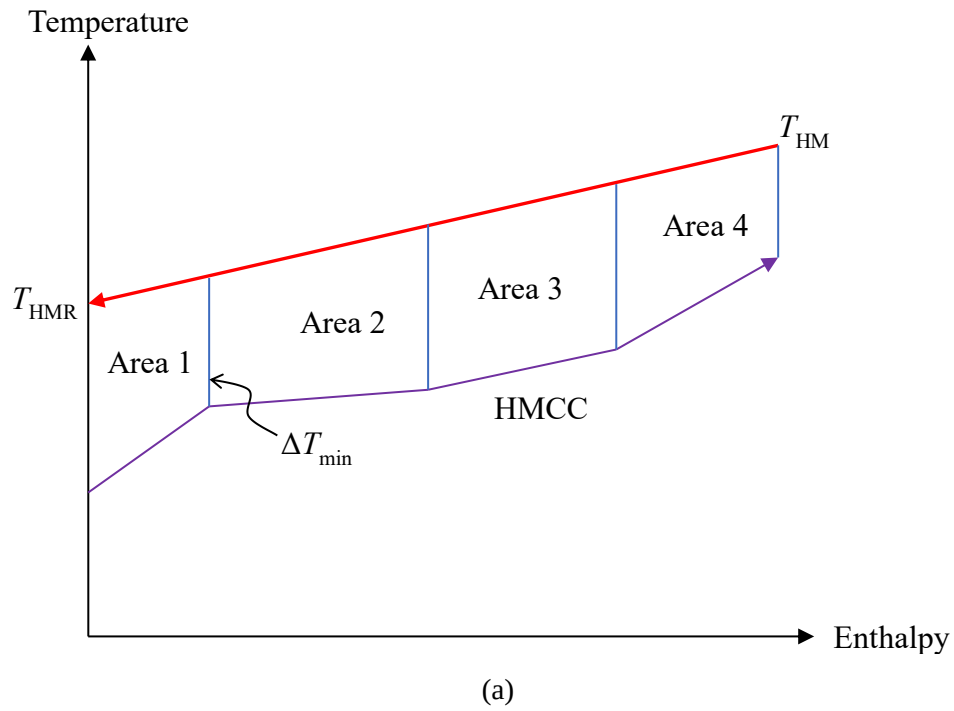


Figure 5.1. Area targeting by changing the slope of supply line: (a) larger ΔT_{min} (b) smaller ΔT_{min}

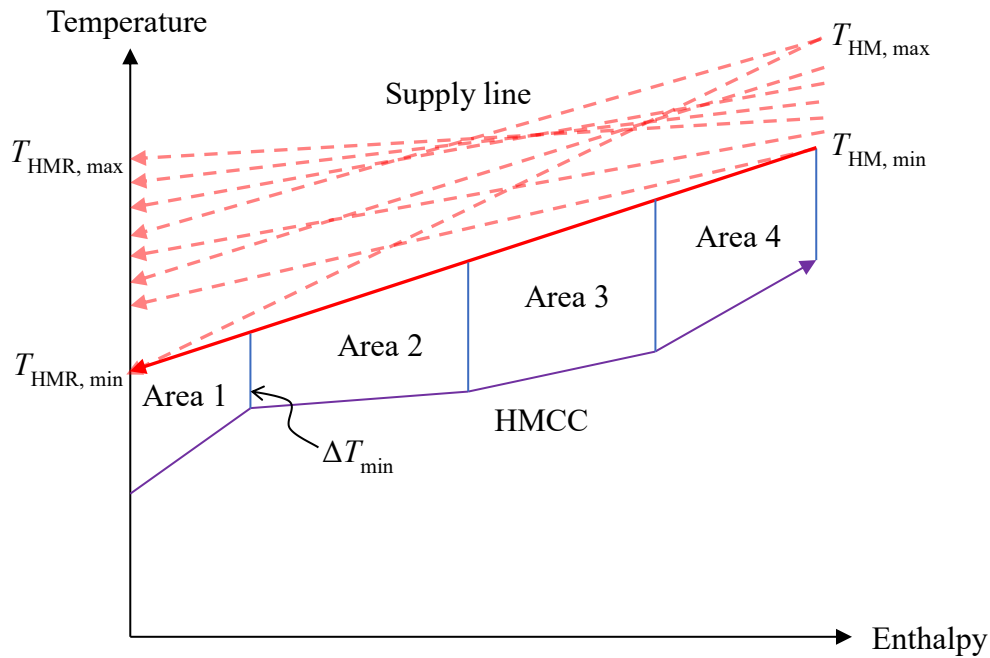


Figure 5.2. Numerous possible orientations for supply line when their inlet and outlet temperatures are varied

In order to help the computational of the model, the maximum and the minimum heating medium flowrates has to be determined in priori and be added to the model as part of the constraints. It is known that smallest heating medium flowrate occurs when the slope is at the steepest. For a heating medium with variable inlet and outlet temperatures, this scenario occurs when the heating medium is operated at its maximum inlet temperature and minimum approach temperature, as shown in Figure 5.3.

On the other hand, the maximum heating medium flowrate occurs when the temperature difference between the T_{HM} and T_{HMR} is the smallest. When this scenario take place, the gradient of the supply line is at the least steep. Note that, this case can occur at the minimum T_{HM} value ($T_{HM, min}$) as shown in Figure 5.4 or at maximum T_{HMR} ($T_{HMR, max}$), or any temperature in between. What is critical is the temperature difference between T_{HM} and T_{HMR} , which has to be the smallest difference. For design case, this difference can be set as 1 °C. Although, the subsequent flowrate can be extremely large, it ensures that all possible heating medium flowrate is covered when determining the optimum solution.

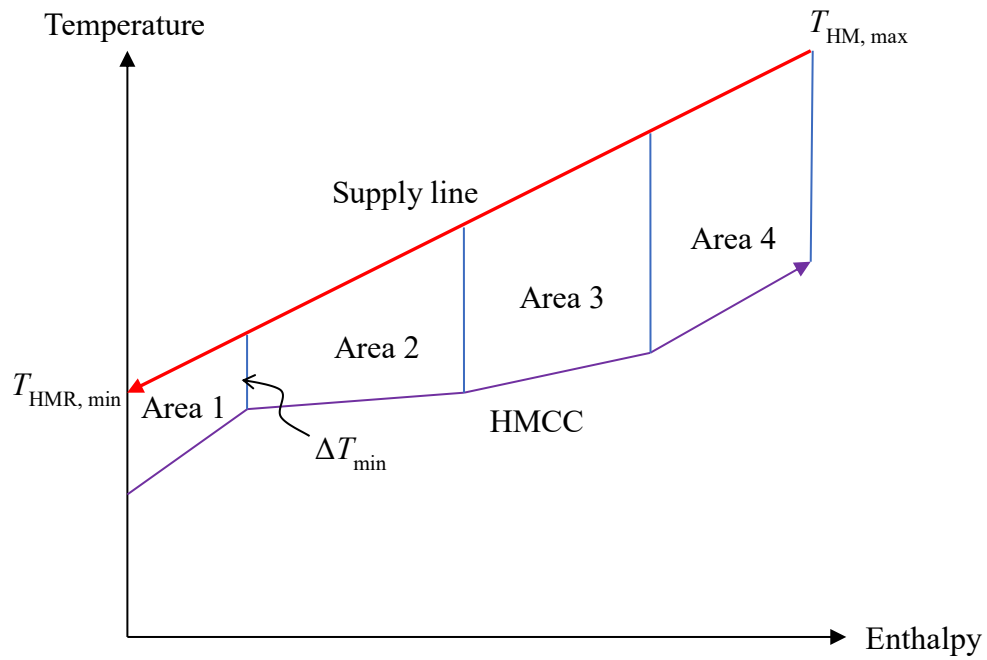


Figure 5.3. Determining the lowest possible heating medium flowrate

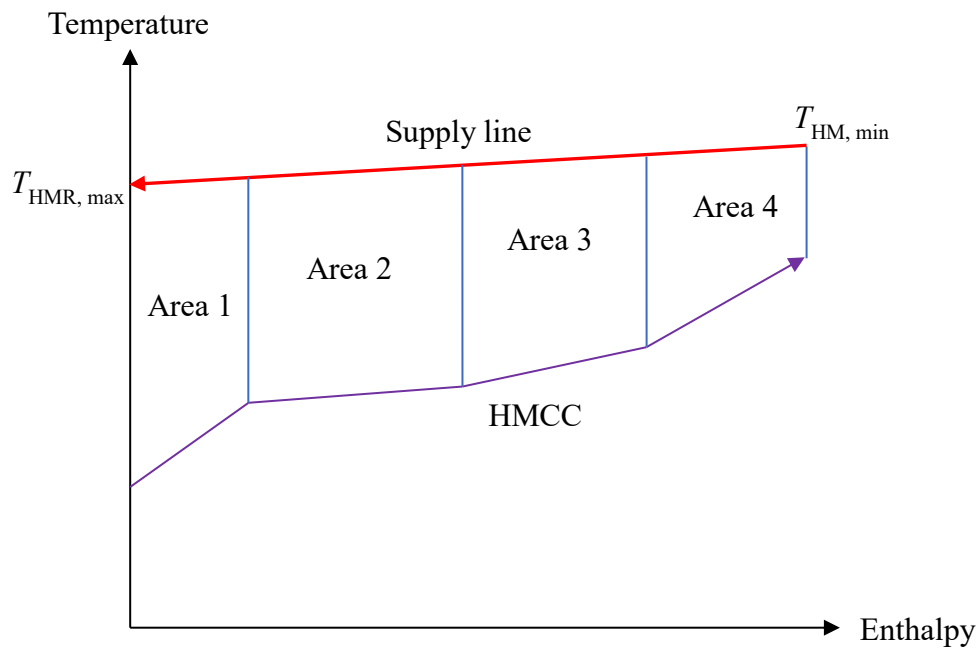


Figure 5.4. Determining the highest possible heating medium flowrate minimum T_{HM}

The graphical representation of the ATM for HEN area targeting is shown in Figure 5.5 and it is deduced into a set of equations in the following section.

5.4 Formulation for the ATM framework

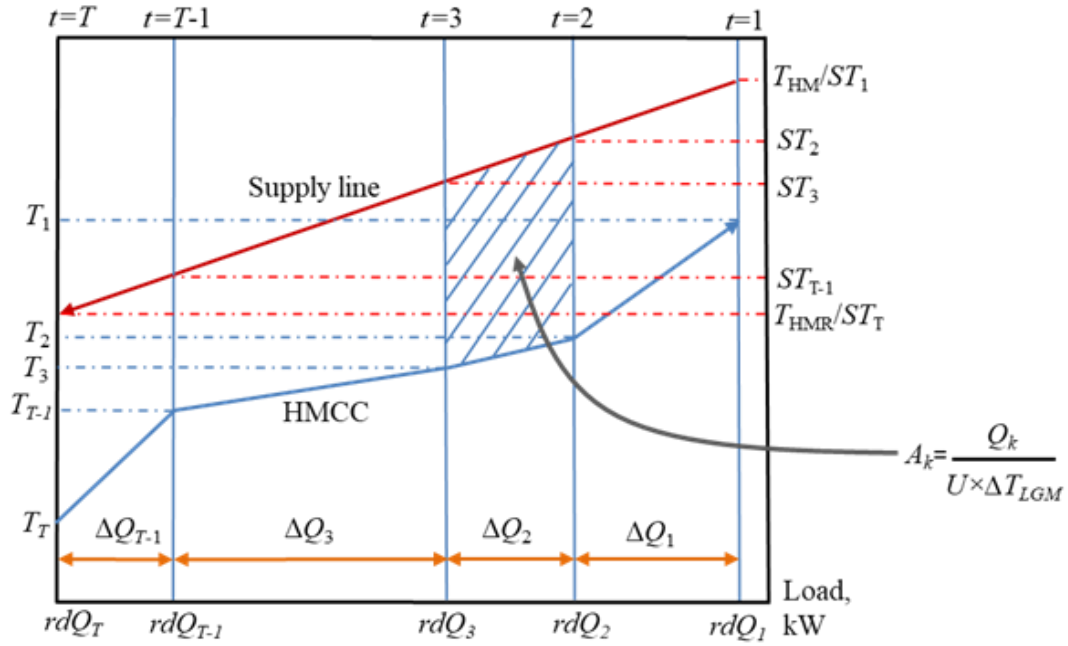


Figure 5.5. ATM approach in determining the heat transfer area required for HEN in thermal heating utility system

The generic ATM framework are summarised in Tables 5.1 and 5.2. Table 5.1 is used to determine the heating medium flowrate, while Table 5.2 determines the heat exchanger area. Both tables work coherently in order to determine the optimum TAC, in which information from Table 5.1 is to be used by Table 5.2. Table 5.1 is first discussed as follows.

In the first column of Table 5.1, the supply and target temperatures (T_i) of all the heating medium users are arranged together, in descending order since the problem involves heating. Note that each temperature represents a turning point in the HMCC (see discussion for Figure 5.1). Next, all heating medium users are located at their respective temperature intervals in the following column.

Table 5.1.

Framework for targeting optimum heat capacity flowrate of heating medium

Temperature, T_t (°C)	Heat capacity flowrate, $F_{m,t}$ (kW/°C)	Interval Heat Load, ΔQ_t (kW)	Heat Load Cascade, dQ_t (kW)	Capacity flowrate, F_t (kW/°C)
T_1			$dQ_1 = 0$	$F_1 = dQ_1/(T_{HM} - T_1 - \Delta T_{min})$
	↑	ΔQ_1		
T_2			dQ_2	$F_2 = dQ_2/(T_{HM} - T_2 - \Delta T_{min})$
	↑	ΔQ_2		
T_3			dQ_3	$F_3 = dQ_3/(T_{HM} - T_3 - \Delta T_{min})$
⋮	↑			
T_{T-1}	2 3		dQ_{T-1}	$F_{T-1} = dQ_{T-1}/(T_{HM} - T_{T-1} - \Delta T_{min})$
		ΔQ_{T-1}		
T_T	1		dQ_T	$F_T = dQ_T/(T_{HM} - T_T - \Delta T_{min})$

Eq. (5.1) is used to calculate heat load for each temperature interval, and documented in the third column:

$$\Delta Q_t = \sum_m F_{m,t}(T_t - T_{t+1}) \quad \text{for } t = 1, 2, 3 \dots T - 1 \quad (5.1)$$

In the fourth column, heat load cascade at each level t is calculated by adding heat load at each interval, as shown in Eq. (5.2):

$$dQ_t = \begin{cases} 0 & t = 1 \\ dQ_{t-1} + \Delta Q_{t-1} & t = 2, 3 \dots T - 1, T \end{cases} \quad (5.2)$$

Note that net heat load of level 1 is always set to zero, as there is no heat load entry at the first level. The last entry in this column represents the total heat load required by the heating medium users. In the next column, the heat capacity flowrate of the heating medium required for level t (F_t) is calculated using Eq. (5.3), i.e. division between heat load at temperature interval t (dQ_t) and the difference between the heating medium supply temperature (T_{HM}), the level temperature (T_t), and ΔT_{min} .

$$F_t = \frac{dQ_t}{(T_{HM} - T_t - \Delta T_{min})} \quad \forall t \quad (5.3)$$

The minimum approach temperature (ΔT_{min}) has to be included as it provides the minimum driving force for heat transfer to take place between the hot and the cold stream. For practical heat transfer consideration, the value of ΔT_{min} is normally set to 10 °C, or larger.

By incorporating ΔT_{min} value into Eq. (5.3), the heating medium user temperature can be determined for every ΔT_{min} value. This is a modification to the conventional approaches where the ΔT_{min} is deducted from process temperatures e.g. Problem Table Analysis for HEN synthesis (1978) and the algebraic approach proposed by Bade and Bandyopadhyay (2014).

The return temperature of the heating medium (T_{HMR}) is calculated using Eq. (5.4), as follows,

$$T_{HMR} = T_{HM} - \frac{dQ_T}{F_{opt}} \quad (5.4)$$

where dQ_T is the total heat load needed by the heating medium system (given by the last entry in column 4), F_{opt} is total optimum heating medium flowrate required to meet the energy demand of the entire system; the latter is obtained by identifying the largest value among the heat capacity flowrate (F_i) of the heating medium required at each level t , as shown in Eq. (5.5).

$$F_{opt} = MAX\{F_t\} \quad \forall t \quad (5.5)$$

Next, Table 5.2 that is used for determining the optimum heat exchange area is explained. For ease of reference, the first two columns in Table 5.2 are set to follow those in Table 5.1.

In the third column, the heat load cascade of the heating medium users is constructed in reverse order following Eq. (5.6). The heat load values are calculated in this order so that the heating medium temperature ST_t (calculated using Eq. (5.7)) produces temperature in descending order. Note that at level $t = 1$, rdQ_t takes the value of dQ_T which is the total heat load required by the system (see last entry in column 4, dQ_T in Table 5.1).

$$rdQ_t = \begin{cases} dQ_T & t = 1 \\ rdQ_{t-1} - \Delta Q_{t-1} & t = 2, 3 \dots T-1, T \end{cases} \quad (5.6)$$

In the following column, the corresponding temperature of the heating medium supply line (ST_t) at each temperature level t (corresponding to the turning points of the HMCC) is determined using Eq. (5.7):

$$ST_t = \frac{(rdQ_t - dQ_T)}{F_{opt}} + T_{HM} \quad \forall t \quad (5.7)$$

Note that in Eq. (5.7), ST_t is determined using the gradient of the supply line, F_{opt} (determined using Eq. (5.5)) and the cumulative heat load at each temperature level, rdQ_t (calculated using Eq. (5.6)). The first entry of ST in column 4 represents the heating medium temperature T_{HM} (as first term in Eq. (5.7) becomes zero), while the last entry represents the heating medium return temperature T_{HMR} .

The minimum approach temperature (ΔT_{min}) for the HEN is determined from the lowest value among temperature differences between the heating medium and that of the HMCC at level t , given by Eq. (5.8):

$$\Delta T_{min} = \text{MIN}\{ST_2 - T_2, ST_3 - T_3, \dots, ST_{T-1} - T_{T-1}, ST_T - T_T\} \quad (5.8)$$

This value represents the smallest difference between the source and the sink temperature at a temperature level t , which defines the pinch location of the HEN. The hot and cold pinch temperatures can then be identified as the temperature sets where pinch is located.

As discussed earlier, the optimum heating medium flowrate (F_{opt}) should be bound between its maximum (F_{max}) and minimum values (F_{min}), as shown by Equation (5.9).

$$F_{min} \leq F_{opt} \leq F_{max} \quad (5.9)$$

The value of F_{min} can be calculated by equation (5.10).

$$F_{min} = \frac{dQ_T}{\Delta T_{HM}} \quad (5.10)$$

Table 5.2.

ATM area targeting Framework

Temp., T_i (°C)	HM users	Heat load cascade, rdQ_t (kW)	Corresponding temperature of supply line, ST_i (°C)	Log mean temperature, T_{LGMt} (°C)	Hot area, $A_{t,h}$ (m ²)	Cold area $A_{t,c}$ (m ²)	Total area, A_{HEN} (m ²)
T_1		rdQ_1	T_{HM}				$\sum_t (A_{t,h} + A_{t,c})$
				$\frac{(ST_1 - T_1) - (ST_2 - T_2)}{\ln[(ST_1 - T_1) - (ST_2 - T_2)]}$	$\frac{F_{opt}(ST_1 - ST_2)}{h_{HM}T_{LGM1}}$	$\sum_m \frac{F_m(T_1 - T_2)}{h_m T_{LGMT1}}$	
T_2		rdQ_2	$\frac{(rdQ_2 - dQ_T)}{F_{opt}} + T_{HM}$				
				$\frac{(ST_2 - T_2) - (ST_3 - T_3)}{\ln[(ST_2 - T_2) - (ST_3 - T_3)]}$	$\frac{F_{opt}(ST_2 - ST_3)}{h_{HM}T_{LGM2}}$	$\sum_m \frac{F_m(T_2 - T_3)}{h_m T_{LGMT2}}$	
T_3		rdQ_3	$\frac{(rdQ_3 - dQ_T)}{F_{opt}} + T_{HM}$				
				$\frac{(ST_3 - T_3) - (ST_{t-1} - T_{t-1})}{\ln[(ST_3 - T_3) - (ST_{t-1} - T_{t-1})]}$	$\frac{F_{opt}(ST_3 - ST_{t-1})}{h_{HM}T_{LGM3}}$	$\sum_m \frac{F_m(T_3 - T_{t-1})}{h_m T_{LGMT3}}$	
T_{T-1}	2 3	rdQ_{T-1}	$\frac{(rdQ_{T-1} - dQ_T)}{F_{opt}} + T_{HM}$				
				$\frac{(ST_{t-1} - T_{t-1}) - (ST_t - T_t)}{\ln[(ST_{t-1} - T_{t-1}) - (ST_t - T_t)]}$	$\frac{F_{opt}(ST_{t-1} - ST_t)}{h_{HM}T_{LGMt-1}}$	$\sum_m \frac{F_m(T_{T-1} - T_T)}{h_m T_{LGMT-1}}$	
T_T	1	rdQ_T	$\frac{(rdQ_T - dQ_T)}{F_{opt}} + T_{HM}$				

Note that the value of $F_{\max}$ can be determined by solving Eqs. 5.1-5.5, with T_{HM} set to the highest possible value and $\Delta T_{\min}$ to the smallest allowable value.

In the fifth column, the log mean temperature ($T_{\text{LGM}t}$) between the hot and cold stream is calculated for each interval t , using Eq. 5.11.

$$T_{\text{LGM}t} = \frac{[(ST_t - T_t) - (ST_{t+1} - T_{t+1})]}{\ln[(ST_t - T_t) - (ST_{t+1} - T_{t+1})]} \quad t = 1, 2, 3 \dots T - 1 \quad (5.11)$$

Next, in column 6, the heat transfer area required based on enthalpy intervals at each interval t is determined using equation (5.12) (Townsend and Linnhoff, 1984) for the hot side. Note that as mentioned earlier, the ATM allows log mean temperature to be used in calculating the driving force, rather than other approximation equations which underestimate (e.g. Chen approximation) or overestimate (e.g. Paterson approximation) the driving force, which then leads to inaccurate area for the HEN (Yee et al., 1990).

$$A_{t,h} = \frac{F_{opt}(ST_t - ST_{t+1})}{h_{\text{HM}}T_{\text{LGM}t}} \quad t = 1, 2, 3 \dots T - 1 \quad (5.12)$$

where h_{HM} is the film heat transfer coefficient of the heating medium. In the subsequent column, the area required for the cold stream side is calculated using Eq. (5.13).

$$A_{t,c} = \sum_m \frac{F_m(T_t - T_{t+1})}{h_m T_{\text{LGM}t}} \quad t = 1, 2, 3 \dots T - 1 \quad (5.13)$$

where h_m is the film heat transfer coefficient of the heating medium users. In the final column, the sum of the total area required at each interval t (A_{HEN}) is determined by adding the values of hot and cold area calculated in column 5 and 6, following Eq. 5.14.

$$A_{\text{HEN}} = \sum_t A_{t,h} + A_{t,c} \quad (5.14)$$

The hot ($T_{p,h}$) and cold ($T_{p,c}$) pinch temperature of the optimum solutions can be obtained through Eqs. 5.15-5.18, where ω_t is a binary variable.

$$(ST_t - T_t)\omega_t = \omega_t \Delta T_{\min} \quad t = 2, 3, 4 \dots T \quad (5.15)$$

$$\sum_t \omega_t = 1 \quad (5.16)$$

$$T_{P,h} = \sum_t ST_t \omega_t \quad t = 2,3,4 \dots T \quad (5.17)$$

$$T_{P,c} = \sum_t T_t \omega_t \quad t = 2,3,4 \dots T \quad (5.18)$$

Up to this point, the two frameworks determine the flowrate target of the heating medium and the area target for heat transfer, for a given range of $\Delta T_{\min}$. Using these information, the pump power, capital (for heat exchangers and pump) and utility costs of the thermal heating utility system can be determined. Detailed pump power and cost calculations for heat exchangers, pump and utility are given in Appendix C and D respectively.

The supply and return temperature of the heating medium would affect the cost of the waste heat recovery unit (WHRU) and hence the overall cost of the heating medium system. Note that, a lower supply and/or return temperature of the heating medium means that bigger temperature driving force exists between the flue gas and the heating medium at the WHRU, which subsequently leads to lower cost of the WHRU (smaller area). Besides, a lower return temperature reduces the pump and utility costs but increases the cost of HEN. Note however, lowering the supply temperature has a different effect, where all cost elements (i.e. pump, utility and HEN) will increase. Besides, the heating medium temperatures also effect the cost of pipings. For example, when both the heating medium supply and return are at maximum, a higher flowrate will be required to supply heat to its users. A higher heating medium flowrate will require larger pipes to meet velocity limits and to avoid larger pressure drop across pipe and fittings. Thus, there exists a clear trade-off among the WHRU design, head loss and piping cost in the optimisation model. The sizing of WHRU is provided in Appendix B while the costing of a WHRU is shown in Appendix D. The pipe and fitting losses calculation together with piping costing is shown in Appendix E.

Finally, the optimisation function is set to minimise the TAC1 of the heating medium system, which consists of the cost of heat exchanger (C_{HEN}), WHRU (C_{WHRU}), pump (C_{PP}), piping (C_{PIPE}) and utility (C_U), given by Eq. (5.19). The problem will also determine the optimum $\Delta T_{\min}$ that gives the minimum cost solution.

$$TAC1 = C_{HEN} + C_{WHRU} + C_{PP} + C_U + C_{Pipe} \quad (5.19)$$

5.5 Industrial case study

Industrial case study in Section 4.5 is revisited. All data used remains the same apart from the heating medium temperature is now set to vary between 100 and 130 °C. Note that the maximum temperature should not exceed 130 °C so to avoid Napthenate scale formation in the pipe lines.

WHRU is a finned tube-type heat exchanger, commonly used for offshore application since it enhances heat transfer for fluids with lower film heat transfer coefficients. The addition of fins to the external tube surface increases the surface area for heat exchange, which leads to improved overall heat transfer coefficient, U_{inn} . The steps involved in calculating this enhanced overall heat transfer coefficient, U_{inn} is given in Appendix B.

According to Smith (2005), pressure drop for short transmission pipe (<100 m) is usually nominated by the pressure drop in pipe fittings. Thus, ATM model includes not only the head loss in piping but also the head loss in valves, bends, entrance and exit to equipment as well. The number of valves, bends, equipment entrances and exits that are in line with the heating medium loop are specified in Table 5.3. For piping cost calculation, the information required is given in Table 5.4

Table 5.3.

Fittings specification for heating medium system

Location	Type	Function	Total
Heat exchanger	Butterfly valve	Isolation	9 (3×3 user)
	Check valve	Safety	3 (1×3 user)
	Globe valve (50% opening)	Control	3 (1×3 user)
	Entrance	-	3 (1×3 user)
	Exit	-	3 (1×3 user)
	Bends	-	12 (4×3 user)
Heating medium pump	Butterfly valve	Isolation	4
	Check valve	Safety	1
	Bends	-	4
WHRU	Gate valve (SDV)	Isolation	1
	Gate valve	Isolation	3
	Check valve	Safety	1
	Entrance	-	1
	Exit	-	1
	Bends	-	4

Table 5.4.

Data for pump power and equipment costing calculation

Parameters	Value
Pipe cost per unit length, $J1$	0.82 \$/kg
Installation cost, $J2$	1.85 \$/m ^{0.48}
Pipe length, L	80 m

Two scenarios are analysed for this case study. In Scenario 1, the T_{HM} is set at 130 °C and the output temperature is allowed to vary. The pipe and fitting losses are assumed to be 0.5 bar similar to the case in Section 4.5, while the piping cost are excluded. Solving the objective function in Eq. (5.19) (excluding C_{PP}) subject to the constraints in Eqs. 5.1-5.8, 5.11-5.18, B1-B8 (for WHRU area) C1 (for pump power) and D1-D5 (capital and utility cost) yields the minimum TAC of the heating medium system as \$197,027, with the ATM results given in Tables 5.5 and 5.6. The model contains 33 variables and 35 constraints, solved using Lingo 14 and the computational time required was 4 s using Intel i5 6200U 2.30GHz processor. The optimum ΔT_{min} determined by the ATM is found at 23.7 °C. Note that in the work of Section 4.5, the optimum ΔT_{min} value was reported as 20 °C, as only a limited ΔT_{min} values (10, 20, 30 °C, etc.) were considered. Hence, the TAC was determined as \$204,357, which is higher than that reported by the current ATM approach.

Table 5.5.

ATM results for targeting heating medium flowrate for industrial case study (Scenario 1)

Temperature, T_i (°C)	Heat capacity flowrate, $F_{m,t}$ (kW/°C)	Interval Heat Load, ΔQ_t (kW)	Heat Load Cascade, Q_t (kW)	Heat capacity flowrate, F_{opt} (kW/°C)
90			0	0
		5067		
65			5067	122.78
		5274		
57	3		10341	209.89
		8218		
39	2		18560	275.89
		933		
30	1		19493	255.57

Table 5.6.

ATM results for targeting HEN for industrial case study (Scenario 1)

Temp., T_i (°C)	HM Users	Heat load cascade, rdQ_i (kW)	Heating medium temperature, ST_i (°C)	Log mean temperature, T_{LGMi} (°C)	Hot area, $A_{t,h}$ (m ²)	Cold area, $A_{t,c}$ (m ²)	Total area, A_{HEN} (m ²)
90		19493.00	130.00				732.28
				43.23	58.61	78.14	
65		14425.58	111.63				
				40.82	64.60	105.79	
57	3	9151.52	92.52				
				29.23	140.59	249.24	
39	2	933.43	62.73				
				26.44	17.65	17.65	
30	1	0	59.35				

In Scenario 2, both the inlet and outlet temperatures of the heating medium are allowed to vary. The pipe and fitting losses are calculated based on the equations in Appendix E, while the pipe costing is included as part of the optimisation function. For this case, $F_{\min}$ value was calculated by solving Eqs. (5.1) -(5.5) as 229.13 kW/°C, while $F_{\max}$ was determined as 19,493 kW/°C using Eq. (5.10). These two values are added to Eq. (5.9) as part of the constraints. For Eq. (E43), the last two terms, J3 and J4 are removed. This is because the right of way cost is not foreseen in FPSO. While, the insulation cost seems unsuitable for this case study because according to SHELL design engineering practice (Royal Dutch Shell, 2012) barrier insulation is sufficient for temperature below 220 °C which is not equivalent to thermal insulation.

Solving the objective function in Eq. (20), subject to the constraints in Eq. (5.19), Eqs. (5.1) - (5.8), (5.11)-(5.18), B1-B8 (for WHRU area), C1 (for pump power), D1-D5 (capital and utility cost) and E1-E43 (pipe and fitting losses and piping cost) yields the minimum TAC of the heating medium as \$195,913, with the ATM results given in Tables 5.7 and 5.8. The model contains 69 variables and 80 constraints, solved using Lingo 14 and the computational time required was 230 s using Intel i5 6200U 2.30GHz processor. The optimum $\Delta T_{\min}$ determined by the ATM is found at 24.0 °C, resulting

with the cold and hot Pinch temperatures as 39 °C and 63 °C. The T_{HM} and T_{HMR} values are 130 °C and 59.66 °C respectively (first and final entries of column 4; Table 5.8). Key result of case study is shown in Table 5.9.

Note that, even with the addition of piping cost in Scenario 2, it produces a lower optimum cost than in case study 1. This was due the pipe and fitting losses, which was estimated higher than the actual value in Scenario 1 (0.5 bar). The actual value calculated in Scenario 2 (0.263 bar) is less than half of the assumed value, driving the pump and utility cost lower. The velocity of fluid that produces the optimum solution in Scenario 2 is 0.91 m/s, which is equal to the specified lower limit. For the conventional design, the velocity of the fluid that produces the lowest pump and piping cost is also 0.91 m/s. Note that however, larger pipe size is required compared to the ATM result in order to achieve the same velocity. Since the velocity is equal, pressure drop calculated for ATM (0.263 bar) and conventional design (0.259 bar) is relatively similar with the small difference is attributed to the Reynolds number. Based on this result, a conclusion can be drawn that it is more economical to pay for larger pipe, than to pay the penalty for the pressure drop in pipe and fittings for this case study.

Table 5.7.

ATM results for targeting heating medium flowrate for industrial case study (Scenario 2)

Temperature, T_t (°C)	Heat capacity flowrate, $F_{m,t}$ (kW/°C)	Interval Heat Load, ΔQ_t (kW)	Heat Load Cascade, Q_t (kW)	Heat capacity flowrate, F_{opt} (kW/°C)
90			0	0
		5067		
65			5067	123.67
		5274		
57	3		10341	211.15
		8218		
39	2		18560	277.11
		933		
30	1		19493	256.57

Table 5.8.

ATM results for targeting for HEN for industrial case study (Scenario 2)

Temp., T_i (°C)	HM Users	Heat load cascade, rdQ_i (kW)	Heating medium temperature, ST_i (°C)	Log mean temperature, T_{LGMi} (°C)	Hot area, $A_{t,h}$ (m ²)	Cold area, $A_{t,c}$ (m ²)	Total area, A_{HEN} (m ²)
90		19493.00	130.00				728.03
				43.27	58.56	78.07	
65		14425.58	111.71				
				40.95	64.40	105.46	
57	3	9151.52	92.68				
				29.47	139.44	247.20	
39	2	933.43	63.02				
				26.74	1745	17.45	
30	1	0	59.66				

Comparing with the conventional design (see Figure 1.1), where the heating medium is fed to all heat exchangers at 130 °C and returns at 95 °C, the TAC is determined as \$250,096 (see detailed results in Table 5.9). Thus, an annual saving of \$46,616 (corresponds to 19% reduction) can be achieved through the series design in Figure 5.7 (b). Similar to the earlier example, savings achieved

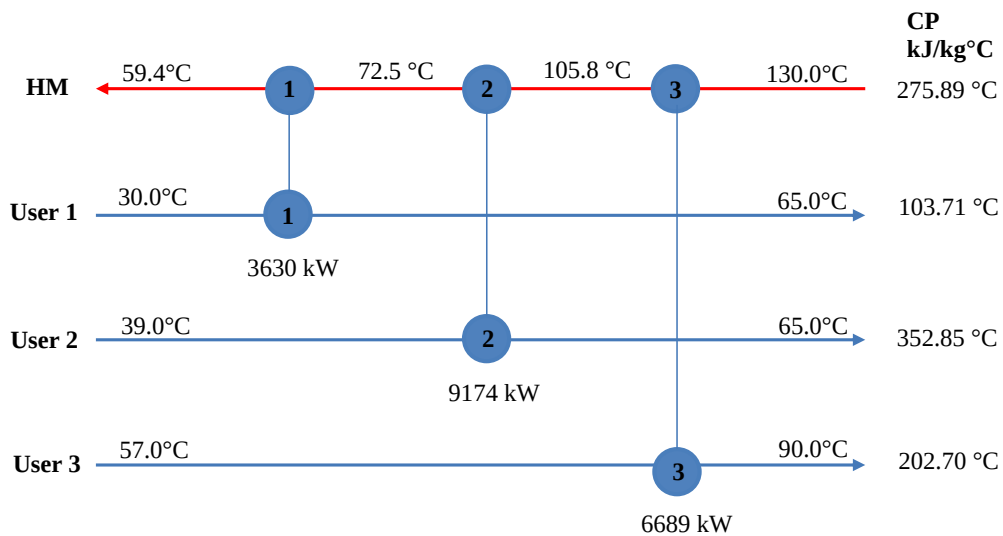
**Figure 5.6.** HEN for case study with lowest total area in grid diagram form

Table 5.9.

Comparison of key results and cost between ATM, detailed and conventional design for industrial case study

Description		ATM (Scenario 1)	ATM (Scenario 2)	Detailed design 1 (Figure 8)	Detailed design 2 (Figure 9)	Conventional design
HEN area (m ²)		732.28	728.03	768.52	819.93	483.86
WHRU area (m ²)		366.97	367.48	367.48	367.48	444.34
Pump shaft power (kW)		59.40	56.83	56.83	56.83	114.15
Heating medium flowrate (kW/°C)		275.89	277.11	277.11	277.11	556.94
Pipe and fittings losses (bar)		0.5	0.263	0.263	0.263	0.259
Velocity (m/s)		-	0.9144	0.9144	0.9144	0.9144
Pipe Diameter (m)		-	0.3540	0.3540	0.3540	0.4992
Annualised capital cost	HEN (\$/y)	80,127	79,777	83,113	87,350	59,654
	WHRU (\$/y)	44,131	44,185	44,185	44,185	52,323
	Pump (\$/y)	10,323	10,075	10,075	10,075	14,785
	Piping (\$/y)	-	2,127	2,127	2,127	3,344
Total annualised capital cost (\$/y)		134,581	136,164	139,500	143,737	130,106
Total utility cost (\$/y)		62,446	59,743	59,743	59,743	119,990
TAC (\$/y)		197,027	195,907	199,243	203,480	250,096

from utility cost due to series design configuration (as compared to the conventional parallel design) contributed to the largest portion of the savings (81%).

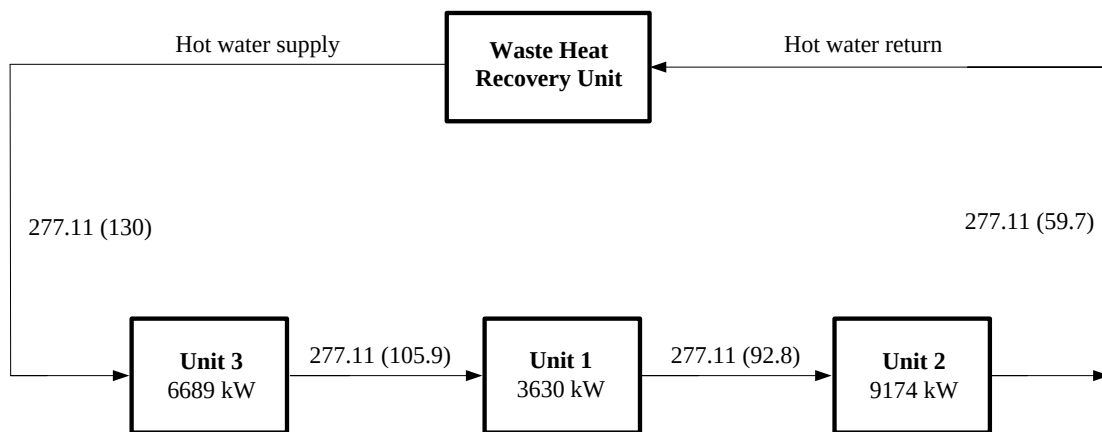
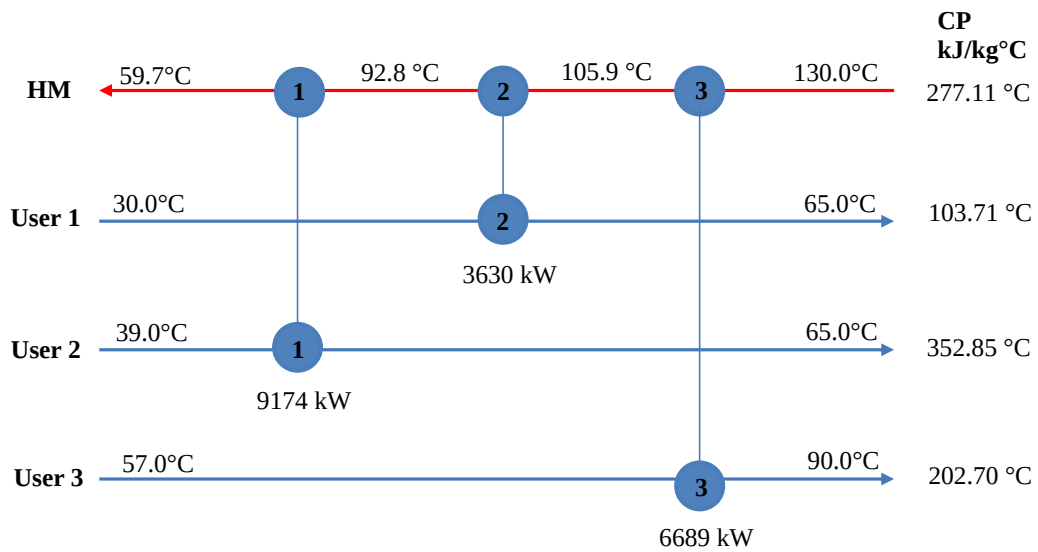


Figure 5.7. (a) Final HEN for case study in grid diagram form (b). Final configuration of heating utility network for case study in conventional PFD (heat capacity flowrate in kW/°C: temperature in °C, given in parenthesis)

Comparing with the conventional design (see Figure 1.1), where the heating medium is fed to all heat exchangers at 130 °C and returns at 95 °C, the TAC is determined as \$250,096 (see detailed results in Table 5.9). Thus, an annual saving of \$46,616 (corresponds to 19% reduction) can be

achieved through the series design in Figure 5.7 (b). Similar to the earlier example, savings achieved from utility cost due to series design configuration (as compared to the conventional parallel design) contributed to the largest portion of the savings (81%).

5.6 Summary

This section proposes an extended ATM which allows area and total cost targeting for a heating medium system, which is a special case of HEN. The proposed method allows the lowest TAC of the system that includes the regeneration cost of heating medium to be determined by identifying the optimum $\Delta T_{\min}$. The main advantage of the ATM as compared to other techniques is that, it allows the total annualised cost to be determined without the need for designing the final HEN. Besides, since it is algebraic in nature, it is carried out using optimisation software to obtain a rigorous target, which is not possible using graphical approach. The result obtained for the industrial case study in scenario 1 is evident of this, where the current solution is found to be superior than the result calculated through graphical approach. Besides, with the implementation of pipe and fitting losses in scenario 2, a better solution was achieved as the actual pressure loss was found to be lower than previously assumed. The application of process integration for the industrial case study shows reduction of TAC by 19% compared to the conventional design. In both hypothetical and industrial case study, the optimum solution found that the highest temperature of the heating medium is preferable when the inlet temperature of the heating medium is varied.

CHAPTER 6

SIMULTANEOUS OPTIMISATION OF HEATING UTILITY AND PHASE SEPARATION SYSTEMS FOR OIL AND GAS PLATFORM

6.1 Introduction

This chapter is arranged as follows. First, the problem statement for simultaneous optimisation of heating utility and phase separation systems is provided. Next, a revised automated targeting model (ATM) is presented to determine minimum utility targets for the entire systems, with a newly proposed algorithm in performing automated sorting of the heating medium temperatures in descending order; the latter enables the temperature intervals to be formed automatically, where the heating medium users may be located. In the following section, the design models for the separation units are presented. An industrial case study is next used to illustrate the newly proposed simultaneous optimisation approach. The case study is fairly complex due to the series arrangement of the crude oil heater, where the upstream outlet temperatures of light and heavy oil heaters effect the inlet temperature of crude oil heater (Figure 1.2). Due to this arrangement, the chemical properties of the crude oil inlet stream have to be calculated each time the upstream heaters outlet temperature is changed. In order to avoid large iterations, a localised search approach is adopted to solve the model.

6.2 Problem Statement

The formal problem for heating utility and phase separation systems design can be stated as follows. Given is a set of heating medium users (heat exchangers) $m \in M$, which are to be heated by a single source of heating medium (water or oil). The heating medium has a specified supply temperature, T_{HM} while its return temperature, T_{HMR} is set as a variable. For the heating medium user (heat exchanger) m , the inlet temperature is specified as $T_{in, m}$ while its outlet temperatures, $T_{out, m}$ are set as discrete variables with minimum inlet temperature being 10°C above $T_{in, m}$ while the maximum temperature being 10°C below T_{HM} . The heat load and heat capacity flowrates of user m varies according to their outlet temperatures. Apart from the heat exchanger network (HEN), the heating utility system also consists of pump and waste heat recovery unit (WHRU); hence their costs are to be included as part of the problem analysis.

For every outlet temperature of the heating medium users, a set of oil property data, e.g. density, viscosity, volumetric flowrate, etc. are made available (generated from process simulation software). These data are used for optimum sizing of the separators. The main objective is to synthesise an optimum heating utility and phase separation systems with lowest total annualised cost (TAC), lower than the conventional design in Figure 1.1). The overall optimisation process is summarised in Figure 6.1.

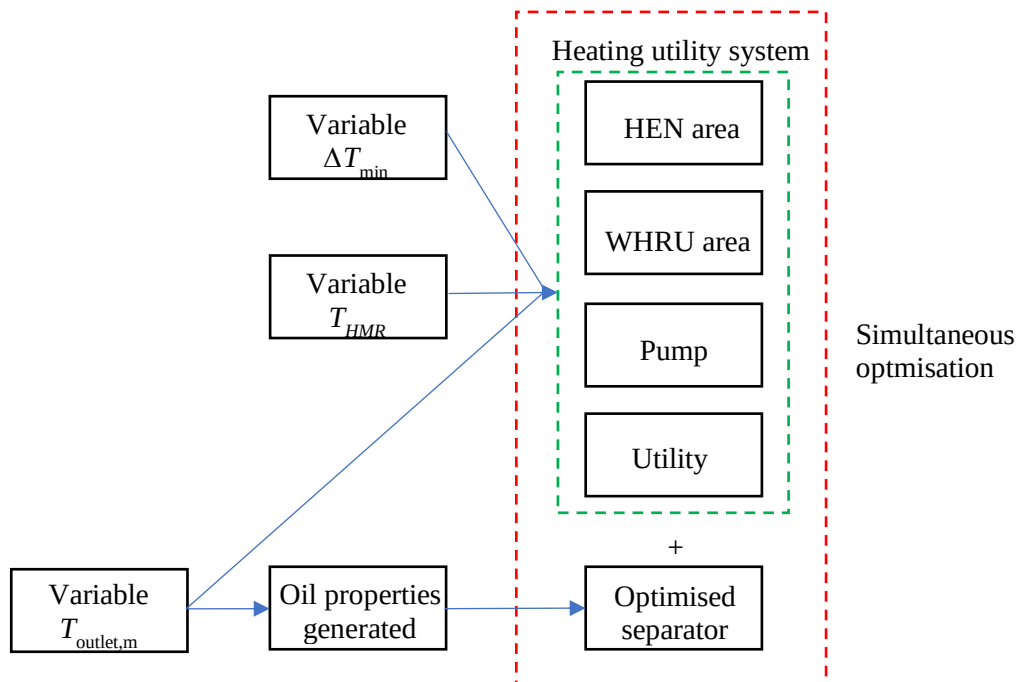


Figure 6.1. The overall simultaneous optimisation process for heating utility and phase separation system

6.3 Revised ATM with automated temperature intervals creation

The limitation of the ATM method proposed in Chapter 5 is that, it is not fully automated, as it requires user intervention in sorting the temperature levels, in order to locate the heating medium users within the corresponding intervals. When optimisation is to be performed for heating medium and phase separation systems simultaneously, the outlet temperatures of the heating medium users change according to the heat input to the individual heat exchangers. As such, the model becomes

cumbersome when user intervention is required whenever the outlet temperatures are changed. To overcome this, a new algorithm is proposed for generating temperature levels and to locate the heating medium user automatically into temperature intervals without user intervention.

In the first step, all given inlet and outlet temperatures are to be sorted from highest to lowest levels. In order to do so, the inlet and outlet stream temperatures are converted to common temperature abbreviations that are in arbitrary arrangement using Equations 6.1 and 6.2.

$$T_i = T_{inlet_i} \quad \text{for } i \leq M \quad \forall i \in I \quad (6.1)$$

$$T_i = T_{outlet_{(i-M)}} \quad \text{for } i > M \quad \forall i \in I \quad (6.2)$$

where M represents the total number of heating medium users and i represent the specified stream temperatures. Since the outlet temperatures are set as variables in this model, the optimisation operation will select one outlet temperature among the available options for the heating medium user, m . Equations 6.3 and 6.4 will ensure the preceding statement hold true, where σ is binary variables, while subscript n represent the available outlet temperatures options.

$$T_{outlet_m} = \sigma_{m,n} T_{outlet_{m,n}} \quad \forall m \in M \quad \forall n \in N \quad (6.3)$$

$$\sum_n \sigma_{m,n} = 1 \quad \forall m \in M \quad \forall n \in N \quad (6.4)$$

Note that the total number of heating medium users, M is related to the total number of specified stream temperatures, I by Equation 6.5. This is because each user will only have one set of specified inlet and outlet temperatures.

$$I = 2M \quad (6.5)$$

Note that, the value of I is also equivalent to the total number of temperature level, T for a given problem. The logic in arranging the highest to the lowest temperatures is derived from the notion that if there is I number of streams temperatures, the highest temperature can be bigger than other temperatures by $I-1$ times. Similarly, the second highest temperature will be bigger than other temperatures by $I-2$ times and so on. This relationship can be deduced to a set of equation shown below where t is the order of superiority (also equivalent to temperature level).

$$O_t = I - t \quad \forall t \in I \quad (6.6)$$

The next step is to compare each temperature T_i (apart from its own temperature) to one another using a mirror set of temperature T_k . This is done to determine how many times each temperature is bigger than the others. Binary value of ψ will produce the value of 1 when a temperature is bigger than the another.

$$\psi_{i,k} T_i \geq \psi_{i,k} T_k \quad \text{for } i \neq k \quad \forall i \quad \forall k \quad \epsilon I \quad (6.7)$$

$$(1 - \psi_{i,k}) T_i \leq \psi_{i,k} T_k \quad \text{for } i \neq k \quad \forall i \quad \forall k \quad \epsilon I \quad (6.8)$$

Note that the total summation of ψ value represents the total number of times all the temperatures can be larger than each other. Equation 6.9 shows how this can be calculated.

$$\sum_i \sum_k \psi_{i,k} = \frac{I(I-1)}{2} \quad \text{for } i \neq k \quad \forall i \quad \forall k \quad \epsilon I \quad (6.9)$$

The following step in sorting the stream temperature involves the comparison of the calculated total ψ value of each temperature to O through Equations 6.10-6.12.

$$\sum_k \psi_{i,k} \xi_{i,t} \leq O_t \quad \text{for } i \neq j \quad \forall i \quad \forall k \quad \forall t \quad \epsilon I \quad (6.10)$$

$$\sum_i \xi_{i,t} = 1 \quad \forall i \quad \forall t \quad \epsilon I \quad (6.11)$$

$$\sum_t \xi_{i,t} = 1 \quad \forall i \quad \forall t \quad \epsilon I \quad (6.12)$$

Temperature with the highest value will have ψ value of $I-1$, the second highest temperature will have ψ value of $I-2$ (see Figure 6.2), etc. Note that ξ is a binary variable introduced in these equations to relate the calculated O value of each temperature with ψ . Each ξ value can only relate one temperature to one level of superiority of t . The latter requirement is enforced by Equations 6.11-6.12.

Finally, Eq 6.13 will arrange the temperature levels according to descending order, i.e. from highest to lowest temperature, so that energy targeting may be carried out.

$$\sum_i \xi_{i,t} T_i = T_t \quad \forall i \quad \forall t \quad \epsilon I \quad (6.13)$$

Note that there is a possibility that one or more stream temperature could have the same value. These however does not affect the logic function of Eqs. 6.4-6.13 which are used to perform

the sorting of stream temperatures. Note however that there are some numerical complications during the calculation of the log mean temperature difference. The method to overcome this problem will be discussed in later section.

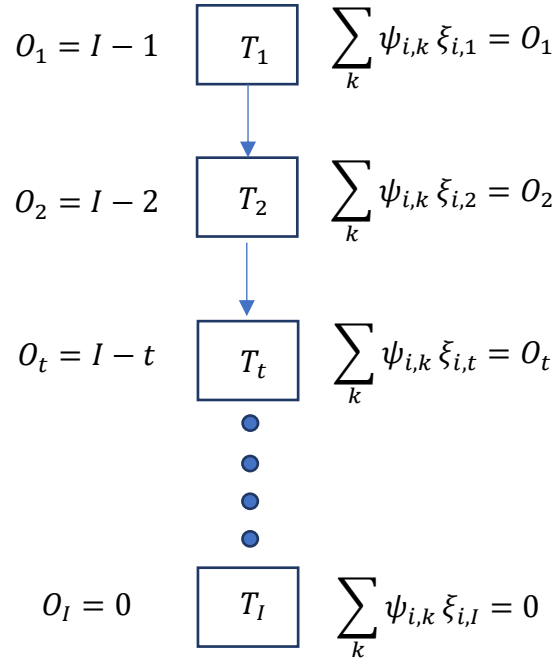


Figure 6.2. General framework for sorting temperature in descending order

In the following step, each process stream is to be located in its corresponding temperature intervals. In order to do so, the temperature interval is first generated using Equations 6.14.

$$T_{int_z} = T_t \quad \forall z \in Z \quad (6.14)$$

Note that the total number of intervals, Z may be related to the total number of streams temperatures, I as shown in Equation 6.15.

$$Z = I - 1 \quad (6.15)$$

The next step involves the comparison of the inlet and outlet temperatures of a stream with the interval temperature. For an outlet temperature to exist in an interval, its temperature has to be larger than or equal to that of the interval. This is tested using Equation 6.16, in which binary value δ will take a value of 1 for the equation to remain valid.

$$Toutlet_m \geq \delta_{m,z} Tint_z \quad \forall m \in M \quad \forall z \in Z \quad (6.16)$$

Note that from Equations 6.4 – 6.11, the total temperature levels are larger than the other temperatures in the intervals, X_i can be related to a value as shown in Equation 6.17.

$$X_i = \sum_k \psi_{i,k} \quad \forall i \quad \forall k \in I \quad (6.17)$$

This value is also equalling to the amount each outlet temperature can exists in the temperature intervals, $Tint_z$. Thus, the total amount for binary value δ of an outlet temperature, $Toutlet_m$ in total intervals, z is determined using Equation 6.18.

$$\sum_z \delta_{m,z} = X_{(m+M)} \quad \forall m \in M \quad \forall z \in Z \quad (6.18)$$

On the other hand, for an inlet temperature to exist in an interval, it has to be smaller than the interval temperature, $Tint_z$. Note that this is the reverse operation of the outlet temperature discussed above (Equation 6.16). Equations 6.19 is used to compare the inlet temperature of user m with the temperature intervals, $Tint_z$.

$$Tinlet_m \geq (1 - \tau_{m,z}) Tint_z \quad \forall m \in M \quad \forall z \in Z \quad (6.19)$$

Binary value τ in Equation 6.19 will take a value of 0 to ensure a valid equation. The total amount of binary τ for an inlet temperature, $Tinlet_m$ in intervals z , is determined using Equation 6.20.

$$\sum_z \tau_{m,z} = Z - X_m \quad \forall m \in M \quad \forall z \in Z \quad (6.20)$$

The value of X , i.e total value a temperature is larger than other temperature is deducted from Z in Equation 6.20 since we are now determining whether the inlet temperature, $Tinlet_m$ is lower than the interval temperature, $Tint_z$ in Equation 6.19.

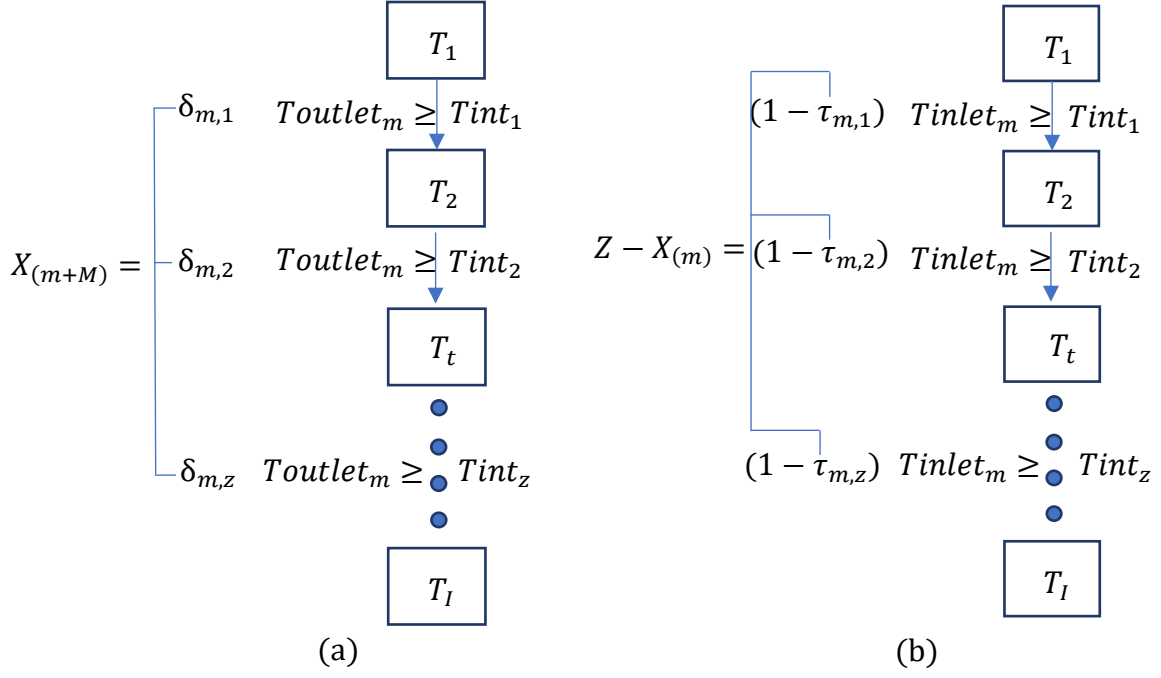


Figure 6.3. (a) Framework for determining outlet temperature in intervals, z (b) Framework for determining inlet temperature in intervals, z

Several modifications are required on the ATM equations proposed in Chapter 5 in order to automate the calculation when outlet temperature is varied. First, Equation 5.1 is modified to Equation 6.21 by adding binary τ and δ for each user m at intervals, z .

$$\Delta Q_t = \sum_m \tau_{m,z} \delta_{m,z} F_{m,t} (T_t - T_{t+1}) \quad t = 1, 2, 3 \dots T - 1 \quad \forall m \in M \quad \forall z \in Z \quad (6.21)$$

The product of binary τ and δ will determine whether the user m , with T_{inlet_m} and T_{outlet_m} temperature exist in a particular interval, z i.e. both binary value of τ and δ will require the value of 1 for a set of temperatures to exist in the interval.

Note that, the value of heat capacity flowrate of each user, F_m is no longer a constant since the outlet temperatures are variables. Due to this, the model will choose the appropriate value of F_m based on the outlet temperature that produces the lowest cost. The relationship between the outlet temperature and heat capacity flowrate of user m is provided by its binary variable, σ as shown in Equation 6.22.

$$F_m = \sigma_{m,n} F_{m,n} \quad \forall m \in M \quad \forall n \in N \quad (6.22)$$

Next, calculation of the log mean temperature has to be modified, since there is always a possibility that two or more users might have the same inlet or outlet temperatures, or both. For this scenario, the level temperature T_t and the heating medium temperature, ST_t will have the same value at levels t and $t+1$. Thus, the denominator of Equation 5.11 will produce the value of infinity. In order to overcome this problem, an arbitrary temperature difference between T_t and ST_t is chosen to avoid mathematical complication when there are two identical temperatures at level t and $t+1$.

Equations 6.23-6.24 ensure this operation, where the binary variable γ becomes 0 when there is no heat load, ΔQ at a particular temperature level. Temperature difference at level t , dt then becomes an arbitrary value AB .

$$\Delta Q_t \leq dQ_T \gamma_t \quad t = 1, 2, 3 \dots I - 1 \quad (6.23)$$

$$(T_t - ST_t) + AB(1 - \gamma_t) = dt_t \quad t = 1, 2, 3 \dots I - 1 \quad (6.24)$$

This new approach for calculating the temperature difference is redefined for Equation 5.11 as shown in Equation 6.25. Note that, the arbitrary value chosen will have no significance in calculating the area for heat transfer because the heat load, dt is always zero when there are two identical temperatures at level t and $t+1$, which will result in zero heat transfer area, according to Equations 5.12-5.13.

$$T_{LGMt} = \frac{[(dt_t) - (T_{t+1} - ST_{t+1})]}{LN[(dt_t) - (T_{t+1} - ST_{t+1})]} \quad t = 1, 2, 3 \dots T - 1 \quad (6.25)$$

Next, Equation 5.13 in which the area of the heat exchanger for the cold stream is determined. Similar to determining the heat load in Equation 6.21, binaries τ and δ are introduced in Equation 6.26 to determine whether a stream belongs to a particular interval. Both binaries of τ and δ requires to be one in order for area calculation to occur for a particular user m , at interval l .

$$A_{t,c} = \sum_m \frac{\tau_{m,l} \delta_{m,l} F_m (T_t - T_{t+1})}{h_m T_{LGMt}} \quad t = 1, 2, 3 \dots T - 1 \quad (6.26)$$

To ease the computation of the model, two constraints i.e. minimum ($F_{\min}$) and maximum heating medium flowrate ($F_{\max}$) are to be calculated first. The former is calculated by solving Equations 5.5 subject to constraints 6.1-6.22, 5.2-5.4 and setting $\Delta T_{\min}$ to the smallest allowable value. On the other hand, $F_{\max}$ is calculated using Equation 6.27.

$$F_{\max} = \frac{\text{MIN}\{dQ_{T,n}\}}{\Delta T_{HM}} \quad (6.27)$$

where the $dQ_{T,n}$ is the total heat load required by the entire system for a possible outlet temperature option, n , ΔT_{HM} is the smallest temperature difference between T_{HM} and T_{HMR} . Note that, Equation 6.27 is a revision of Equation 5.10 to take into account multiple heat load options. $F_{\min}$ and $F_{\max}$ are used as part of the constraints shown in Equation 5.9.

The flowrate target of the heating medium and the area target for HEN that are determined from these equations will be used to calculate the capital (for heat exchangers and pump) and utility costs of the thermal heating utility system. In order to do so, the WHRU and pump are first sized according to Appendix B and C. Then, detailed cost calculations for heat exchangers and pump as well as the utility cost is carried out according to Appendix D.

6.4 Separator sizing

Two types of separator are considered in this work, i.e. three phase and electrostatic separators. The three phase separator is first discussed. This kind of separator is used to separate gas, oil and produced water. The separation process allows the fluids to be processed easily and less costly than if they were processed together since gas and oil have different processing requirements. Besides, due to corrosive nature of the produced water, removing it allows cheaper material to be used for downstream equipment.

On the other hand, an electrostatic separator functions by producing electric field which enhances the coalescence of water droplets dispersed in crude oil. The calculation for the amount of power required for the electric field generation is beyond the scope of this study. The separator is sized based on gravity settling, with the assumption that the droplets size of water is large enough

after electric field treatment. The sizing model is similar to the three phase separator, however without allocated space for gas separation (since gas separation is performed in three phase separators).

The TAC2, that includes the capital cost of HEN, pump, WHRU, separators and utility cost is determined based on Equation 6.28. Note that the difference between TAC2 and TAC1 is that the former includes annualised capital cost of separators and excludes the piping cost.

$$TAC2 = C_{HEN} + C_{WHRU} + C_{PP} + C_U + C_{vss,n} \quad (6.28)$$

6.5 Solution strategy

The complexity of solving simultaneous optimisation of heating utility and phase separation systems lies largely on the inlet temperature of the crude oil heater. The latter depends on the outlet temperatures of the light and heavy oil heaters (located upstream of the crude oil heater – see Figure 1.2), as well as the mass flowrates of crude oil that is separated at their respective production separators. In order to solve the problem effectively, the light and heavy oil separators are sized based on the selected range of outlet temperatures for their respective heaters. The lower limit of the selected temperature depends on the minimum temperature difference set for each user, while the upper limit depends on the temperature of heating medium. The data required for the separator sizing i.e. fluid viscosity, density and volumetric flowrate are obtained from commercial process simulation software (Aspen HYSYS v8.8). The results of the separator sizing are then analysed to identify the appropriate range of the outlet temperatures that should be used; the latter will determine the inlet temperature of the crude oil heater. As will be shown in the result section, in contrary to common believe, heating the crude oil does not necessarily reduce the size of the separator. Thus, certain outlet temperatures of light and heavy oil heaters need not be considered in simultaneous optimisation of heating utility and phase separation systems.

Once the inlet temperature to the crude oil heater is determined, fluid data for electrostatic separator are obtained from process simulation for the selected range of outlet temperatures for crude oil heater. This allows the separator to be sized accordingly based on gravity settling. With information on the minimum cost of all three separators are now available, heat integration is carried

out with fixed outlet temperature for the light and heavy oil heaters, while the outlet temperature for crude oil heater is set as variables based on the previously chosen range. The heating utility system with the separator cost is then solved to identify the optimum outlet temperature of crude oil that produces the lowest cost. The process is repeated with a different choice of outlet temperature for light oil. After all the chosen light oil temperature has been completed, the optimum result obtained for variable light and crude oil heaters with fixed heavy oil outlet temperature are examined for change in gradient direction (i.e. from negative to positive or vice versa). In other words, when outlet temperature of light oil heater is increased, the result obtained for simultaneous optimisation should show consistent increment/decrement in cost. In the event that the result shows inconsistency, that region of light oil outlet temperatures i.e. temperatures that is just higher and lower of the temperature that caused change in gradient direction is further discretised and the process of retrieving data, solving for minimum cost for electrostatic separator, and heat integration is repeated with new set of light oil and crude oil heater outlet temperatures. Note that, since all the possible temperatures of crude oil heater are considered, no further discretisation is required. After all the light oil heater temperatures were considered and no change in the direction of the gradient was found, a new heavy oil heater outlet temperature is chosen. The process mentioned before for set of light oil and crude oil heater temperatures is repeated with new sets of temperatures. Once all the heavy oil outlet temperature is completed, the gradient is evaluated again for changes in direction with now all the temperatures are variables. If there is change in gradient direction, that region of heavy oil outlet temperature (temperatures just higher and lower) is further discretised and the process mentioned above is repeated. On the other hand, if there is no change in gradient direction observed, the optimum result is identified among the lowest cost. The overall solution strategy is summarised in Figure 6.4.

For the HEN, there are many possible options for the crude oil heater outlet temperatures to be chosen for which determines the optimum solution. This imposes computational difficulties to the model. In order to overcome this, minimum ($F_{\min}$) and maximum ($F_{\max}$) flowrate among all possible options has to be included as part of the constraint in the model.

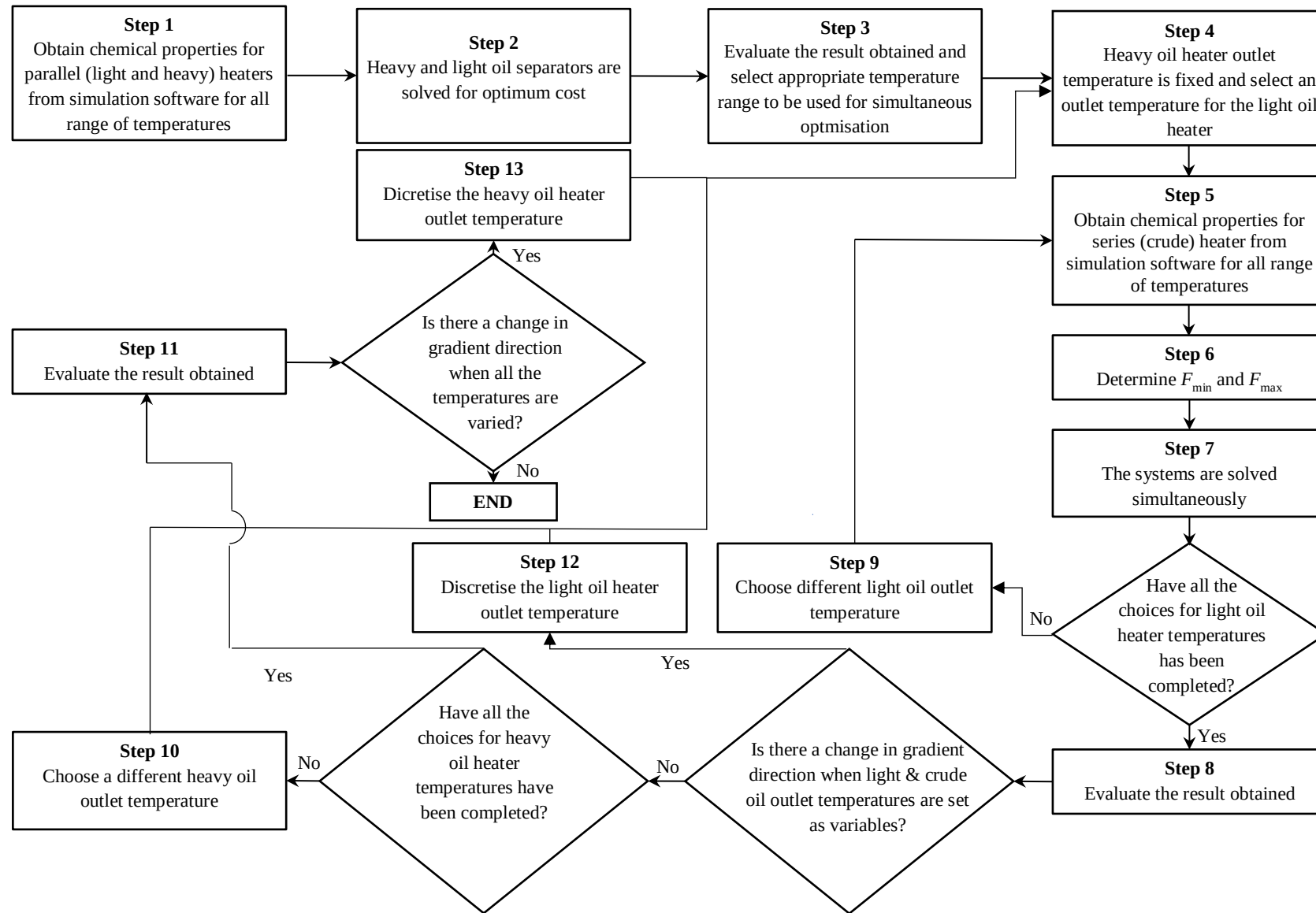


Figure 6.4. Solution strategy for simultaneous heating utility and phase separation systems

Minimum heating medium flowrate is determined by solving the HEN for minimum heating medium flowrate, with $\Delta T_{\min}$ set to lowest value. To obtain the maximum heating medium flowrate, the biggest possible heat load is to be divided by the smallest temperature deference between the heating medium inlet (T_{HM}) and return temperatures (T_{HMR}).

6.6 Case Study

An industrial case study (Figure 1.1 and 1.2) based on actual data from a *floating production storage and offloading* (FPSO) facility is used here to illustrate the application of the proposed method. The data for heating medium users (heat exchangers) is given in Table 6.1. The outlet temperature of all users is set as variables, with lowest operating temperature being 10°C above their inlet temperatures, while their maximum temperature being 10°C below the heating medium supply temperature, 130°C. Note that the maximum temperature should not exceed 130°C so to avoid Napthenate scale formation in the pipelines. As mentioned earlier, the heating medium user 3 (crude oil heater, Figure 1.2) is a unique case since its inlet temperature to the crude oil heat exchanger is dependent on the outlet temperature from heavy and light oil heater (users 1 and 2). Due to this series arrangement, new set of data is required for its inlet temperature, depending on the outlet temperatures of users 1 and 2. For instance, the minimum outlet temperature of user 3 can be determined as 42.6°C i.e. when users 1 and 2 are operated with the lowest outlet temperatures of 40°C and 49°C, respectively. On the other hand, the maximum outlet temperature of user 3 is set to 110°C, which is 10°C below the maximum outlet temperature at 120°C. The total heat load and heat capacity of all users hence become variables, due to the uncertain outlet temperature of the users. For simplicity, the film heat transfer coefficient is kept constant for all outlet temperature range.

Table 6.1.

Data for heating medium users

Heater	User, m	Inlet Temperature, $T_{\text{in}, m} (^{\circ}\text{C})$	Outlet Temperature, $T_{\text{out}, m} (^{\circ}\text{C})$	Cold streams film heat transfer coefficient, $h_m (\text{kWm}^2/^{\circ}\text{C})$
Heavy oil	1	30	40-120	2
Light oil	2	39	49-120	1
Crude oil	3	42.6-110	62.6-120	1.5

Data for the heating medium which is used to supply heat to the heating medium users are given in Table 6.2. The return temperature (T_{HMR}) of the heating medium (after heat exchange with the user) is also set as variable and is to be determined as part of the optimisation problem. The supply temperature (T_{HM}) is fixed at 130°C, unlike the case study in Section 5.5.

Table 6.2.

Physical properties for heating medium

Heating medium	Heating medium inlet temperature, T_{in} (°C)	Film heat transfer coefficient of heating medium, h_{HM} (kWm ² /°C)	Heating medium average specific heat capacity, CP_{HM} (kW/m ² °C)	Density (kg/m ³)
Water	130	2	4.326	984.12

Flue gas (obtained from gas turbine generator and gas turbine compressor) is used at the WHRU to raise the temperature of the heating medium to its desired value. Data of the flue gas is given in Table 6.3.

Table 6.3.

Flue gas data for industrial case study

Medium	Flue gas inlet temperature, $T_{in,g}$ (°C)	Flue gas inlet temperature, $T_{out,g}$ (°C)
Flue gas	433	130

As part of the optimisation study, capital cost of pump and WHRU, as well as the operating cost of pump are considered. Data required for pump and equipment costing calculations are given in Table 6.4. Note that, unlike the industrial case study in Section 5.5, piping and fitting loss calculation as well as pipe costing is not part of the scope of this industrial study. The piping and fittings losses, $hl_{P\&F}$ is assumed as 0.5 bar.

Table 6.4.

Data for pump power and equipment costing calculation

Parameters	Value
Enhanced overall heat transfer, U_e (kW/m ² °C)	0.302
Head loss at equipment and differential height required, d	4.5 bar
Head loss at piping and fittings, $hl_{P\&F}$	0.5 bar
Pump efficiency, η	60%
i	10%
n	25 years
a	59,844
b	748
c	1
WHRU, C_B	\$ 2.3339x10 ⁵
WHRU, D_B	200 m ²
WHRU, M	0.89
Pump, C_B	\$ 2.1247x10 ⁴
Pump, D_B	4 kW
Pump, M	0.55
Duration, W	8760 hr
Electrical tariff, G	0.12 \$/hr

For separator sizing, new set of stream properties are required for different outlet temperature of the heating medium user. The data is generated using simulation software, i.e. Aspen HYSYS v8.8. The simulation flowsheet for the industrial case study is shown in Figure 6.5. The design basis and assumptions for the simulation are as follows:

- Peng-Robinson is chosen as the fluid package
- Outlet sediment and water (S&W) in oil stream is assumed at 10% (volume in oil stream) for heavy oil and light oil separators.
- Outlet S&W in oil stream is assumed at 0.5% (volume in oil stream) for electrostatic separator.
- Water cut for oil stream entering the electrostatic separator shall not exceed 20% by volume.

Note that a logical unit operation “Adjust” is used to set the wash water flowrate based on different concentration of crude oil entering Mixer 4. The wash water is added to reduce the concentration of salt in the crude oil. Other information that the adjust function takes into account for calculating the required wash water flowrate is given in Table 6.5.

Table 6.5.

Specification for wash water calculation

Item	Value
S&W volume % for outlet stream to crude oil storage	0.5
Inlet salt content for stream S21, Slt_{in}	4.4 kg NaCl per barrel of brine S21
Outlet salt content for outlet stream to crude oil storage, Slt_{out}	3.165 kg NaCl per barrel of brine for outlet stream to crude oil storage
Wash water salt content, Slt_{ww}	0.95 kg NaCl per barrel of wash water
Mixing efficiency between wash water and stream S21, E	80%

The outlet salt content specified is equivalent to 35 pounds per thousand barrels (PTB) of crude oil, which is within the range of 10-250 PTB transportations specifications for crude oil to refineries (Abdel-Aal et al., 2003). The data for wash water salt content is obtained from Manning and Thompson (1995).

The amount of dilution water is calculated based on Equations 6.29-6.31 for single stage desalting (Manning and Thompson, 1995).

$$DW = AC(Slt_{in}BC - Z)/E(Z - Slt_{ww}BC) \quad (6.29)$$

$$AC = VF_1(1 - VF_1) \quad (6.30)$$

$$BC = VF_2(1 - VF_2) \quad (6.31)$$

where DW is dilution water flowrate (bbl per bbl of clean crude)

AC is bbl brine in S21 per bbl of clean crude

BC is bbl brine outlet to storage tank per bbl of clean crude

VF_1 volume fraction of brine in crude stream for S21

VF_2 volume fraction of brine in crude stream for outlet stream to crude oil storage

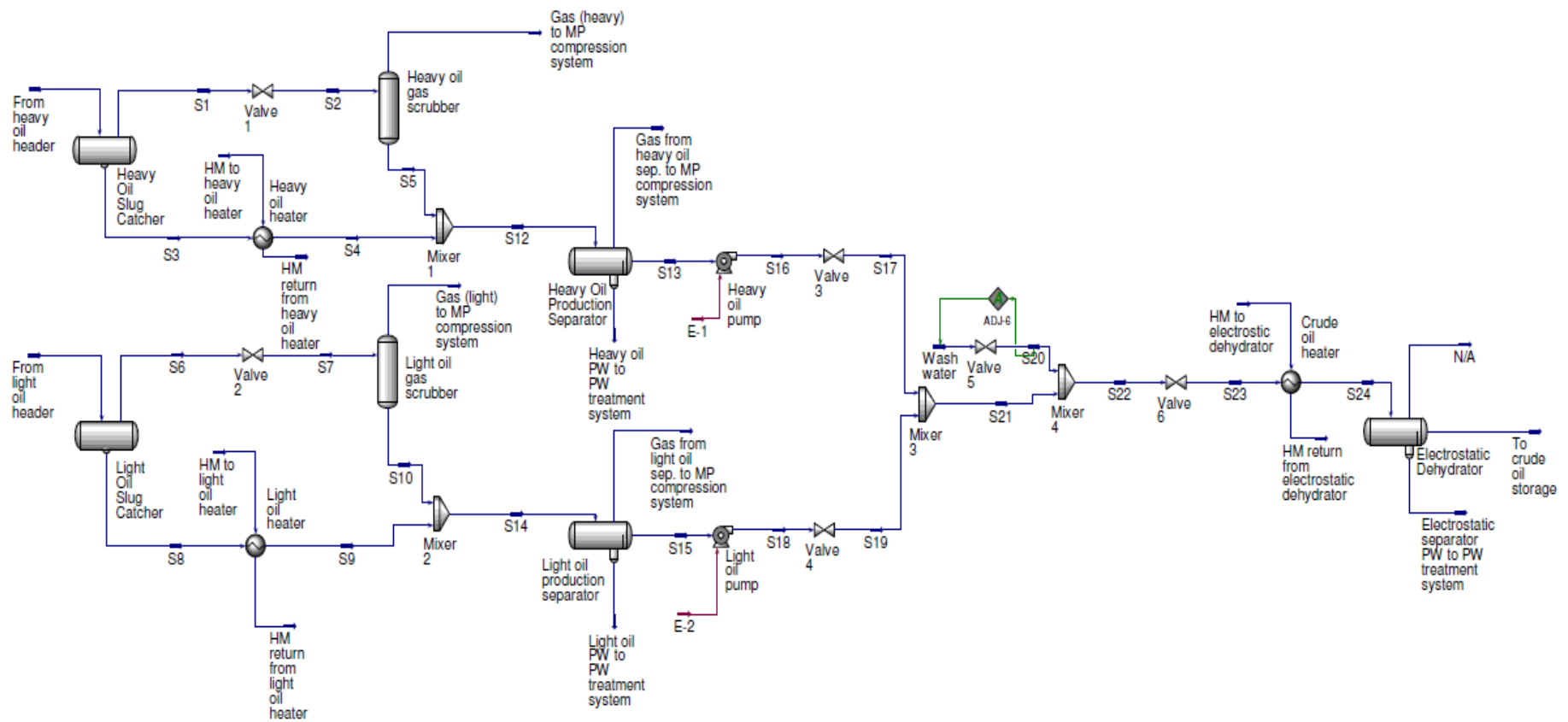


Figure 6.5: Process simulation for phase separation system with Aspen HYSYS

These equations are added to the adjust function in Aspen HYSYS to calculate the required wash water flowrate.

Process specifications for each separator are given in Table 6.6. Note that, for the oil and water residence time (column 8 and 9), American Petroleum (API) specification 12J (1991) is used as reference (Table 6.7). The specification allows equal retention times for both the oil and water phases.

Table 6.6.

Carry over, droplet size and design pressure specification for separators

Separator	Droplet size (μ)			Design pressure (kPa)	Souders Brown coefficient (m/s)	Retention time (min)	
	Oil in gas	Oil in water	Water in oil			Oil	Water
Heavy oil	≤ 300	≤ 300	≤ 600	4,500	0.2438*	5	5
Light oil	≤ 300	≤ 300	≤ 300	3,000	0.2438*	3	3
Electrostatic	N/A	≤ 300	≤ 250	4,500	N/A	5	5

*Data from Manning and Thompson (1995)

Table 6.7.

API 12J specification for retention time

Oil Gravity ($^{\circ}$ API)	Temperature ($^{\circ}$ C)	Typical Retention time (min)
Above 35	-	3 to 5
Below 35	37.8	5 to 10

Other specification for the separation sizing that is applicable to all separators are given in Table 6.8.

Table 6.8.

Separation sizing specification

Item		Value
Steel density*		7849 kg/m ³
Corrosion allowance, CA^*		0.0032 mm
Maximum allowable stress, S^*		1,034.21 barg
Fractional weld efficiency, E^*		1.0
CE index**		704.1
Maximum diameter, L	3 phase separator	1.9 m
	2 phase separator	3.25 m
Maximum length		13 m

*Data from Seider et al. (2009)

**CE Index based on Oct 2014 (Economic Indicators, 2015)

Based on the above information, optimisation is first performed for the light and heavy oil separators. The required physical property data for the oil are obtained from Aspen HYSYS for every 1°C between 40-120°C for heavy oil separator, and between 49-120°C for light oil separator. Then, the model is solved with the constraints given in Equations F1-F32 and F34-F36, with objective to minimise the annualised capital cos of separator via Equation F37. Result of the optimisation of the light and heavy oil separators are shown in Figures 6.6 and 6.7.

Based on the optimisation results in Figures 6.6 and 6.7, two different trends are observed for heavy and light oil separators. For the latter, its optimum cost increases linearly as its outlet temperature is rising. On the other hand, the optimum cost of the heavy oil separator reduces initially, until the outlet temperature reaches 48°C before it increases gradually. For both cases, the separator sizing is controlled by liquid capacity and not by gas flowrate. This outcome is expected as most gases are removed at heavy and light oil slug catcher. The two different results obtained for light and heavy oil separators can be attributed to the liquid residence time and liquid droplet velocity. Note that the increase of temperature leads to reduced viscosity and density of the crude oil. These two changes in properties have opposite effect to the size of the separator. Reduced viscosity increases the droplet velocity of the fluid, thus reduces the time required for the liquid droplet (water droplet in oil) to settle.

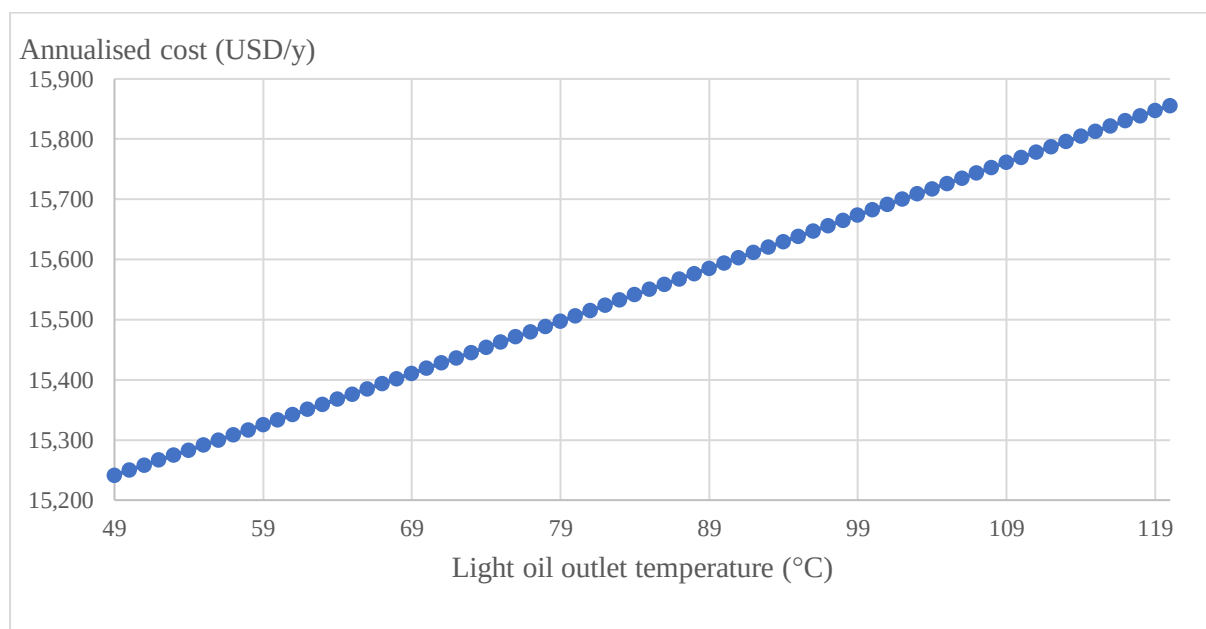


Figure 6.6. Optimum light oil separator cost against its heater outlet temperature

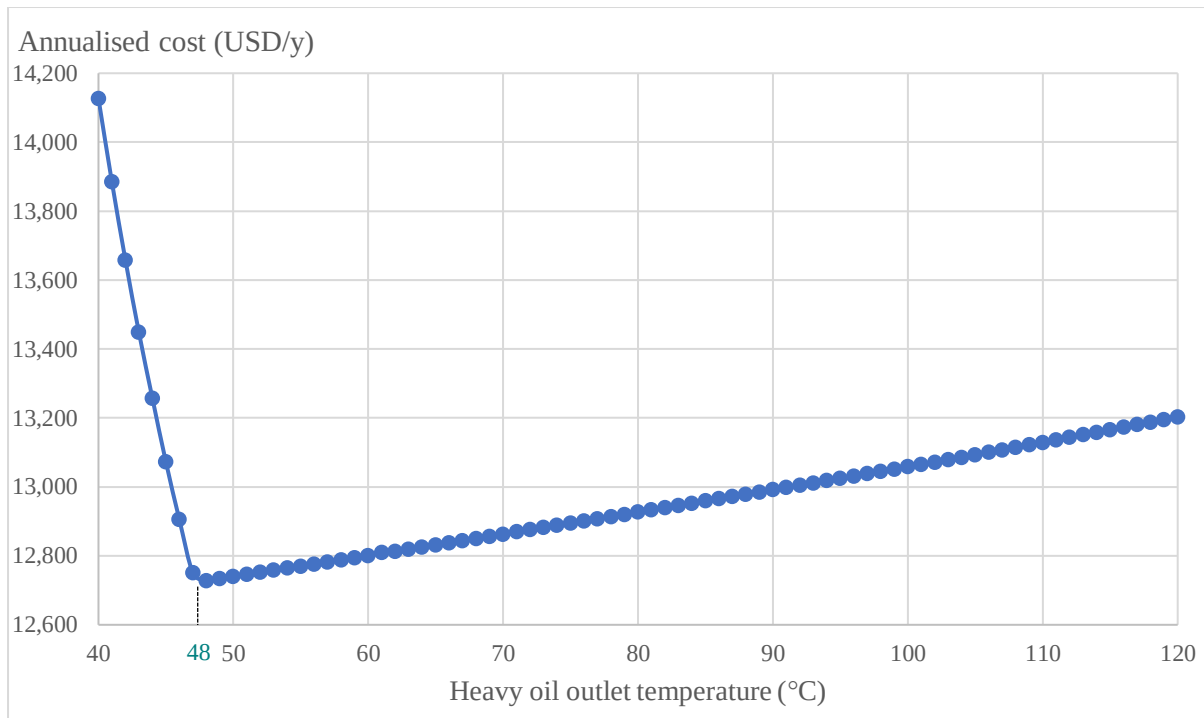


Figure 6.7. Optimum heavy oil separator cost against its heater outlet temperature

On the other side, reduced density increases the volumetric flowrate, thus larger vessel volume is required to meet the specified residence time. This is useful in explaining the costing trend of the heavy oil separator in Figure 6.7. When the separator is operated between 40–48°C, the size of the separator is controlled by the liquid droplet velocity. Thus, the benefits from viscosity reduction outweighs the impact of increment in volumetric flow. From 49°C to 120°C, the size of the separator is controlled by the liquid residence time, and hence the increased volumetric flow outweighs the benefits brought by the increase of liquid droplet velocity. The same situation applies for light oil separator as well, where the residence time is the controlling factor for the sizing of the separator for all ranges of the temperatures that were considered.

Based on Figure 6.4, step 3 is now applied whereby appropriate outlet temperature range is chosen for heavy and light oil heaters, in order to reduce the numbers of iteration. It is known that the increase of crude oil temperature will lead to overall increased heating requirements of the heating utility system. Consequently, the cost of heating utility system increases as larger heat exchanger area and higher heating medium flowrate will be required to achieve the required enthalpy load. Thus, it is

clear that in order to derive any benefit from providing more heat to the crude oil, there must be a reduction in the cost of the separator. Due to this reason, for heavy oil heater, only four temperatures between 40-48°C are considered i.e. 40°C, 43°C, 46°C, 48°C. For the light oil separator, there should be no addition of heat to the lowest outlet temperature that was specified, since it will lead to increased costs for both heating utility system and light oil separator. However, there is a possible scenario in which by providing more heat to the light oil heater, the total cost of entire system may be reduced due to temperature increase in the inlet stream to the crude oil heater. Hence, crude oil entering the final heater at higher temperature will require less heat to achieve the target outlet temperature. In order to study this scenario six temperature points between 49-120°C i.e. 49°C, 63°C, 77°C, 91°C, 105°C, 120°C are chosen for simultaneous heating utility and phase separation system optimisation.

To illustrate the iterative procedure of the model a single case is used as an example. First, the outlet temperatures 40°C and 49°C are chosen for heavy and light oil separators. Based on these outlet temperatures, the inlet temperatures to crude oil heater is determined as 42.6°C. Based on this inlet temperatures, the minimum outlet temperature of the crude oil heater is determined as 52.6°C, while the maximum temperature as 119.6°C (~ 120°C). Between these values, each outlet temperature was discretised for every 1°C and the process data i.e. volumetric flowrate, viscosity and density are obtained from the simulation model (Figure 6.5). From the process data, the electrostatic dehydrator is sized using Equations F3-F4, F6-F16, and F23-F36, with the objective is to minimise its total annualised cost, following Equation F37. Figure 6.8 shows the optimisation result for the electrostatic dehydrator. The result obtained shows similar trend as the heavy oil separator. Note that there is no viable solution between 52.6°C (minimum outlet temperature) to 70.6°C, as the calculated separator length is beyond the maximum value of 13 m (see Table 6.8).

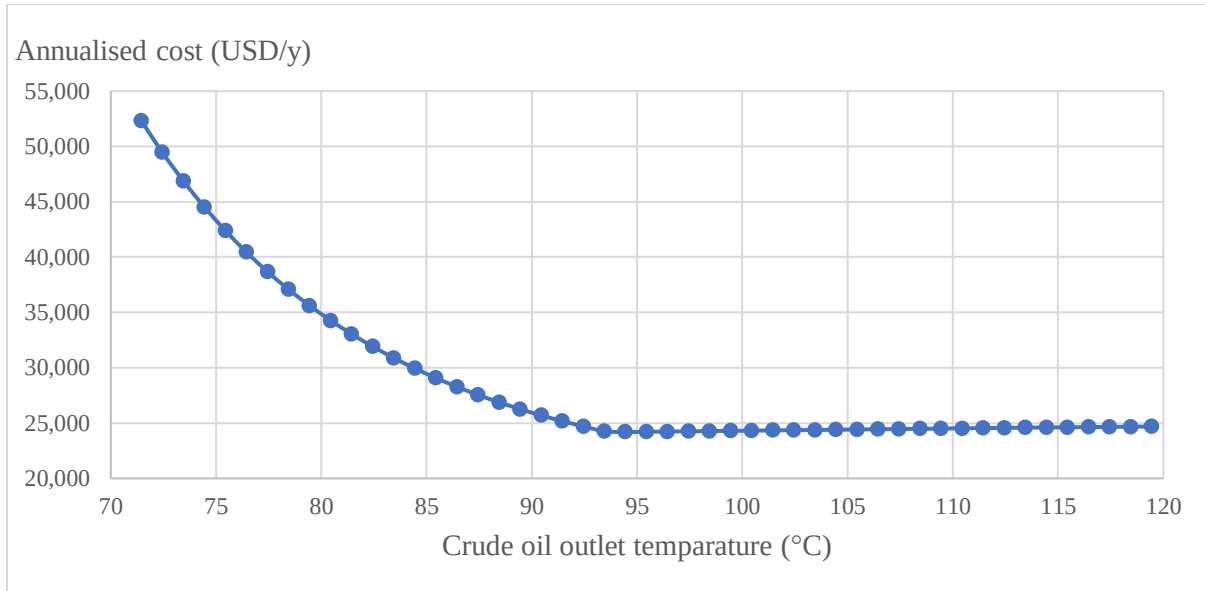


Figure 6.8. Optimum electrostatic dehydrator annualised cost against its heater outlet temperature

In the next step, heat integration is carried out based on the outlet temperatures of the three heaters. All possible outlet temperatures for crude oil temperature shown in Figure 6.8 will be considered, while the heavy and light oil separator outlet temperatures are set to 40°C and 49°C respectively. First, the $F_{\min}$ value is determined as 107.5 kW/°C based on Equations 6.1-6.22 and 5.2-5.4, with the optimisation function set to minimum heating medium flowrate, following Equation 5.5, with $\Delta T_{\min}$ set to 10°C. Next, the value of $F_{\max}$ is determined as 18815.38 kW/°C using Equation 5.10, with ΔT_{HM} set to 1°C. These two values are utilised in Equation 5.9. The model is then solved for minimum TAC2 (Eq. 6.28), subject to Equations 6.1-6.26, 5.2-5.9, 5.12, 5.14-5.18, B1-B8, C1 and D1-C5. The optimum solution is found to be \$182,807/y with the optimum outlet temperatures for the crude oil heater at 76.61°C. The process is then repeated for all possible combinations of outlet temperature of light oil (i.e. 49, 63, 71, 91, 105 and 120°C) and heavy oil separators (i.e. 40, 43, 46, 48°C) with the results summarised in Table 6.9. The table shows that when the heavy oil heater outlet temperatures are fixed, the optimised cost of the two systems increases when light oil heater temperatures increase. Since there is no change in gradient direction, there was no need for further discretisation for the light oil heater outlet temperatures. Similarly, when the heavy oil temperatures are increased a clear trend was observed where the optimum cost (highlighted) of the systems

increases as well. Due to no change in gradient direction, there was no requirement for further discretisation of heavy oil heater outlet temperature. Note that, the trend of crude oil outlet temperatures on the TAC is not a concern for further discretisation since all the possible outlet temperatures are considered in the analysis. Based on total of 948 different combinations of outlet temperatures that were considered, the final results show that the optimum solution (TAC2) as \$182,807/y. The corresponding heat and mass balances from Aspen HYSYS simulation of the optimum case are given in Appendix H.

Table 6.9.

Optimum result for different combination of heavy oil, light oil and crude oil outlet temperatures

Heavy oil heater outlet temperature (°C)	Light oil heater outlet temperature (°C)	Crude oil heater outlet temperature (°C)	Optimum Cost (USD/y)	Remarks
40	49	52.61-119.61	182,807	52.61-70.61°C *
	63	61.41-119.41	218,910	62.41-71.41°C*
	77	70.28-119.28	258,622	70.28-72.28°C *
	91	79.23-119.23	305,752	
	105	88.24-119.24	369,337	
	120	97.96-119.96	487,955	
43	49	53.30-119.30	183,379	53.30-70.30°C*
	63	62.09-119.09	219,220	62.09-72.09°C*
	77	70.95-119.95	258,689	70.95-72.95°C*
	91	79.88-119.88	304,368	
	105	88.88-119.88	369,722	
	120	98.60-119.60	488,795	
46	49	53.99-119.99	184,274	53.99-70.99°C*
	63	62.76-119.76	219,842	62.76-71.76°C*
	77	71.61-119.61	259,092	71.61-72.61°C*
	91	80.54-119.54	304,673	
	105	89.53-119.53	370,409	
	120	99.24-119.24	489,736	
48	49	54.45-119.45	185,129	54.55-70.45°C*
	63	63.13-119.13	220,532	63.13-72.13°C *
	77	72.06-119.06	259,613	72.06-73.06°C*
	91	80.98-119.98	305,177	
	105	89.96-119.96	371,129	
	120	99.66-119.66	490,629	

*No electrostatic separator available between the specified temperature that meets separator length below 13 m limit

Table 6.10.

Heaters outlet temperatures and heat loads for optimum solution for outlet temperatures of heavy (40°C) and light oil separators (49°C)

Unit, m	Inlet Temperature, $T_{in,m}$ (°C)	Outlet Temperature, $T_{out,m}$ (°C)	Cold streams film heat transfer coefficient, h_m (kWm ² /°C)	Load, Q_m (kW)	Heat capacity flowrate, F_m (kW/°C)
1	30	40	2	1,033	103.28
2	39	49	1	3,427	342.66
3	42.6	76.6	1.5	6,097	179.33
Total				10,577	625.27

The relationship of different outlet temperatures to TAC is examined by running the model as a single case without binary option, σ . Note that in this case, the possible outlet temperature options, n will be 1 since the model is carried one by one. Using this approach TAC for each different combination of outlet temperatures can be obtained individually. The results are summarised in Figures 6.9 – 6.12.

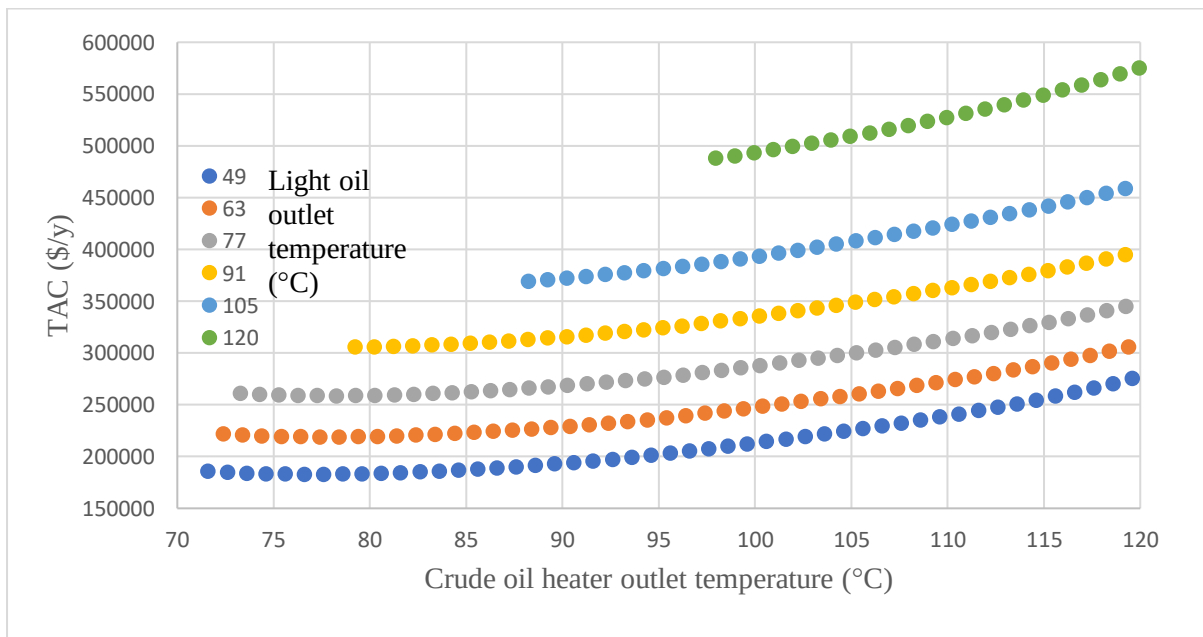


Figure 6.9. TAC vs light oil and crude oil heater outlet temperatures for heavy oil heater outlet temperature of 40°C

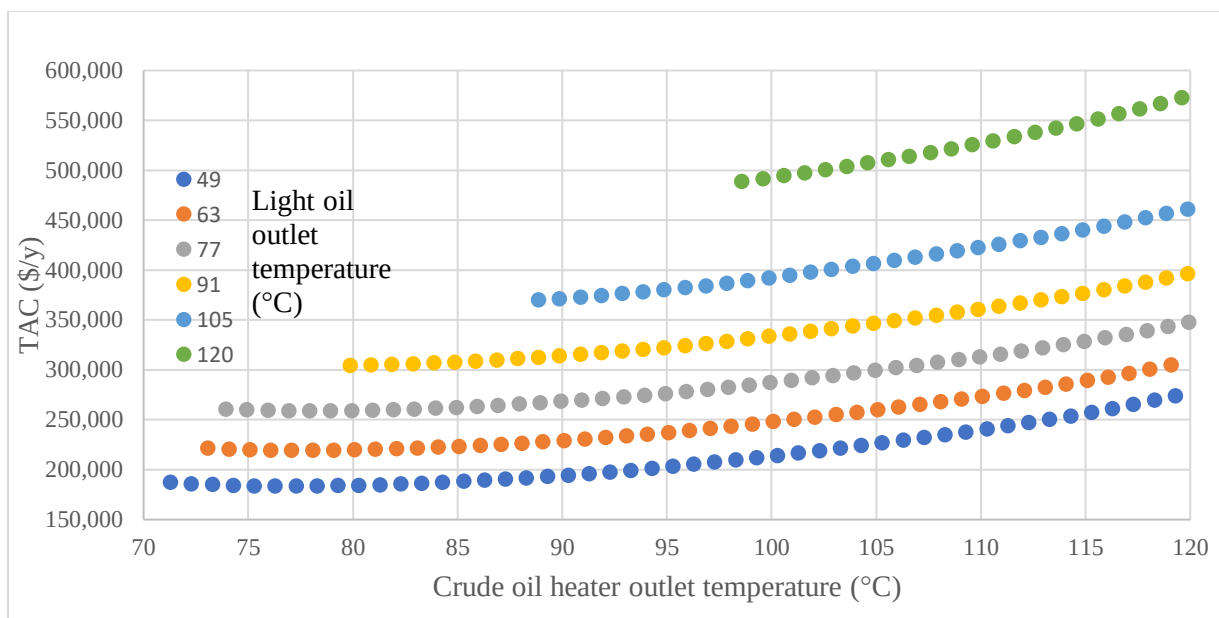


Figure 6.10. TAC vs light oil and crude oil heater outlet temperatures for heavy oil heater outlet temperature of 43°C

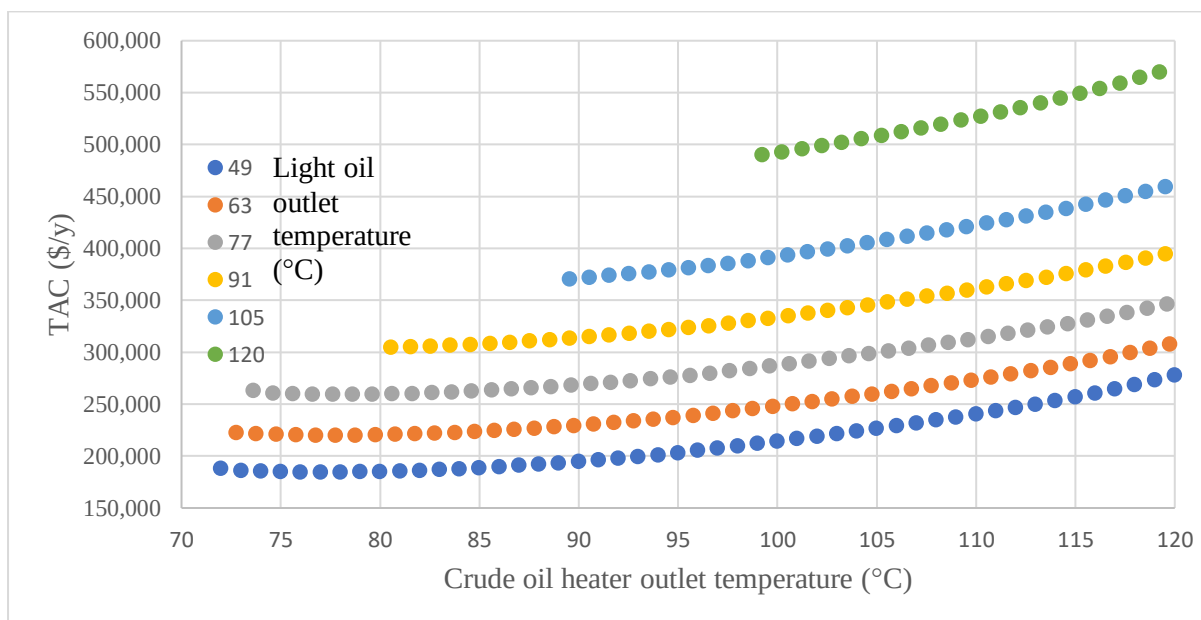


Figure 6.11. TAC vs light oil and crude oil heater outlet temperatures for heavy oil heater outlet temperature of 46°C

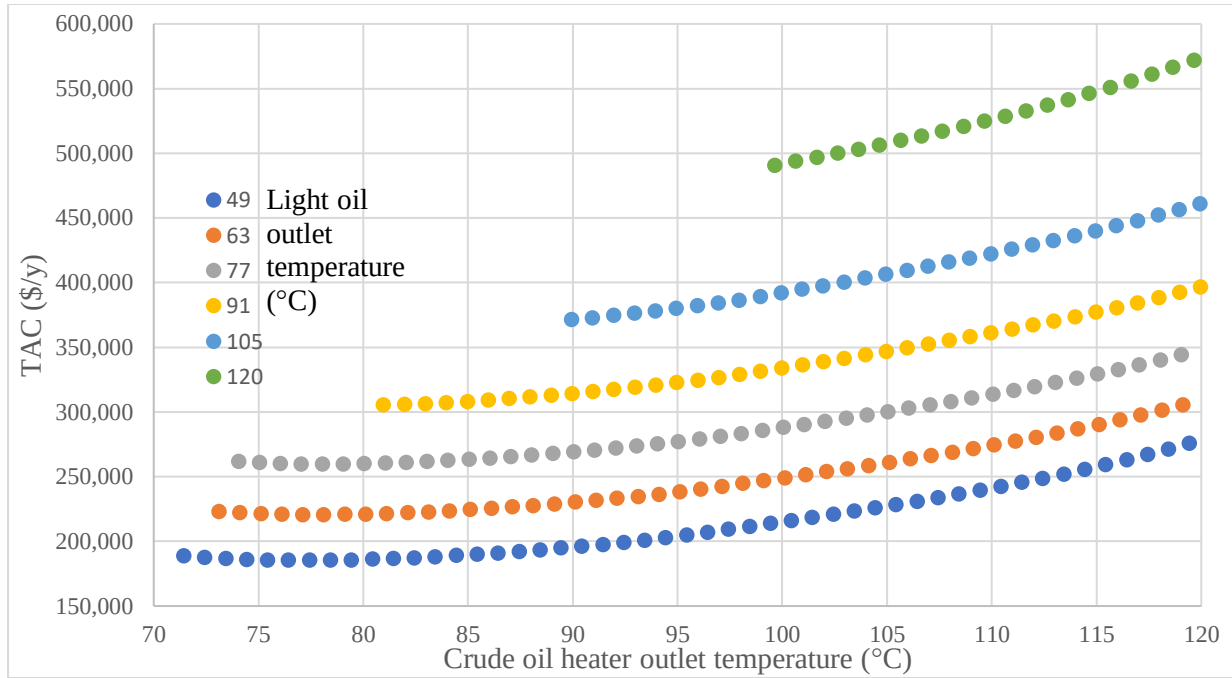


Figure 6.12. TAC vs light oil and crude oil heater outlet temperatures for heavy oil heater outlet temperature of 48°C

Note that from Figures 6.9– 6.12, it is observed that the outlet temperature for crude oil heater decreases with the increase of heavy and light oil heater outlet temperatures. This is due to the increase of inlet temperature of crude oil heater (as a results of higher outlet temperatures of heavy and light oil heaters located upstream to crude oil heater), thus eliminating the lower outlet temperatures as viable options. In all figures, the first three lowest outlet temperatures (49°C, 63°C and 77°C) of light oil heater show reduction in TAC initially when crude oil outlet temperatures are increased. This is because the cost of crude oil separator reduces at a faster rate at low outlet temperatures (see Figure 6.8), thus compensating the increase in the cost of HEN which requires to meet higher enthalpy demand. After initial reduction, the TAC increases with the increase of crude oil outlet temperature. As shown in Figure 6.8, at high outlet temperatures, heating crude oil has an inverse effect of separator cost. As explained earlier, this is due to size of the separator that is controlled by the residence time.

At this end, it is meaningful to compare the result obtained from simultaneous optimisation with two other scenarios. In the first scenario, the conventional design is used where all heat

exchangers receive the heating medium at the supply temperature in a parallel configuration, as shown in Figure 1.1. Data of the heating medium users is given in Table 6.11. Note that there is slight difference for user 3 data as compared to the one used for industrial case study for Chapters 4 and 5. This is due to simplification of the flowsheet shown in Figure 6.5 to ease process simulation when outlet temperatures of heaters were changed. The supply and the return temperatures of the heating medium are set to 130°C and 95°C, respectively and other chemical properties are identical to those in Table 6.2. Other data for the case study, i.e. flue gas, pump power and equipment costing remain identical as those in Tables 6.3 and 6.4. The result for costing calculation is summarised in column 2 of Table 6.12.

Table 6.11.

Cold stream data for scenarios 1 and 2

Unit, m	Inlet Temperature, $T_{in, m}$ (°C)	Outlet Temperature, $T_{out, m}$ (°C)	Cold streams film heat transfer coefficient, h_m (kWm ² /°C)	Load, Q_m (kW)	Heat capacity flowrate, F_m (kW/°C)
1	30	65	2	3,630	103.71
2	39	65	1	9,174	352.85
3	58.4	90	1.5	5,816	184.04
Total				18,620	640.60

In scenario 2, only heat integration is carried out for the heating utility system without simultaneous optimisation of the phase separation system. All data used for this scenario remain identical as those in scenario 1, while the return temperature of the heating medium is allowed to vary as part of the optimisation study. Note that, the HEN configuration may be different from the conventional design (Figure 1.1). The model is solved for optimum TAC2, subject to Eqs. 6.1-6.26, 5.2-5.8, 5.12, 5.14-5.18, B1-B8, C1 and D1-D5 without the binary σ . The result is summarised in column 3 of Table 6.12. As shown, the optimum TAC is determined as \$243,913/y, with ΔT_{min} located at 23.6°C and heating medium return temperature of 59.0°C.

For simultaneous optimisation approach, the optimum cost for heating utility system is determined as \$182,807/y, with $\Delta T_{\min}$ located at 22.7°C and heating medium return temperature at 55.1°C. Based on this information, the HEN is constructed using the pinch design method (Linnhoff et al., 1982) and design configuration is shown in Figures 6.13 and 6.14. Note that the discrepancies between the targeted and actual HEN area is within 10% accuracy for scenario 2 (0.8%) and simultaneous optimisation (5.4%).

As shown in Table 6.12, performing heat integration alone (Scenario 2) reduces the overall TAC from 306,313 \$/y to 247,221 \$/y (detailed design), or a reduction of 19.3%. Bulk of the savings (81%) is achieved from pumping cost, due to smaller heating medium flowrate required to meet the energy demand. Note that, performing heat integration on heating utility system does change the cost of separator since the outlet temperature of the heaters (inlet temperature to separators) are fixed. By performing simultaneous optimisation of heating utility and phase separation systems, the overall TAC of the two systems are significantly reduced to 183,040 \$/y, equivalent to 40.2% reduction, as compared to the conventional design. Based on these results, it may be concluded that using more energy at heating utility system in conventional design (18,620 kW, Table 6.10) was disadvantageous, as the cost of separator did not reduce significantly (Table 6.12, row 14). With simultaneous optimisation approach, lower heat load (10,557 kW, Table 6.9) was supplied to the heating utility system. Although the total cost of separator increased by 26% (i.e. to 69,318 \$/y from 54,970 \$/y in conventional design), large savings of 55% was achieved at the heating utility system (i.e. to 251,343 from 113,722 in conventional design), thus reducing the overall cost of both systems. As with Scenario 2, the biggest contributor to the overall savings is still the utility cost which contributed 69% of the overall savings.

Table 6.12.

Comparison of key results and cost between conventional design (Scenario 1), ATM supertargeting (Scenario 2) and simultaneous optimisation

Description	Conventional design (Scenario 1)	ATM Supertargeting (Scenario 2)	Detailed design (Scenario 2, Figure 6.13)	Simultaneous optimisation	Detailed design (simultaneous optimisation, Figure 6.14)
HEN area (m ²)	460.99	697.24	737.38	358.09	360.93
WHRU area (m ²)	145.34	349.99	349.99	195.12	195.12
Pump shaft power (kW)	119.92	56.46	56.46	30.36	30.36
Heating medium flowrate (kW°C ⁻¹)	556.94	262.24	262.24	140.98	140.98
HEN (\$/y)	57,770	77,239	80,547	49,289	49,523
WHRU (\$/y)	52,323	42,309	42,309	25,153	25,153
Annualised Pump (\$/y)	15,192	10,039	10,039	7,136	7,136
capital cost Heavy oil separator (\$/y)	12,831	12,831	12,831	14,127	14,127
Light oil separator (\$/y)	15,376	15,376	15,376	15,241	15,241
Electrostatic separator (\$/y)	26,762	26,762	26,762	39,950	39,950
Annualised capital cost for HEN (\$/y)	125,284	129,587	132,896	81,578	81,812
Total utility cost (\$/y)	126,059	59,356	59,356	31,910	31,910
TAC for HEN (\$/y)	251,343	188,943	192,251	113,489	113,722
TAC for separators (\$/y)	54,970	54,970	54,970	69,318	69,318
Overall TAC (\$/y)	306,313	243,913	247,221	182,807	183,040

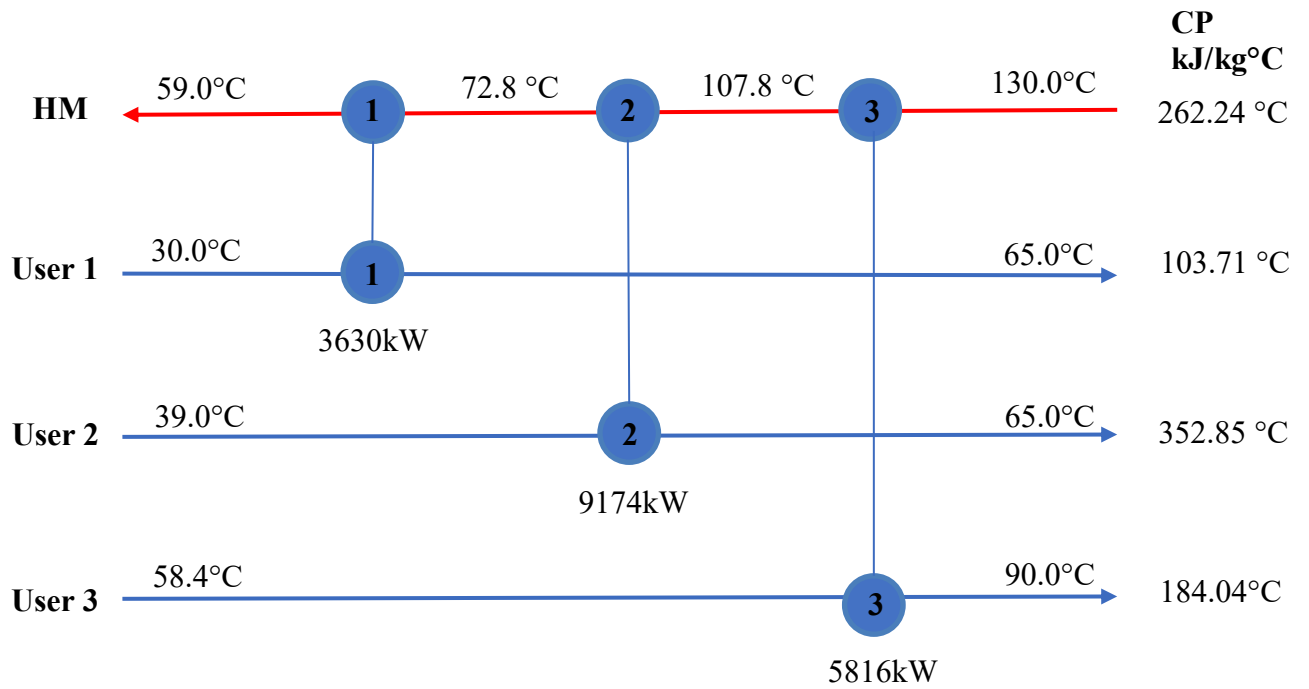


Figure 6.13. HEN for ATM supertargeting (Scenario 2)

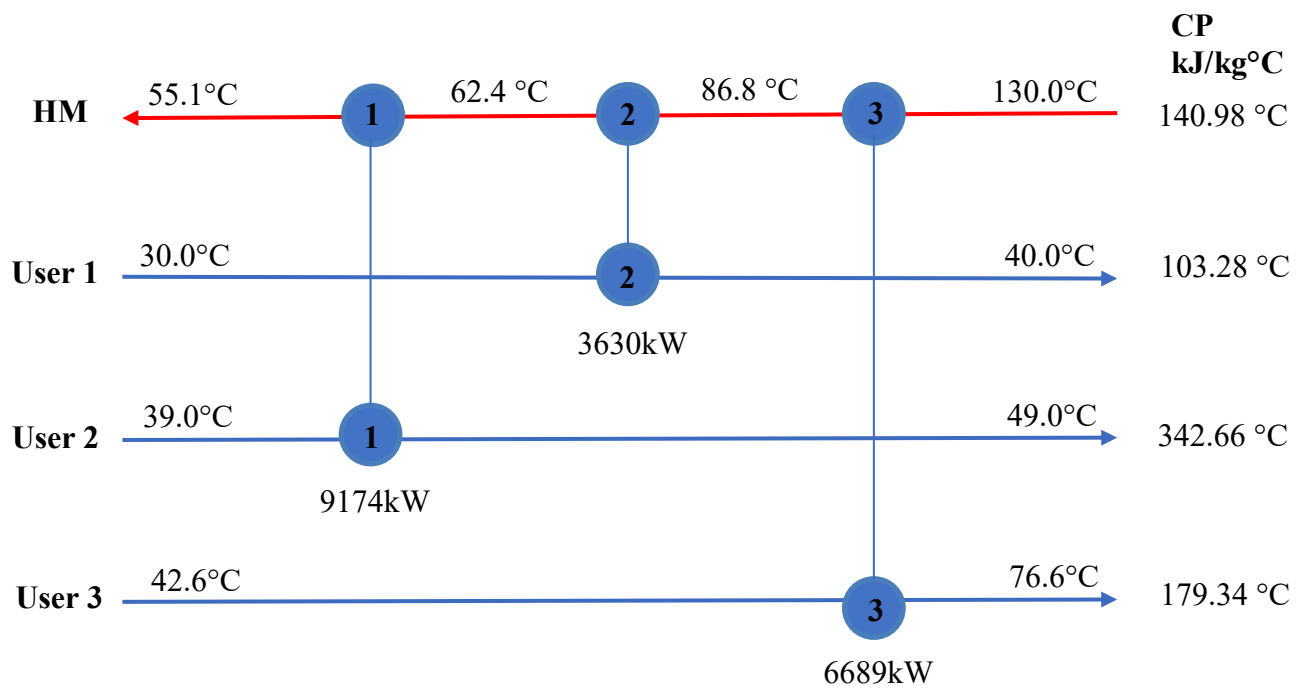


Figure 6.14. HEN for simultaneous optimisation

6.7 Summary

This chapter proposes a new approach in simultaneous optimisation of heating utility and phase separation systems of oil field platform, in which the systems exhibit high interactions among them. In order to solve the HEN with variable enthalpy problem, an improved ATM has been presented. An FPSO problem has been used to illustrate the proposed model. The problem is complex in nature, due to the existence of series arrangement in the phase separation system where temperature changes in the upstream crude oil streams directly effecting the downstream processes. In order to reduce the number of simulations for data extraction, a localised search approach is used to solve the problem. In optimising the separators, it is observed that the cost of separator does not necessarily reduce when crude oil temperature entering the separator (heaters outlet temperatures) is increased. Two different factors i.e. residence time and liquid droplet velocity plays a different role on the cost of separator for a liquid controlled unit. For cases where liquid droplet velocity is the limiting factor, the cost of separator reduces. The opposite holds true for separator that is controlled by residence time. Based on the simulation result for simultaneous optimisation of heating utility and phase separation systems, a lower heater outlet temperature was found to produce the lowest TAC. This is because the cost of separators was found to increase at higher temperature for heavy oil and electrostatic separator, while no cost benefit was found for the light oil separator. Thus, limited compensation from the phase separation system was unable to outweigh the increase in cost of HEN when total heat load is increased. The simultaneous optimisation approach leads a reduced TAC of 40.2% as compared to the conventional design.

CHAPTER 7

CONCLUSIONS, CONTRIBUTIONS AND FUTURE WORKS

7.1 Conclusions and contributions

This thesis provided several key contributions in the area of heat integration and flowsheet optimisation. First, a systematic procedure was proposed for the design of heating utility system. In the proposed approach, apart from heat exchanger network (HEN), other associated elements in the heating utility system (i.e. heating medium pump and waste heat recovery unit (WHRU)) were also included in the optimisation model. Besides, a revised nearest neighbour algorithm (NNA) is proposed for the design of HEN. The revised NNA provides an alternative approach to the established pinch design method (Linnhoff and Hindmarsh, 1983) for synthesising the HEN. Although the proposed approach may not produce the minimum HEN area, it is capable to design a HEN that fulfils the optimum minimum approach temperature determined at the targeting stage.

Apart from that, this thesis also studied in detail the effect of heating medium temperatures on the cost of the entire system. In Chapter 4, the impact on changing the heating medium return temperature on the capital cost of HEN, pump and operating cost of the pump was examined. While, Chapter 5 studied how the cost of heating medium system can be impacted by setting both supply and return temperature as variables. Note that, setting both temperatures as variables posed computational difficulties in determining the optimum solution. In this thesis, a new procedure is proposed for determining the minimum and maximum heating medium flowrates that are to be used as the lower and upper bounds in the automated targeting model (ATM) formulation. Besides, detailed pressure drop calculation that includes fitting losses at bends, valves and entrance and exit losses were considered in determining the optimum solution for the heating medium system.

A framework for simultaneous optimisation of heating utility and phase separation system was proposed in this thesis. In simultaneous optimisation, the outlet temperatures of the heating medium users, as well as the total heat loads required are no longer fixed. The existing supertargeting approach does not cater for these changes. In order to overcome this limitation, a new algorithm is

proposed to sort out the changing temperature levels and to locate the heating medium user in the correct temperature intervals without any user intervention. For the separator sizing, this thesis has shown that increasing the temperatures of crude oil does not always equals to reduce size and cost of separator. Based on the result obtained, it was concluded that the sizing of the separator is governed by either by liquid residence time and liquid droplet velocity. Only when the separator is governed by liquid droplet velocity, the cost of separator shows reduction when temperature of crude oil is increased. Besides, the industrial case study shows that more savings was achieved when interaction of the heating utility and phase separation systems was considered simultaneously, as compared to the case where they were treated independently.

7.2 Future works

During the period of this research, there were several areas that was found suitable for future work. These key areas are identified and elaborated as follows.

Heat exchanger pressure drop calculation

In chapter 5, detailed pressure drop was only considered for pipe and fittings. However, pressure drop for the heat exchangers were assumed values, due to limitation of the ATM model. The latter is meant for targeting, in which actual design of HEN remains as unknown during area targeting, thus preventing the calculation for pressure drop at each heat exchanger. Note that, a series design of heat exchangers will produce larger pressure drop as compared to a parallel design. As such, the impact of pressure drop on the HEN and the overall cost of the heating utility system requires further investigation. Model like stage wise superstructure (SWS, Yee and Grossmann, 1990) can be used for this instance since it is capable to determine the HEN before arriving at the optimum solution. The addition of heat exchangers, pipe and fitting losses however may increase the computation complexity of the model. As a solution strategy, one may use the values calculated from ATM as initial targets for SWS, in order to ease the computational difficulty.

Heating utility system design retrofit on existing platform

The results obtained in Chapters 4 and 5 have shown that the application of process integration on a new oil and gas platform can reduce the cost of the utility system. Such integration also can be carried out on existing platforms, however its viability requires further investigation. This is because any modification works on oil and gas platform have addition complication. For example, stationary platforms which are normally located in the middle of the sea or dessert have limited access to instruments and heavy machinery. Besides, carrying out modification on a life platform possess many safety issues and required careful handling. In the event when the platform is shutdown, the turndown cost against the possible savings should be examined in order to arrive at the actual savings.

Sensitivity analysis

The result obtained in Chapter 4 shows that the difference between the optimum solution and the next closest values are just 7% apart. A small change in the utility cost or other parameters for the equipment costing might produce a different solution point. Thus, it will be beneficial to carry out a sensitivity analysis on these parameters to understand which of them will affect total annualised cost (TAC) the most when altered.

Multi-period consideration

Oil and gas wells do not have same throughput during their lifetime. The production of oil normally increases to its peak and then subsequently decreases. Produced water, on the other hand has an opposite characteristic, where it started with minimum flowrate but continues to increase throughout the life of the wells. The common practise of designing the heating utility or phase separation systems at peak condition do not guarantee optimum condition, as the peak condition only occur at some part of the lifetime operation of the platform. Thus, the systems may be overdesigned for other periods of the platform lifetime. By considering multi-period optimisation, these systems can be designed based on the trade-off between different period of platform lifetime, instead at the peak condition.

Link process simulation to optimisation software directly

In the simultaneous optimisation model presented in Chapter 6, the outlet temperatures of the heat exchangers were set as variables. Due to this setting, new process data are required for both separators design and HEN optimisation for each different temperature. In the proposed model, this operation was carried by first discretising the range of temperatures chosen for analyses, followed by data acquisition from process simulation software (i.e. Aspen HYSYS) for the selected temperatures. The data was collected in Excel and linked to the optimisation software (i.e. LINGO) using build in @OLE function (Lingo: The modelling language and optimiser, 2018). Although the user intervention is moderate, this process can be improved by directly linking the two software. Doing so allows the outlet temperatures to be taken as continuous variables rather than discrete variables. The model can then be operated with the optimisation software choosing the outlet temperatures automatically. Hence, the process data can be retrieved from simulation software for calculation without user intervention. This direct link between process simulation and optimisation software will reduce the calculation time significantly.

Expanding simultaneous optimisation to more systems in the platform

In this thesis, a systematic procedure for simultaneous optimisation of heating utility and phase separation system has been proposed in Chapter 6. As shown, by considering interaction between two systems more savings was able to be achieved. Beside these two systems, there are several other systems in the offshore platform that interact with the heating utility and phase separation systems. These include the compression system that compresses gas from phase separation system (before delivering to onshore processing facility or/and for use as gas injection), cooling medium system that provides cooling to gas (after each stage of compression), as well as the produced water treatment system where the produced water is treated prior to discharge. More savings can be achieved by simultaneously optimising these systems. However, with increased number of systems in the model, the computation difficulty will also increase due to the increase of process variables in the

model. Thus, there lies the challenge for a systematic approach that is capable in considering all variables associated to different systems, while solving the model in an acceptable computation time.

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Appendices

Appendix A-Identification of temperature profiles for hot streams for heat exchangers

In order to design the heating utility network using the revised NNA, hot stream temperatures of the various heat exchangers are required. These data are may be derived using the cold stream temperatures (which correspond to the heating medium users). By assuming that the heat exchanger is operated as counter-current configuration, the minimum approach temperature, $\Delta T_{\min}$ is added to the inlet and outlet temperatures of the cold stream. Note that, in the event where the calculated value of the inlet temperature of the hot stream exceeds the heating medium temperature, T_{HM} , the latter value shall be used.

A simple example is illustrated here. For the industrial case study, unit 1 has the cold stream with inlet and outlet temperatures of 57 °C and 90 °C. The corresponding inlet and outlet temperatures of the hot stream can be determined easily as 130°C and 97°C, as shown in Figure A1. Similarly, hot stream temperatures of all other heat exchanger units can be determined as such.

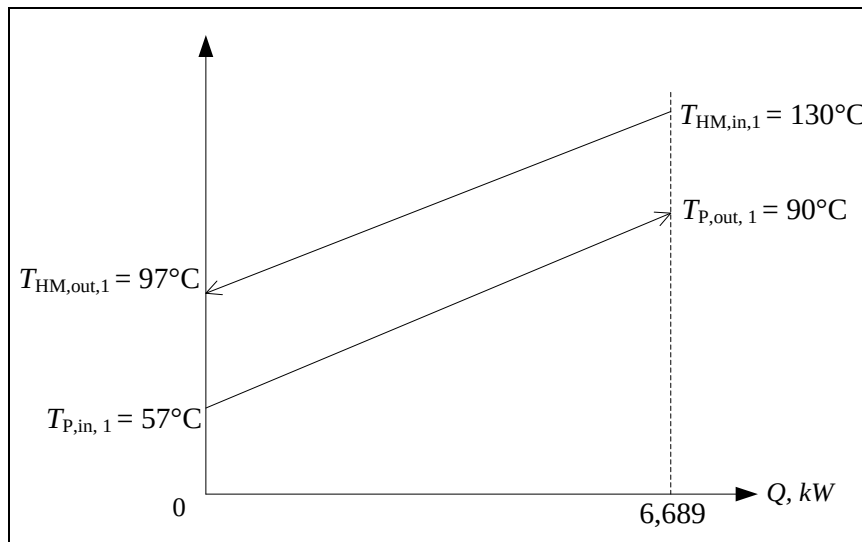


Figure A1. Hot and cold temperature profiles of unit 1 in industrial case study

Appendix B-WHRU area calculation

Approach reported by Nitsche and Gbadamosi (2016) is used to calculate the area for WHRU for finned tube heat exchanger type.

$$\frac{1}{U_{inn}} = \frac{1}{h_{oin}} + \frac{1}{h_i} + \frac{s}{\lambda_{tb}} \quad (B1)$$

$$h_{oin} = h_o \times \eta_w \times \frac{A_o}{A_{in}} \quad (B2)$$

$$q_{in} = U_{inn} \times A_{in} \times T_{WHRU} \quad (B3)$$

where h_o = outer film heat transfer coefficient

h_{oin} = outer film heat transfer coefficient based on the core tube surface area

λ_{tb} = heat conductivity of the core tube

s = tube wall thickness of the core tube

η_w = weighted fin efficiency

A_o = total outer surface area of the tube

A_{inn} = inner surface area of the tube

q_{in} = heat duty in relation to the inner tube surface area

Assumption has to be made on the geometry of the fin in order to determine the weighted fin efficiency (η_w) (Eq. B4). For this case, the chosen parameters for the fin geometry and the resulting surface area of the core tube (A_C), fin (A_{Fin}), total inner (A_{in}) and outer area (A_o) of the tube are listed in Table B1.

$$\eta_w = \frac{\eta_F \times A_{Fin} \times A_C}{A_o} \quad (B4)$$

$$\eta_F = \frac{\tanh X_{DF}}{X_{DF}} \quad (B5)$$

$$X_{DF} = X + \left(1 + 0.35 \times \ln \frac{Dia_{Fin}}{Dia_C} \right) \quad (B6)$$

$$X = H_F \times \sqrt{\frac{2 \times h_o}{w_{Fin} \times \lambda_{Fin}}} \quad (B7)$$

where A_{Fin} = fin surface area
 A_C = core tube surface area
 λ_{Fin} = heat conductivity of the fin
 H_{Fin} = fin height
 w_{Fin} = fin width
 η_{Fin} = fin efficiency
 Dia_C = core tube diameter
 Dia_{Fin} = fin diameter

Table B1.**Fin geometry for WHRU**

Fin height, H_F (mm)	Fin width, w_{Fin} (mm)	Core tube diameter, Dia_C (mm)	Fin diameter, Dia_{Fin} (mm)	Tube wall thickness of the core tube, s (mm)	Outer Area, A_o (m ² /m)	Core tube surface area, A_C (m ² /m)	Fin surface area, A_{Fin} (m ² /m)	Inner Area, A_{in} (m ² /m)
10	0.3	20	40	2	0.55	0.07	0.48	0.05

For the chosen fin geometry in Table B1, the weighted fin efficiency (η_w) is determined as 82% from Eq. B4, which is acceptable as it meets the minimum value of 80% suggested by Nitsche and Gbadamosi (2016). For gas stream with low heat film transfer coefficient, a large fin height is normally chosen, as it is more economical (2016). For this case, the inner (h_{inn}) and outer (h_o) heat film transfer coefficients are taken as 0.3 kW/m²°C and 0.04 kW/m²°C; while the heat conductivity of the core tube (λ_{tb}) and fin (λ_{Fin}) are assumed as 0.05 kW/m°C.

The WHRU log mean temperature (T_{WHRU}) is calculated based on Equation B8 that follows, i.e.,

$$T_{WHRU} = \frac{(T_{in,g} - T_{HM}) - (T_{out,g} - T_{HMR})}{LN[(T_{in,g} - T_{HM}) - (T_{out,g} - T_{HMR})]} \quad (B8)$$

where $T_{in, g}$ is the temperature of the flue gas entering the WHRU, while $T_{out, g}$ is the exhaust gas temperature leaving WHRU after heat transfer with the heating medium. The flue gas is assumed to have a fixed flowrate.

Appendix C-Pump power calculation

The shaft power required (P_{st}) to circulate the heating utility is calculated using Equation C1,

$$P_{st} = \frac{1.1F_{opt}g(d + hl_{p\&f})}{1 \times 10^3 CP_{HM}\eta} \quad (C1)$$

where g is the gravitational force, d is the head loss at equipment and differential height required to transfer the heating medium to its required users, $hl_{p\&f}$ is the head loss at piping and fittings, CP_{HM} is the specific heat capacity of heating medium and η is the pump efficiency. A design margin of 10% is included in the heating medium flowrate for pump design purpose.

Appendix D-Capital and operating costs

Capital cost correlations for heat exchangers (CC_{HEN}), pump (C_P) and WHRU (C_{WHRU}) are calculated by correlation given by Smith (2005) that follow.

$$C_{HEN} = N \left[a + b \left(A_{HEN}/N \right)^c \right] AF \quad (D1)$$

Where N = number of units or shells, whichever is appropriate

a = installation cost

b = area cost

c = constant

For this case study a , b and c is taken as 260,800, 3260 and 1 (Smith, 2005)

$$C_{PP} = C_B AF \left(\frac{P_s}{P_B} \right)^M \quad (D2)$$

where C_{PP} = pump cost with capacity P_s

C_B = known base cost of equipment with capacity P_B

M = constant depending on equipment type

For pump, the values for C_B and M are taken as \$US 9.2954 X 10⁴ and 0.55.

The capital cost of WHRU is calculated using Eq. C3. Note that the capital cost correlation for WHRU is based on that for air cooled heat exchanger.

$$C_{WHRU} = C_B AF \left(\frac{A_{WHRU}}{A_B} \right)^M \quad (D3)$$

where C_{WHRU} = WHRU cost with capacity A_{WHRU}

C_B = known base cost of equipment with capacity A_B

M = constant depending on equipment type

For WHRU, the values for C_B and M are taken as \$US 1.0171 X 10⁶ and 0.89.

Note that the capital cost correlation for WHRU is based on that for air cooled heat exchanger. Note that, the base costs of heat exchanger and pump have been adjusted according to Oct 2014 CE Index value of 652 and 941 (Economic Indicators, 2015).

The annualised factor, AF shown in Equations D1, D2 and D3 is used for the total capital cost of the equipment throughout the lifespan of the offshore platform (typically between a period of 20-30 years).

$$AF = \frac{i(1+i)^n}{(1+i)^n - 1} \quad (D4)$$

where i and n represent the interest rate and number of years respectively.

Next, the operating cost of the heating medium system (C_U) based on the power consumption of the circulation pump is determined using Equation D5.

$$C_U = P_s W G \quad (D5)$$

where P_s is the pump power, W is the duration of usage and G is the electricity tariff.

Appendix E-Pressure drop and piping cost calculation

The conventional approach for calculating pipe diameter (e.g. Chang et al. 2016) has been to fix the velocity of the fluid (v) to an acceptable value and using it to calculate the diameter. This approach does not necessarily reflect the actual practice in the industry where the velocity of the fluid can fall in an acceptable range of limit. In this work, the flow limit is derived based on American Petroleum Institute (API) 14E (1991), which recommends that the flow velocity to be lower than the maximum velocity of 4.572 m/s, in order to minimise flashing ahead of control valve. On the other way, it is recommended to limit the minimum velocity to 0.9144 m/s, in order to minimise deposition of sand and other solid. Hence, the fluid velocity is given by the constraint in Eq (E1).

$$0.9144 \leq v \leq 4.572 \quad (E1)$$

The inside diameter, Dia_{inn} is calculated using Eq. E2, based on velocity obtained from Eq. E1.

$$Dia_{inn} = \sqrt{\frac{4 \times F_{opt} \times 1.1}{v \times CP_{HM} \times \rho \times \pi}} \quad (E2)$$

The outside diameter, Dia_o and weight, wt_{PP} of the pipe is estimated using Eqs. E3-E4 based on Schedule 80 pipe (Stijepovic et al., 2011).

$$Dia_o = 1.101Dia_{inn} + 0.006349 \quad (E3)$$

$$wt_{PP} = 1330Dia_{inn}^2 + 75.18Dia_{inn} + 0.9268 \quad (E4)$$

The Reynolds number, Re is determined using Eq. E5

$$Re = \frac{\rho Dia_{in} v}{\mu} \quad (E5)$$

The Fanning factor, cf is calculated based on Reynolds number:

$$cf_1 = \frac{16}{Re} \quad Re \leq 2000 \quad (E6)$$

$$cf_2 = 0.046Re^{-0.2} \quad 2000 < Re < 20,000 \quad (E7)$$

$$cf_3 = 0.079Re^{-0.25} \quad Re \geq 20,000 \quad (E8)$$

Eqs. E9-E12 ensures only one Fanning factor is chosen based on the Reynolds number.

$$Re\varphi_1 \leq 2000 \quad (E9)$$

$$Re\varphi_3 \geq 20,000 \quad (E10)$$

$$Re(1 - \varphi_3) < 20,000 \quad (E11)$$

$$1 = \varphi_1 + \varphi_2 + \varphi_3 \quad (E12)$$

where φ_1 , φ_2 and φ_3 are binaries number.

The head loss on straight pipe, hl_{pp} is calculated by Equation E13

$$hl_{pp} = (\varphi_1 cf_1 + \varphi_2 cf_2 + \varphi_3 cf_2) \frac{2Lv^2}{9.81Dia_{inn}} \quad (E13)$$

The head loss across globe valve, hl_{gv} is calculated based on Eqs. E14-D19

$$CL_{gv1} = \frac{100}{Re^{0.33}} \quad Re \leq 5000 \quad (E14)$$

$$CL_{gv2} = 6 \quad Re > 5000 \quad (E15)$$

$$Re\varphi_{gv1} \leq 5000 \quad (E16)$$

$$Re > 5000(1 - \varphi_{gv1}) \quad (E17)$$

$$1 = \varphi_{gv1} + \varphi_{gv2} \quad (E18)$$

$$hl_{gv} = \left(\frac{\varphi_{gv1} CL_{gv1}}{\alpha^{0.5}} + \frac{\varphi_{gv2} CL_{gv2}}{\alpha^{0.5}} \right) \frac{2v^2}{9.81} \quad (E19)$$

where φ_{gv1} and φ_{gv2} are binaries, α is valve opening fraction, CL is loss coefficient.

The head loss across butterfly valve, hl_{bfly} is calculated based on Eqs. E20-E21

$$CL_{bfly} = 0.24 \quad (E20)$$

$$hl_{bfly} = \frac{CL_{bfly} 2v^2}{9.81} \quad (E21)$$

The head loss across gate valve, hl_{gtv} is calculated based on Eqs. E22-E27

$$CL_{gtv1} = \frac{1200}{Re} \quad Re \leq 6000 \quad (E22)$$

$$CL_{Lgtv2} = 2 \quad Re > 6000 \quad (E23)$$

$$Re\varphi_{gtv1} \leq 6000 \quad (E24)$$

$$Re > 6000(1 - \varphi_{gtv1}) \quad (E25)$$

$$1 = \varphi_{gtv1} + \varphi_{gtv2} \quad (E26)$$

$$hl_{gtv} = (\varphi_{gtv1}CL_{gtv1} + \varphi_{gtv2}CL_{gtv2}) \frac{2v^2}{9.81} \quad (E27)$$

where φ_{gtv1} and φ_{gtv2} are binaries.

The head loss across check valve, hl_{chv} is calculated based on Eqs. E28-E27

$$CL_{chv1} = \frac{1200}{Re^{0.86}} \quad Re \leq 1700 \quad (E28)$$

$$CL_{chv2} = 2 \quad Re > 1700 \quad (E29)$$

$$Re\varphi_{chv1} \leq 1700 \quad (E30)$$

$$Re > 1700(1 - \varphi_{chv1}) \quad (E31)$$

$$1 = \varphi_{chv1} + \varphi_{chv2} \quad (E32)$$

$$hl_{chv} = (\varphi_{chv1}CL_{chv1} + \varphi_{chv2}CL_{chv2}) \frac{2v^2}{9.81} \quad (E33)$$

where φ_{chv1} and φ_{chv2} are binaries.

The head loss across bend, hl_{bend} is calculated based on Eqs. E34-E39

$$CL_{bend1} = \frac{840}{Re} \quad Re \leq 1100 \quad (E34)$$

$$CL_{bend2} = 0.8 \quad Re > 1100 \quad (E35)$$

$$Re\varphi_{bend1} \leq 1100 \quad (E36)$$

$$Re > 1100(1 - \varphi_{bend1}) \quad (E37)$$

$$1 = \varphi_{bend1} + \varphi_{bend2} \quad (E38)$$

$$hl_{bend} = (\varphi_{bend1}CL_{bend1} + \varphi_{bend2}CL_{bend2}) \frac{2v^2}{9.81} \quad (E39)$$

where φ_{bend1} and φ_{bend2} are binaries.

The entrance loss, hl_{ent} is calculated based on Eq. E40

$$hl_{ent} = \left[0.5 \left(1 - \frac{A_2}{A_1} \right) \right] \left(\frac{v_2^2}{2g} \right) \quad (E40)$$

where A_1 and A_2 upstream and downstream pipe areas (m^2), v_2 is downstream velocities (ms^{-1})

The exit loss, hl_{exit} is calculated based on Eq. E41

$$hl_{exit} = \left[1 - \frac{A_1}{A_2}\right]^2 \left(\frac{v_1^2}{2g}\right) \quad (E41)$$

where v_1 is upstream velocities (ms^{-1})

The total head loss in piping and fittings are summed up as shown in Eq. E42.

$$hl_{p\&f} = hl_{pp} + hl_{gv} + hl_{bfly} + hl_{gtv} + hl_{chv} + hl_{bend} + hl_{ent} + hl_{exit} \quad (E42)$$

The piping cost is calculated based on Eq. E43 that follows.

$$C_{pp} = L \times AF(J_1 wt_{pp} + J_2 Dia_o^{0.48} + J_3 + J_4 Dia_o) \quad (E43)$$

where J_1 is pipe cost per unit weight

J_2 is installation cost

J_3 is right-of-way cost

J_4 is insulation cost

Appendix F-Separator sizing calculation

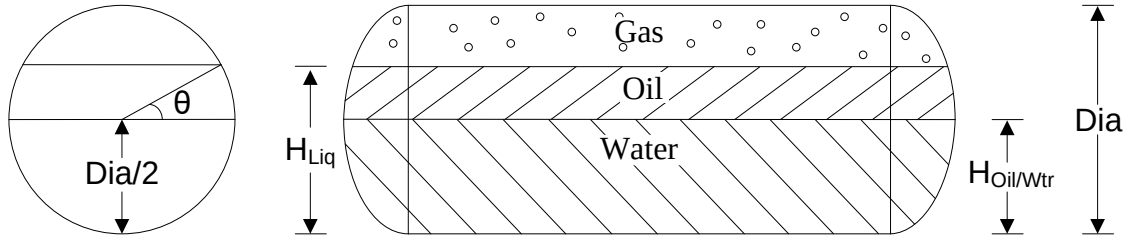


Figure F1: Interface level diagram for three phase separator

The sizing of the three phase separator is obtained from Manning and Thompson (1995). For verification of the coding in Lingo, an example case from Manning and Thompson (1995) was used. The formulation of that case and its result is presented in Appendix G. For gas capacity calculation, Equations F1 and F2 are used to determine the maximum superficial gas velocity (V_{max} , in m/s) and cross sectional area for gas flow (A_{gas} , in m^2), respectively:

$$V_{max} = K_{SB} \sqrt{\rho_{oil} - \rho_{gas} / \rho_{gas}} \quad (F1)$$

$$A_{gas} = (F_{gas} / V_{max}) \quad (F2)$$

where K_{SB} = Souders-Brown coefficient (m/s)

ρ_{oil} = Oil density (kg/m^3)

ρ_{gas} = Gas density (kg/m^3)

F_{gas} = Gas flow rate (m^3/s)

For this work, K_{SB} takes the value of 0.2438 m/s. To calculate the cross sectional area for liquid flow, A_{Liq} Equations F3 – F4 are used:

$$vol_{Liq} = (F_{wtr} r t_{wtr} + F_{oil} r t_{oil}) \quad (F3)$$

$$A_{Liq} = vol_{Liq} / L_{eff} \quad (F4)$$

where vol_{Liq} = liquid volume (m^3)

F_{wtr} = Feed rate of water (m^3/min)

F_{oil} = Feed rate of oil (m³/min)

rt_{wtr} = Theoretical water residence time (min)

rt_{oil} = Theoretical oil residence time (min)

L_{eff} = Portion of vessel shell length available for separation (m)

The minimum inside diameter of separator, Dia_{min} is calculated using Equations F5-F6.

$$A_{min} = A_{Li} + A_{gas} \quad (F5)$$

$$Dia_{min} = [A_{min}/(\pi/4)]^{0.5} \quad (F6)$$

where A_{min} = Cross sectional area for total flow (m²)

The actual inside diameter of separator, Dia is ensured to be larger than the minimum diameter required by inequality constraint of Equation F7.

$$Dia \geq Dia_{min} \quad (F7)$$

Height of liquid level in separator, Hgt_{Li} is calculated using Equations F8-F11

$$A_{tot} = (\pi/4)(Dia)^2 \quad (F8)$$

$$R = A_{Li}/A_{tot} \quad (F9)$$

$$R = 0.5 + (\pi/4)(ht_{Dia} - 0.5)(ht_{Dia} - ht_{Dia}^2)^{0.5} + (1/\pi)\sin^{-1}(2ht_{Dia} - 1) \quad (F10)$$

$$ht_{Dia} = Hgt_{Li}/Dia \quad (F11)$$

where A_{tot} = total cross sectional area (m²)

R = Horizontal cylinder partial volume coefficient (liquid filled fraction of the vessel-
dimensionless)

ht_{Dia} = Depth from liquid level to bottom of vessel divided by vessel inside diameter
(dimensionless)

Next, Equations F12 – F16 are used to calculate height of water-oil interface in separator, $Hgt_{oil/wtr}$.

$$vol_{wtr} = (F_{wtr}rt_{wtr}) \quad (F12)$$

$$A_{wtr} = vol_{wtr}/L_{eff} \quad (F13)$$

$$R_{wtr} = A_{wtr}/A_{tot} \quad (F14)$$

$$R_{wtr} = 0.5 + (\pi/4)(ht_{wtr} - 0.5)(ht_{wtr} - ht_{wtr}^2)^{0.5} + (1/\pi)\sin^{-1}(2ht_{wtr} - 1) \quad (F15)$$

$$ht_{wtr} = Hgt_{oil/wtr}/Dia \quad (F16)$$

where vol_{wtr} = Water volume (m³)

A_{wtr} = Cross sectional area for water flow (m²)

ht_{wtr} = Depth from water level to bottom of vessel divided by vessel inside diameter
(dimensionless)

R_{wtr} = Horizontal cylinder partial volume coefficient (water filled fraction of the vessel-
dimensionless)

$Hgt_{oil/wtr}$ = Height of water-oil interface in separator (m)

Gas capacity based on oil drop in gas phase is calculated using Equations F17-F22.

$$A_{actg} = A_{tot} - A_{Liq} \quad (F17)$$

$$V_{actg} = F_{gas}/A_{actg} \quad (F18)$$

$$V_{od} = (Dia - Hgt_{Liq})/rt_{tg} \quad (F19)$$

$$Dia_{od} = 1 \times 10^6 (V_{od}^2 3CD_{gas}\rho_{gas})/[4g(\rho_{oil}-\rho_{gas})] \quad (F20)$$

$$Re = Dia_{od}V_{od}\rho_{gas}/\mu_{gas} \quad (F21)$$

$$CD_{gas} = 24/RE + 0.3/RE^{0.5} + 0.34 \quad (F22)$$

where A_{actg} = Actual cross sectional area for gas flow (m²)

V_{actg} = Actual superficial gas velocity (m/s)

V_{od} = Free-rising velocity of oil drop (m/s)

Dia_{od} = Diameter of oil drop (μ)

rt_{tg} = Theoretical gas residence time (s)

CD_g = Drag coefficient for gas (Dimensionless)

g = Acceleration due to gravity (m/s)

μ_{gas} = Viscosity of gas (cP)

Liquid capacity based on water drop in oil phase is calculated using Equations F23-F25.

$$V_{wd} = (Hgt_{Liq} - Hgt_{oil/wtr})/60rt_{oil} \quad (F23)$$

$$Dia_{wd} = 1 \times 10^6 (V_{wd}^2 3CD_{wtr} \rho_{oil})/[4g(\rho_{wtr} - \rho_{oil})] \quad (F24)$$

$$Re = D_{wd} V_{wd} \rho_{oil} / \mu_{oil} \quad (F25)$$

where V_{wd} = Free-settling velocity of water drop (m/s)

Dia_{wd} = Diameter of water drop (μ)

CD_{wtr} = Drag coefficient for water (dimensionless)

ρ_{wtr} = Water density (kg/m³)

μ_{oil} = Viscosity of oil (cP)

Liquid phase capacity based oil drop in water phase is calculated using Equations F26-F28:

$$V_{od} = Hgt_{oil/wtr}/60rt_{wtr} \quad (F26)$$

$$Dia_{od} = 1 \times 10^6 (V_{od}^2 3CD_{oil}\rho_{wtr})/[4g(\rho_{wtr}-\rho_{oil})] \quad (F27)$$

$$Re = D_{od}V_{od}\rho_{wtr}/\mu_{wtr} \quad (F28)$$

where V_{od} = Free-settling velocity of oil drop (m/s)

D_{od} = Diameter of oil drop (μ)

CD_{oil} = Drag coefficient for oil (Dimensionless)

μ_{wtr} = Viscosity of water (cP)

Evans (1980) recommendation included below as part of the vessel sizing constraints

$$Hgt_{Liq}/Dia \leq 0.8 \quad (F29)$$

Also minimum vapour space constraint is added according to Shell Design Engineering Practice (2012)

$$Dia - Hgt_{Liq} \geq 0.8 \quad (F30)$$

The ratio between effective length, L_{eff} and vessel diameter, Dia is ensured within the guidelines of Shell Design Engineering Practice (2012)

$$3 \geq L_{eff}/Dia \leq 5 \quad (F31)$$

The total length (L) of the separator is calculated using Equations F32 that follows.

$$L = (4/3)L_{eff} \quad (F32)$$

The 2 phases separator is sized according to equations F3-F4, F6-E16, F23-F32 while equation F5 is modified as follow.

$$A_{min} = A_{Liq} \quad (F33)$$

The cost of the separator with length, L and diameter Dia is calculated based on the proposed approach by Seider et. al. (2009) discussed below.

The cylindrical shell thickness is computed from ASME pressure-vessel as shown in Equation F34.

$$TP = \frac{PS_{int}Dia}{4SE - 0.4PS_{int}} + CA \quad (F34)$$

where TP = vessel thickness to withstand internal pressure (m)

PS_{int} = Internal design gauge pressure (barg)

S = Maximum allowable stress of the shell material at the design temperature (barg)

E = Fractional weld efficiency (Dimensionless)

CA = Corrosion allowance (m)

The dry vessel weight (WT_{Dry} , in kg) is calculated using Equation F35.

$$WT_{Dry} = \pi(Dia + TP)(L + 0.8Dia)TP\rho_{st} \quad (F35)$$

An additional 15% margin is included on top of the dry vessel weight as an estimation of vessel internals weight

$$WT_{Act} = 1.15WT_{Dry} \quad (F36)$$

where WT_{Act} = Total vessel weight including vessel internals (kg)

ρ_{st} = steel density(kg/m³)

The cost of the separator is calculated from Equation F37.

$$C_{vss,m,n} = A \exp\{8.9552 - 0.2330[\ln(2.2046WT_{act})] + 0.04333[\ln(2.2046WT_{act})]^2\} \quad (F37)$$

Note that for each outlet temperature there will be a corresponding separator cost. Equation F38 is used to model this relationship with binary variable σ providing the link between the two variables.

$$C_{vss,n} = \sum_m (\sigma_{m,n} C_{vss,m,n}) \quad \forall m \in M \quad \forall n \in N \quad (F38)$$

Appendix G-Verification of three phase separator calculation

Mathematical formulation

```

!Pressure (bar);
Pressure=30;

!Ksb (m/s);
Ksb=0.24384;
trw=3;
tro=3;
G=9.81;
steelDen=7849.054;

Leff=9.144;
D=2.5908;
!Density (kg/m3);
VDen=26.43053;
LDen=1033.514;
ODen=757.3547;

!V=Vapour, L=liquid, O=oil, M=mixture;
!Flowrate (m3/hr);
VFlow=1499.552;
LFlow=401.1363;
OFlow=414.7283;

!Visosity (cP);
VVis=0.012;
LVis=0.45;
OVis=0.75;

!Dimension;
Leff/D>=3;
Leff/D<=5;
L<=13;
!Gas Capacity-Soulders-Brown K factor;
Vmax=Ksb*((ODen-VDen)/VDen)^(1/2);
Av=Vflow/(Vmax*3600);

!Liquid Phase Capacity-Retention time;
Vl=(Lflow*trw/60+OFlow*tro/60);
Al=Vl/Leff;
Amin=Av+Al;
Dmin=(Amin/(@pi()/4))^(1/2);
D>=Dmin;

Atot=(@pi()/4)*(D)^2;
ML=Al/Atot;
ML=0.5+(4/@pi())*(hdl-0.5)*(hdl-hdl^2)^0.5+(1/@pi())*@asin(2*hdl-1);

```

```

HL=hdl*D;

Vw=Lflow*trw/60;
Aw=Vw/Leff;
Mw=Aw/Atot;
Mw=0.5+(4/@pi())*(hdw-0.5)*(hdw-hdw^2)^0.5+(1/@pi())*@asin(2*hdw-1);
How=hdw*D;
Aactv=Atot-Al;
Vactg=Vflow/(3600*Aactv);

!Evans recommendation;
HL/D<=0.8;

D-HL>=0.3;

!Gas Gapacity-Oil Drop in Gas Phase;

Grt=Leff/Vactg;
Vod=(D-HL)/Grt;
dod=(1000000)*(((Vod^2)*3*Cdg*Vden)/(4*g*(Oden-Vden)));
REg=((dod/1000000)*Vod*Vden)/(Vvis/1000);
Cdg=(24/REg)+(0.3/(REg^(1/2)))+0.34;

!Liquid Capacity-Water Drop in Oil Phase;
Vwd=(HL-How)/(tro*60);
dwd=(1000000)*(((Vwd^2)*3*Cdw*Oden)/(4*g*(Lden-Oden)));
REw=((dwd/1000000)*Vwd*Oden)/(Ovis/1000);
Cdw=(24/REw)+(0.3/(REw^(1/2)))+0.34;

!Liquid Capacity-Oil Drop in Water Phase;
Voil=how/(trw*60);
doil=(1000000)*(((Voil^2)*3*Cdo*Lden)/(4*g*(Lden-Oden)));
REo=((doil/1000000)*Voil*Lden)/(Lvis/1000);
Cdo=(24/REo)+(0.3/(REo^(1/2)))+0.34;

L=(4/3)*Leff;

i=0.1;!interest;
py=25; !paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;

!pressure in psig and 110%;
Pd=Pressure*14.5038;
CA=0.125;
tp=((Pd*D*1000/25.4)/(2*15000*1-1.2*Pd))+CA;

platethickness=(tp)*25.4/1000;

!vessel weight calc;
DryVesselWeight=@pi()*(D+platethickness)*(L+0.8*D)*platethickness*steelDen;
ActualVesselWeight=DryVesselWeight*1.15;

!Costing calc;
Cv=@exp(8.9552-
0.233*@log(ActualVesselWeight*2.20462)+0.04333*(@log(ActualVesselWeight*2.2
0462)^2));
ACv=anl*CV*704.1/500;

min=acv;

```

3-Phase separator verification

Result

Local optimal solution found.
 Objective value: 25109.87
 Infeasibilities: 0.000000
 Total solver iterations: 22
 Elapsed runtime seconds: 0.09

Model Class: NLP

Total variables: 9
 Nonlinear variables: 3
 Integer variables: 0

Total constraints: 19
 Nonlinear constraints: 3

Total nonzeros: 21
 Nonlinear nonzeros: 3

Variable	Value	Reduced Cost
PRESSURE	30.00000	0.000000
KSB	0.2438400	0.000000
TRW	3.000000	0.000000
TRO	3.000000	0.000000
G	9.810000	0.000000
STEELDEN	7849.054	0.000000
LEFF	9.144000	0.000000
D	2.590800	0.000000
VDEN	26.43053	0.000000
LDEN	1033.514	0.000000
ODEN	757.3547	0.000000
VFLOW	1499.552	0.000000
LFLOW	401.1363	0.000000
OFLOW	414.7283	0.000000
VVIS	0.1200000E-01	0.000000
LVIS	0.4500000	0.000000
OVIS	0.7500000	0.000000
L	12.19200	0.000000
VMAX	1.282296	0.000000
AV	0.3248410	0.000000
VL	40.79323	0.000000
AL	4.461202	0.000000
AMIN	4.786043	0.000000
DMIN	2.468558	0.000000
ATOT	5.271785	0.000000
ML	0.8462413	0.000000
HDL	0.7889400	0.000000
HL	2.043986	0.000000

VW	20.05681	0.000000
AW	2.193440	0.000000
MW	0.4160716	0.000000
HDW	0.4338897	0.000000
HOW	1.124121	0.000000
AACTV	0.8105827	0.000000
VACTG	0.5138800	0.000000
GRT	17.79404	0.000000
VOD	0.3073019E-01	0.000000
DOD	31.14747	0.000000
CDG	11.93071	0.000000
REG	2.108204	0.000000
VWD	0.5110358E-02	0.000000
DWD	161.4212	0.000000
CDW	29.47990	0.000000
REW	0.8330094	0.000000
VOIL	0.6245119E-02	0.000000
DOIL	139.7605	0.000000
CDO	12.52432	0.000000
REO	2.004606	0.000000
I	0.1000000	0.000000
PY	25.00000	0.000000
ANL	0.1101681	0.000000
PD	435.1140	0.000000
CA	0.1250000	0.000000
TP	1.630592	0.000000
PLATETHICKNESS	0.4141703E-01	0.000000
DRYVESSELWEIGHT	38346.76	0.000000
ACTUALVESSELWEIGHT	44098.77	0.000000
CV	161854.3	0.000000
ACV	25109.87	0.000000

Appendix H-Aspen HYSYS simulation result for optimum case

Table H1A: Aspen HYSYS simulation result

Stream Name	From heavy oil header	S1	S2	S3	S4	S5	Gas (heavy) to MP compression system	From light oil header
Vapour Fraction	0.098	0.916	0.916	0.000	0.001	0.000	1.000	0.238
Temperature (°C)	29.906	29.906	29.667	29.906	40.000	29.667	29.667	39.368
Pressure (barg)	18.987	18.987	18.487	18.987	18.490	18.487	18.487	18.987
Std Gas Flow (MMSCFD)	7.391	7.391	7.392	0.000	0.024	0.000	7.867	67.581
Oil Phase Flow (barrel/day)	10332	103.319	103.229	10229	10305	103.229	0.000	28908
Water Phase Flow (barrel/day)	9088	90.884	90.845	8998	9066	90.845	0.000	34578

Table H1B: Aspen HYSYS simulation result (contd.)

Stream Name	S6	S7	S8	S9	S10	Gas (light) to MP compression system	S12	S13
Vapour Fraction	0.969	0.969	0.000	0.001	0.000	1.000	0.001	0.000
Temperature (°C)	39.368	39.072	39.368	49.000	39.072	39.072	39.897	39.897
Pressure (barg)	18.987	18.490	18.987	18.490	18.490	18.490	18.487	18.487
Std Gas Flow (MMSCFD)	84.330	84.334	0.000	0.146	0.000	85.231	0.024	0.000
Oil Phase Flow (barrel/day)	289.08	288.45	28619	28838	288.45	0.00	10408	10392
Water Phase Flow (barrel/day)	345.776	345.376	34232	34486	345.374	0	9157	1155

Table H1C: Aspen HYSYS simulation result (contd.)

Stream Name	Gas from heavy oil sep. to MP compression system	Heavy oil PW to PW treatment system	S14	S15	Gas from light oil sep. to MP compression system	Light oil PW to PW treatment system	S16	S19
Fraction	1.000	0.000	0.001	0.000	1.000	0.000	0.000	0.000
Temperature (°C)	39.897	39.897	48.901	48.901	48.901	48.901	40.094	49.269
Pressure (barg)	18.487	18.487	18.490	18.490	18.490	18.490	31.490	30.490
Std Gas Flow (MMSCFD)	0.042	0.000	0.145	0.000	0.187	0.000	0.000	0.000
Oil Phase Flow (barrel/day)	0.000	16.036	29127	29064	0.000	63.330	10380	29030
Water Phase Flow (barrel/day)	0.000	8002	34831	3229	0.000	31602	1154	3229

Table H1D: Aspen HYSYS simulation result (contd.)

Stream Name	Wash water	S20	S21	S22	S23	S24	To crude oil storage	Electrostatic separator PW to PW treatment system
Fraction	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Temperature (°C)	17.119	17.119	46.703	42.520	42.611	76.610	76.610	76.610
Pressure (barg)	30.487	30.487	30.490	30.487	28.490	28.490	28.490	28.490
Std Gas Flow (MMSCFD)	0.000	0.000	0.000	0.000	0.000	0.000	0.000	0.000
Oil Phase Flow (barrel/day)	0.000	0.000	39390	39228	39241	40624	40620	3.750
Water Phase Flow (barrel/day)	3073	3073	4383	7503	7504	7701	204.121	7496

Appendix I-Industrial case study in chapter 5

Scenario 1

Mathematical formulation

```
sets:
sinks/1..3/:fi,FCC;
intervals:dx,Tkin,Tkout,th,rdx,LGT,heat,cold,adt,rheat;!intervals 1 90-65,
2 65-57, 3 57-39, 39-30;
connections(sinks,intervals):y;
coldstream/1..4/:Tc;

endsets

Data:
n=4;
intervals=1..n;
fi=103.71,352.85,202.70;
Tkin=90,65,57,39;
Tkout=65,57,39,30;
Tc=90,65,57,39;
y = 0 1 1 1
    0 1 1 0
    1 1 0 0;
FCC=2,1,1.5;

enddata

TMIN>=10;
TMIN<=100;

THM=130;

!heat load at each level;
@for(intervals(k):
@sum (sinks(i):(Tkin(k)-Tkout(k))*y(i,k)*fi(i))=dx(k));

!heat load cascade;
adt(1)=dx(1);
@for(intervals (k)| k#LT#4:adt(k+1)=adt(k)+dx(k+1));

!energy limit;
@for(intervals (k):adt(k)<=fmin*(130-tmin-Tkout(k));
```

```

!reverse heat load summation;
@for(intervals (k) | k#LT#4: rheat(k+1)=rheat(k)-dx(k));
rheat(1)=19493;

!to calculate return temp;
THMR=(0-19493)/Fmin+thm;

!source temperature calc at turning point;
@for(intervals(k): ((rheat(k)-19493)/fmin)+thm=th(k));

!to ensure min tmin;
@smin(Th(2)-Tkin(2),Th(3)-Tkin(3),Th(4)-Tkin(4),THMR-Tkout(4))=tmin;

!log mean temp diff;
@for(intervals(k) | k#LT#4:
((Th(k)-Tc(k))-(Th(k+1)-Tc(k+1)))/@log((Th(k)-Tc(k))/(Th(k+1)-
Tc(k+1)))=lgt(k));
((Th(4)-Tc(4))-(Thmr-30))/@log((Th(4)-Tc(4))/(Thmr-30))=lgt(4);

!hot stream area requirement;
FCH=2;
@for(intervals(k) | k#LT#4: ((Fmin*(Th(k)-Th(k+1)))/(FCH*lgt(k)))=heat(k));
@for(intervals(k) | k#EQ#4: ((Fmin*(Th(4)-Thmr))/(FCH*lgt(k)))=heat(k));

!cold stream area requirement;
@for(intervals(k) | k#LT#4:
@sum(sinks(i): ((Tc(k)-Tc(k+1))*fi(i)*y(i,k)/(FCC(i)*lgt(k))))=cold(k));
@for(intervals(k) | k#EQ#4:
@sum(sinks(i): ((Tc(k)-30)*fi(i)*y(i,k)/(FCC(i)*lgt(k))))=cold(k));

!for Hex area calc;
@sum(intervals(k): cold(k)+heat(k))=sumAk;

!for Hex cost;
i=0.1;!interest;
py=25;!paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;
CI=40000*anl;
AC=500*anl;
CeHEX=(6.52/4.358);
CHEX=(CI+(AC*sumAk)/3)*3*CeHEX;

!for WHRU area calc;
GHI=433;!Hot gas in;
GHO=150;!Hot gas out;
GFCH=40;!Gas film heat transfer coefficient;
Lf=50;!Heat conductivity of the fin material;
bf=0.0003;!fin width;
hf=0.01;!fin height;
X=@sqrt((2*GFCH)/(bf*Lf))*hf;
df=40;!fin diameter;
dc=20;!fin core diameter;
s=2;!tube wall thickness of the core tube;
Xdf=X*(1+0.35*@log(df/dc));
nf=@tanh(xdf)/xdf;!fin efficiency;
Ai=0.05;!internal total surface area;
Ao=0.55;!total surface area;
Act=0.07;!core tube surface area;
Af=0.48;!fin surface area;
nw=((nf*Af+Act)/Ao);!weighted fin efficiency;
GFCHw= nw*GFCH*(Ao/Ai);!weighted film heat transfer coefficient;

```

```

WFCH=2000;!water film heat transfer coefficient;
Ui=1/(1/GFCHw+1/WFCH+s/(1000*50));!overall film heat transfer coefficient;
lgtWHRU=((GHI-THM)-(GHO-THMR))/@log((GHI-THM)/(GHO-THMR));
AWHRU=(1000*19493)/(lgtWHRU*Ui);

!for WHRU cost;
bc=156000;!base cost;
bs=200;!base size;
m=0.89;
CWHRU=(bc*(AWHRU/bs)^m)*anl*CeHEX;

!for pump power calc;
lds=984.12;!liquid desnsity;
lhc=4.326;!liquid heat capacity;
fl=(fmin*1.1/lhc)*3600/lds;!flowrate;
h=51.7908381;!pumping height;
eff=0.6;!efficiency;
Ps=(fl*lds*9.81*h)/(3600000*0.6);

!pump cost;
bcp=9840;!base cost;
bsp=4;!base size;
mp=0.55;
CePump=(9.41/4.358);
CPump=(bcp*(Ps/bsp)^mp)*anl*CePump;

!Utility cost;
UCPump=Ps*365*24*0.12;

!Total cost;
TotalCost=CHEx+CWHRU+CPump+UCPump;

min= totalcost;

```

Scenario 1 (Chapter 5)

Result

```

Global optimal solution found.
Objective value:                197027.3
Objective bound:                197025.5
Infeasibilities:                0.000000
Extended solver steps:          151
Total solver iterations:        22170617
Elapsed runtime seconds:        3.90

```

```

Model Class:                    NLP

```

```

Total variables:                33
Nonlinear variables:            14
Integer variables:              0

```

```

Total constraints:              35
Nonlinear constraints:          26

```

```

Total nonzeros:                96
Nonlinear nonzeros:            49

```

Variable	Value	Reduced Cost
N	4.000000	0.000000
TMIN	23.72863	0.000000
THM	130.0000	0.000000
FMIN	275.8924	0.000000
THMR	59.34564	0.000000
FCH	2.000000	0.000000
SUMAK	732.2834	0.000000
I	0.1000000	0.000000
PY	25.00000	0.000000
ANL	0.1101681	0.000000
CI	4406.723	0.000000
AC	55.08404	0.000000
CEHEX	1.496099	0.000000
CHEX	80127.02	0.000000
GHI	433.0000	0.000000
GHO	150.0000	0.000000
GFCH	40.00000	0.000000
LF	50.00000	0.000000
BF	0.3000000E-03	0.000000
HF	0.1000000E-01	0.000000

X	0.7302967	0.000000
DF	40.00000	0.000000
DC	20.00000	0.000000
S	2.000000	0.000000
XDF	0.9074678	0.000000
NF	0.7933225	0.000000
AI	0.5000000E-01	0.000000
AO	0.5500000	0.000000
ACT	0.7000000E-01	0.000000
AF	0.4800000	0.000000
NW	0.8196269	0.000000
GFCHW	360.6358	0.000000
WFCH	2000.000	0.000000
UI	301.8521	0.000000
LGTWHRU	175.9753	0.000000
AWHRU	366.9718	0.000000
BC	156000.0	0.000000
BS	200.0000	0.000000
M	0.8900000	0.000000
CWHRU	44131.34	0.000000
LDS	984.1200	0.000000
LHC	4.326000	0.000000
FL	256.6258	0.000000
H	51.79084	0.000000
EFF	0.6000000	0.000000
PS	59.40412	0.000000
BCP	9840.000	0.000000
BSP	4.000000	0.000000
MP	0.5500000	0.000000
CEPUMP	2.159247	0.000000
CPUMP	10323.32	0.000000
UCPUMP	62445.62	0.000000
TOTALCOST	197027.3	0.000000
FI (1)	103.7100	0.000000
FI (2)	352.8500	0.000000
FI (3)	202.7000	0.000000
FCC (1)	2.000000	0.000000
FCC (2)	1.000000	0.000000
FCC (3)	1.500000	0.000000
DX (1)	5067.500	0.000000
DX (2)	5274.080	0.000000
DX (3)	8218.080	0.000000
DX (4)	933.3900	0.000000
TKIN (1)	90.00000	0.000000
TKIN (2)	65.00000	0.000000
TKIN (3)	57.00000	0.000000
TKIN (4)	39.00000	0.000000
TKOUT (1)	65.00000	0.000000
TKOUT (2)	57.00000	0.000000
TKOUT (3)	39.00000	0.000000
TKOUT (4)	30.00000	0.000000
TH (1)	130.0000	0.000000
TH (2)	111.6323	0.000000
TH (3)	92.51589	0.000000
TH (4)	62.72863	0.000000
RDX (1)	1.234568	0.000000
RDX (2)	1.234568	0.000000
RDX (3)	1.234568	0.000000
RDX (4)	1.234568	0.000000
LGT (1)	43.23141	0.000000
LGT (2)	40.82216	0.000000

LGT(3)	29.22718	0.000000
LGT(4)	26.43776	0.000000
HEAT(1)	58.60901	0.000000
HEAT(2)	64.59825	0.000000
HEAT(3)	140.5897	0.000000
HEAT(4)	17.65165	0.000000
COLD(1)	78.14535	0.000000
COLD(2)	105.7932	0.000000
COLD(3)	249.2437	0.000000
COLD(4)	17.65259	0.000000
ADT(1)	5067.500	0.000000
ADT(2)	10341.58	0.000000
ADT(3)	18559.66	0.000000
ADT(4)	19493.05	0.000000
RHEAT(1)	19493.00	0.000000
RHEAT(2)	14425.50	0.000000
RHEAT(3)	9151.420	0.000000
RHEAT(4)	933.3400	0.000000
Y(1, 1)	0.000000	0.000000
Y(1, 2)	1.000000	0.000000
Y(1, 3)	1.000000	0.000000
Y(1, 4)	1.000000	0.000000
Y(2, 1)	0.000000	0.000000
Y(2, 2)	1.000000	0.000000
Y(2, 3)	1.000000	0.000000
Y(2, 4)	0.000000	0.000000
Y(3, 1)	1.000000	0.000000
Y(3, 2)	1.000000	0.000000
Y(3, 3)	0.000000	0.000000
Y(3, 4)	0.000000	0.000000
TC(1)	90.00000	0.000000
TC(2)	65.00000	0.000000
TC(3)	57.00000	0.000000
TC(4)	39.00000	0.000000

Scenario 2 (Chapter 5)

Mathematical formulation

```
sets:
sinks/1..3/:fi,FCC;
intervals:dx,Tkin,Tkout,th,rdx,LGT,heat,cold,adt,rheat;!intervals 1 90-65,
2 65-57, 3 57-39, 39-30;
connections(sinks,intervals):y;
coldstream/1..4/:Tc;

endsets

Data:
n=4;
intervals=1..n;
fi=103.71,352.85,202.70;
Tkin=90,65,57,39;
Tkout=65,57,39,30;
Tc=90,65,57,39;
y = 0 1 1 1
    0 1 1 0
    1 1 0 0;
FCC=2,1,1.5;

enddata

TMIN>=10;
TMIN<=100;

THM>=100;
THM<=130;

Fmin<=19493.00;
Fmin>=229.1316;

!heat load at each level;
@for(intervals(k):
@sum (sinks(i):(Tkin(k)-Tkout(k))*y(i,k)*fi(i))=dx(k));

!heat load cascade;
adt(1)=dx(1);
@for(intervals (k)| k#LT#4:adt(k+1)=adt(k)+dx(k+1));

!energy limit;
```

```

@for(intervals (k):adt(k)<=fmin*(thm-tmin-Tkout(k)));

!reverse heat load summation;
@for(intervals (k)| k#LT#4:rheat(k+1)=rheat(k)-dx(k);
rheat(1)=19493;

!to calculate return temp;
THMR=(0-19493)/Fmin+thm;

!source temperature calc at turning point;
@for(intervals(k):((rheat(k)-19493)/fmin)+thm=th(k));

!to ensure min tmin;
@smin(Th(2)-Tkin(2),Th(3)-Tkin(3),Th(4)-Tkin(4),THMR-Tkout(4))=tmin;

!log mean temp diff;
@for(intervals(k)| k#LT#4:
((Th(k)-Tc(k))-(Th(k+1)-Tc(k+1)))/@log((Th(k)-Tc(k))/(Th(k+1)-
Tc(k+1)))=lgt(k));
((Th(4)-Tc(4))-(Thmr-30))/@log((Th(4)-Tc(4))/(Thmr-30))=lgt(4);

!hot stream area requirement;
FCH=2;
@for(intervals(k)| k#LT#4:((Fmin*(Th(k)-Th(k+1)))/(FCH*lgt(k)))=heat(k);
@for(intervals(k)| k#EQ#4:((Fmin*(Th(4)-Thmr))/(FCH*lgt(k)))=heat(k));

!cold stream area requirement;
@for(intervals(k)| k#LT#4:
@sum(sinks(i):((Tc(k)-Tc(k+1))*fi(i)*y(i,k)/(FCC(i)*lgt(k))))=cold(k));
@for(intervals(k)| k#EQ#4:
@sum(sinks(i):((Tc(k)-30)*fi(i)*y(i,k)/(FCC(i)*lgt(k))))=cold(k));

!for Hex area calc;
@sum(intervals(k):cold(k)+heat(k))=sumAk;

!for Hex cost;
i=0.1;!interest;
py=25;!paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;
CI=40000*anl;
AC=500*anl;
CeHEX=(6.52/4.358);
CHEX=(CI+(AC*sumAk)/3)*3*CeHEX;

!for WHRU area calc;
GHI=433;!Hot gas in;
GHO=150;!Hot gas out;
GFCH=40;!Gas film heat transfer coefficient;
Lf=50;!Heat conductivity of the fin material;
bf=0.0003;!fin width;
hf=0.01;!fin height;
X=@sqrt((2*GFCH)/(bf*Lf))*hf;
df=40;!fin diameter;
dc=20;!fin core diameter;
s=2;!tube wall thickness of the core tube;
Xdf=X*(1+0.35*@log(df/dc));
nf=@tanh(xdf)/xdf;!fin efficiency;
Ai=0.05;!internal total surface area;
Ao=0.55;!total surface area;
Act=0.07;!core tube surface area;
Af=0.48;!fin surface area;

```

```

nw=(nf*Af+Act)/Ao;!weighted fin efficiency;
GFCHw= nw*GFCH*(Ao/Ai);!weighted film heat transfer coefficient;
WFCH=2000;!water film heat transfer coefficient;
Ui=1/(1/GFCHw+1/WFCH+s/(1000*50));!overall film heat transfer coefficient;
lgtWHRU=((GHI-THM)-(GHO-THMR))/@log((GHI-THM)/(GHO-THMR));
AWHRU=(1000*19493)/(lgtWHRU*Ui);

!for WHRU cost;
bc=156000;!base cost;
bs=200;!base size;
m=0.89;
CWHRU=(bc*(AWHRU/bs)^m)*anl*CeHEX;

!for pump power calc;
lds=984.12;!liquid desnsity;
lhc=4.326;!liquid heat capacity in kJ/kgC;
fl=(fmin*1.1/lhc)*3600/lds;!flowrate in m3/hr;
h=46.61175+hl+hlgv+hlbfly+hlgtv+hchv+hbend+hlent+hlenxt;!pumping height;
eff=0.6;!efficiency;
Ps=(fl*lds*9.81*h)/(3600000*0.6);

!piping loss;

flr=fl/3600;
miu=0.249/1000;!Viscosity;
L=80;!Lenght of pipe;
V<=4.572;!Velocity limit;
V>=0.9144;!Velocity limit;
Din=@sqrt((4*flr)/(v*@pi()));
Dout=1.101*Din+0.006349;
wt=1330*(Din^2)+75.18*Din+0.9268;

Re=(lds*Din*V)/miu;

cf1=16/Re;
cf2=0.046*(RE^(-0.2));
cf3=0.079*(RE^(-0.25));

@bin (b1);
@bin (b2);
@bin (b3);

1=b1+b2+b3;
RE*b1<=2000;
RE>=20000*b3;
RE*(1-b3)<=20000;

!Globe valve;
CLg1=100/(Re^0.33);
CLG2=6;
Re*(bg1)<=5000;
Re>=5000*(1-bg1);
@bin (bg1);
@bin (bg2);
1=bg1+bg2;
hlgv=3*((bg1*(CLg1/(0.5^0.5))+bg2*(CLG2/(0.5^0.5))*v^2)/(2*9.81);

!Butterfly valve;
CLb=0.24;
hlbfly=13*(CLb*v^2)/(2*9.81);

```

```

!Gate valve;
Clgt1=1200/Re;
Clgt2=0.2;
Re*(bgt1)<=6000;
Re>=6000*(1-bgt1);
@bin (bgt1);
@bin (bgt2);
l=bgt1+bgt2;
hlgtv=4*((bgt1*Clgt1+bgt2*Clgt2*v^2))/(2*9.81);

!Check valve;
Clch1=1200/(Re^0.86);
Clch2=2;
Re*(bch1)<=1700;
Re>=1700*(1-bch1);
@bin (bch1);
@bin (bch2);
l=bch1+bch2;
hchv=5*((bch1*Clch1+bch2*Clch2*v^2))/(2*9.81);

!bends;
Clb1=840/Re;
Clb2=0.8;
Re*(bb1)<=1100;
Re>=1100*(1-bb1);
@bin (bb1);
@bin (bb2);
l=bb1+bb2;
hbend=20*((bb1*Clb1+bb2*Clb2*v^2))/(2*9.81);

!exit and entrance loss;
hlent=4*(0.5*(v^2))/(2*9.81);
hlenxt=4*((v^2))/(2*9.81);

hl=((b1*cf1+b2*cf2+b3*cf3)*L^2*v^2)/(din*9.81);
Pcul=0.82*wt+185*(Dout^(0.48));
PP=Pcul*Anl*L;

!pump cost;
bcp=9840;!base cost;
bsp=4;!base size;
mp=0.55;
CePump=(9.41/4.358);
CPump=(bcp*(Ps/bsp)^mp)*anl*CePump;

!Utility cost;
UCPump=Ps*365*24*0.12;

!Total cost;
TotalCost=CHEX+CWHRU+CPump+UCPump+PP;

min= totalcost;

```

Scenario 2 (Chapter 5)

Result

```

Global optimal solution found.
Objective value:                195907.3
Objective bound:                195907.3
Infeasibilities:                0.000000
Extended solver steps:          3783
Total solver iterations:        1442247480
Elapsed runtime seconds:        505.78

```

```

Model Class:                    MINLP

```

```

Total variables:                69
Nonlinear variables:            40
Integer variables:              11

```

```

Total constraints:              80
Nonlinear constraints:          52

```

```

Total nonzeros:                222
Nonlinear nonzeros:            108

```

Variable	Value	Reduced Cost
N	4.000000	0.000000
TMIN	24.02399	0.000000
THM	130.0000	0.000000
FMIN	277.1091	0.000000
THMR	59.65586	0.000000
FCH	2.000000	0.000000
SUMAK	728.0337	0.000000
I	0.100000	0.000000
PY	25.00000	0.000000
ANL	0.1101681	0.000000
CI	4406.723	0.000000
AC	55.08404	0.000000
CEHEX	1.496099	0.000000
CHEX	79776.79	0.000000
GHI	433.0000	0.000000
GHO	150.0000	0.000000
GFCH	40.00000	0.000000
LF	50.00000	0.000000
BF	0.3000000E-03	0.000000
HF	0.1000000E-01	0.000000
X	0.7302967	0.000000

DF	40.00000	0.000000
DC	20.00000	0.000000
S	2.000000	0.000000
XDF	0.9074678	0.000000
NF	0.7933225	0.000000
AI	0.5000000E-01	0.000000
AO	0.5500000	0.000000
ACT	0.7000000E-01	0.000000
AF	0.4800000	0.000000
NW	0.8196269	0.000000
GFCHW	360.6358	0.000000
WFCH	2000.000	0.000000
UI	301.8521	0.000000
LGTWHRU	175.7332	0.000000
AWHRU	367.4775	0.000000
BC	156000.0	0.000000
BS	200.0000	0.000000
M	0.8900000	0.000000
CWHRU	44185.46	0.000000
LDS	984.1200	0.000000
LHC	4.326000	0.000000
FL	257.7575	0.000000
H	49.33174	0.000000
HL	0.1043945	0.000000
HLGV	1.084828	0.000000
HLBFLY	0.1329621	0.000000
HLGTV	0.3409286E-01	0.000000
HCHV	0.4261607	0.000000
HBEND	0.6818572	0.000000
HLENT	0.8523215E-01	0.000000
HLENXT	0.1704643	0.000000
EFF	0.6000000	0.000000
PS	56.83307	0.000000
FLR	0.7159931E-01	0.000000
MIU	0.2490000E-03	0.000000
L	80.00000	0.000000
V	0.9144000	0.000000
DIN	0.3157486	0.000000
DOUT	0.3539882	0.000000
WT	157.2620	0.000000
RE	1141107.	0.000000
CF1	0.1402147E-04	0.000000
CF2	0.2826784E-02	0.000000
CF3	0.2417105E-02	0.000000
B1	0.000000	0.000000
B2	0.000000	160.7670
B3	1.000000	137.3513
CLG1	1.002495	0.000000
CLG2	6.000000	0.000000
BG1	0.000000	-1148.750
BG2	1.000000	0.000000
CLB	0.2400000	0.000000
CLGT1	0.1051611E-02	0.000000
CLGT2	0.2000000	0.000000
BGT1	0.000000	-44.83381
BGT2	1.000000	0.000000
CLCH1	0.7411065E-02	0.000000
CLCH2	2.000000	0.000000
BCH1	0.000000	-561.4698
BCH2	1.000000	0.000000
CLB1	0.7361274E-03	0.000000

CLB2	0.8000000	0.000000
BB1	0.000000	-901.3577
BB2	1.000000	0.000000
PCUL	241.3341	0.000000
PP	2126.985	0.000000
BCP	9840.000	0.000000
BSP	4.000000	0.000000
MP	0.5500000	0.000000
CEPUMP	2.159247	0.000000
CPUMP	10075.13	0.000000
UCPUMP	59742.92	0.000000
TOTALCOST	195907.3	0.000000
FI (1)	103.7100	0.000000
FI (2)	352.8500	0.000000
FI (3)	202.7000	0.000000
FCC (1)	2.000000	0.000000
FCC (2)	1.000000	0.000000
FCC (3)	1.500000	0.000000
DX (1)	5067.500	0.000000
DX (2)	5274.080	0.000000
DX (3)	8218.080	0.000000
DX (4)	933.3900	0.000000
TKIN (1)	90.00000	0.000000
TKIN (2)	65.00000	0.000000
TKIN (3)	57.00000	0.000000
TKIN (4)	39.00000	0.000000
TKOUT (1)	65.00000	0.000000
TKOUT (2)	57.00000	0.000000
TKOUT (3)	39.00000	0.000000
TKOUT (4)	30.00000	0.000000
TH (1)	130.0000	0.000000
TH (2)	111.7130	0.000000
TH (3)	92.68047	0.000000
TH (4)	63.02399	0.000000
RDX (1)	1.234568	0.000000
RDX (2)	1.234568	0.000000
RDX (3)	1.234568	0.000000
RDX (4)	1.234568	0.000000
LGT (1)	43.26973	0.000000
LGT (2)	40.94933	0.000000
LGT (3)	29.46900	0.000000
LGT (4)	26.74115	0.000000
HEAT (1)	58.55710	0.000000
HEAT (2)	64.39764	0.000000
HEAT (3)	139.4360	0.000000
HEAT (4)	17.45138	0.000000
COLD (1)	78.07613	0.000000
COLD (2)	105.4647	0.000000
COLD (3)	247.1984	0.000000
COLD (4)	17.45231	0.000000
ADT (1)	5067.500	0.000000
ADT (2)	10341.58	0.000000
ADT (3)	18559.66	0.000000
ADT (4)	19493.05	0.000000
RHEAT (1)	19493.00	0.000000
RHEAT (2)	14425.50	0.000000
RHEAT (3)	9151.420	0.000000
RHEAT (4)	933.3400	0.000000
Y (1, 1)	0.000000	0.000000
Y (1, 2)	1.000000	0.000000
Y (1, 3)	1.000000	0.000000

Y(1, 4)	1.000000	0.000000
Y(2, 1)	0.000000	0.000000
Y(2, 2)	1.000000	0.000000
Y(2, 3)	1.000000	0.000000
Y(2, 4)	0.000000	0.000000
Y(3, 1)	1.000000	0.000000
Y(3, 2)	1.000000	0.000000
Y(3, 3)	0.000000	0.000000
Y(3, 4)	0.000000	0.000000
TC(1)	90.00000	0.000000
TC(2)	65.00000	0.000000
TC(3)	57.00000	0.000000
TC(4)	39.00000	0.000000

Appendix J-Industrial case study in chapter 6**Scenario 2 (ATM supertargeting)****Mathematical formulation (HEN)**

```
sets:
star:Tcin,Tcout,fi,FCC;
temp/1..6/:Tin,lukaku;
Lot/1..6/:P,T;
intervals:dx,Tkin,Tkout,th,rdx,LGT,heat,cold,adt,rheat,mun,Tw,TdifI;
Connections(temp,temp):Q;
Connections2(temp,lot):L;
Connections3(star,intervals):y,po;

ENDSETS
DATA:
m=3;
star=1..m;
Tcin=30.00,39.00,58.4;
n=5;
intervals=1..n;
P=5,4,3,2,1,0;
fi=103.714286,352.8461538,184.041794;
FCC=2,1,1.5;
Tcout=65,65,90;
ENDDATA

@Sum(star(o):fi(o)*(Tcout(o)-Tcin(o)))=totalload;

!temp sorting;

@for(intervals(k):dx(k)<=Totalload*Tw(k));
@for(intervals(k):@bin (Tw(k)));

@for (temp (i)| i#LE#m:Tin(i)=Tcin(i));
@for (temp (i)| i#GT#m:Tin(i)=Tcout(i-m));

@for (temp (i):
@for (temp (j)| i#NE#j:Q(i,j)*Tin(i)>=Tin(j)*Q(i,j)));
@for (temp (i):
@for (temp (j)| i#NE#j:(1-Q(i,j))*Tin(i)<=Tin(j)));

@for (temp (i):
```

```

@for (temp (j)| i#NE#j:@bin (Q(i,j)))));

@sum (temp (i):
@sum (temp (j)| i#NE#j:(Q(i,j))))=15;

@for (temp (i):
@for (lot (k):
@sum (temp (j)| i#NE#j:(Q(i,j)*L(i,k)))<=P(k)));

@for (temp (i):
@for (lot (k):@bin (L(i,k)))));

@for (temp (i):
@sum (lot (k):(L(i,k)))=1);

@for (lot (k):
@sum (temp (i):(L(i,k)))=1);

@for (lot (k):
@sum (temp (i):(L(i,k)*Tin(i)))=T(k));

!temp in and out sorting;

@for(intervals(k):Tkin(k)=T(k));
@for(intervals(k):Tkout(k)=T(k+1));

!Cold inlet temp. location at interval;
@for (temp (i):
@sum (temp (j)| i#NE#j:(Q(i,j)))=lukaku(i));

@for(star(o):
@for(intervals(k):Tcout(o)>=y(o,k)*Tkin(k)));

@for(star(o):
@for(intervals(k):@bin (y(o,k))));

@for(star(o):
@sum(intervals(k):y(o,k))=lukaku(o+m));

!Hot inlet temp.temp at interval;

@for(star(o):
@for(intervals(k):Tcin(o)>=(1-po(o,k))*Tkin(k)));

@for(star(o):
@for(intervals(k):@bin (po(o,k))));

@for(star(o):
@sum(intervals(k):po(o,k))=(5-lukaku(o)));

TMIN>=10;
TMIN<=100;

THM=130;

!heat load at each level;
@for(intervals(k):
@sum (star(o):(Tkin(k)-Tkout(k))*y(o,k)*(po(o,k))*fi(o))=dx(k));

!heat load cascade;

```

```

adt(1)=dx(1);
@for(intervals (k) | k#LT#5:adt(k+1)=adt(k)+dx(k+1));

!energy limit;
@for(intervals (k):adt(k)<=fmin*(130-tmin-Tkout(k));

!reverse heat load summation;
@for(intervals (k) | k#LT#5:rheat(k+1)=rheat(k)-dx(k));
rheat(1)=Totalload;

!reverse heat load summation;
@sum(intervals (k):dx(k))=totalload;

!to calculate return temp;
THMR=(0-Totalload)/Fmin+thm;

!source temperature calc at turning point;
@for(intervals(k):((rheat(k)-Totalload)/fmin)+thm)=th(k));

!to ensure min tmin;
@smin(Th(2)-Tkin(2),Th(3)-Tkin(3),Th(4)-Tkin(4),Th(5)-Tkin(5),THMR-
Tkout(5))=tmin;

!log mean temp diff;
@for(intervals(k):
(Th(k)-Tkin(k))*TW(k)+100*(1-TW(k))=Tdifi(k));

!log mean temp diff;
@for(intervals(k) | k#LT#5:
(Tdifi(k)-(Th(k+1)-Tkin(k+1)))/@log(Tdifi(k)/(Th(k+1)-Tkin(k+1)))=lgt(k));
(Tdifi(5)-(Thmr-30))/@log(Tdifi(5)/(Thmr-30))=lgt(5);

!hot stream area requirement;
FCH=2;
@for(intervals(k) | k#LT#5:((Fmin*(Th(k)-Th(k+1)))/(FCH*lgt(k)))=heat(k));
@for(intervals(k) | k#EQ#5:((Fmin*(Th(5)-Thmr))/(FCH*lgt(k)))=heat(k));

!cold stream area requirement;
@for(intervals(k) | k#LT#5:
@sum (star(o):((Tkin(k)-
Tkin(k+1))*fi(o)*y(o,k)*(po(o,k))/(FCC(o)*lgt(k))))=cold(k));
@for(intervals(k) | k#EQ#5:
@sum (star(o):((Tkin(k)-
30)*fi(o)*y(o,k)*(po(o,k))/(FCC(o)*lgt(k))))=cold(k));

!for Hex area calc;
@sum(intervals(k):cold(k)+heat(k))=sumAk;

!for Hex cost;
i=0.1;!interest;
py=25;!paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;
CI=40000*anl;
AC=500*anl;
CeHEX=(6.52/4.358);
CHEX=(CI+(AC*sumAk)/3)*3*CeHEX;

!for WHRU area calc;
GHI=433;!Hot gas in;

```

```

GHO=150;!Hot gas out;
GFCH=40;!Gas film heat transfer coefficient;
Lf=50;!Heat conductivity of the fin material;
bf=0.0003;!fin width;
hf=0.01;!fin height;
X=@sqrt ((2*GFCH)/(bf*Lf))*hf;
df=40;!fin diameter;
dc=20;!fin core diameter;
s=2;!tube wall thickness of the core tube;
Xdf=X*(1+0.35*@log(df/dc));
nf=@tanh(xdf)/xdf;!fin efficiency;
Ai=0.05; !internal total surface area;
Ao=0.55; !total surface area;
Act=0.07;!core tube surface area;
Af=0.48; !fin surface area;
nw=((nf*Af+Act)/Ao);!weighted fin efficiency;
GFCHw= nw*GFCH*(Ao/Ai);!weighted film heat transfer coefficient;
WFCH=2000;!water film heat transfer coefficient;
Ui=1/(1/GFCHw+1/WFCH+s/(1000*50));!overall film heat transfer coefficient;
lgtWHRU=((GHI-THM)-(GHO-THMR))/@log((GHI-THM)/(GHO-THMR));
AWHRU=(1000*Totalload)/(lgtWHRU*Ui);

!for WHRU cost;
bc=156000;!base cost;
bs=200;!base size;
mtt=0.89;
CWHRU=(bc*(AWHRU/bs)^mtt)*anl*CeHEX;

!for pump power calc;
lds=984.12;!liquid desnsity;
lhc=4.326;!liquid heat capacity;
fl=(fmin*1.1/lhc)*3600/lds;!flowrate;
h=51.7908381;!pumping height;
eff=0.6;!efficiency;
Ps=(fl*lds*9.81*h)/(3600000*0.6);

!pump cost;
bcp=9840;!base cost;
bsp=4;!base size;
mp=0.55;
CePump=(9.41/4.358);
CPump=(bcp*(Ps/bsp)^mp)*anl*CePump;

!Utility cost;
UCPump=Ps*365*24*0.12;

!Total cost;
HENCost=CHEx+UCPump+CWHRU+Cpump;
Totalcost=HENCost+Sepcost*anl;

min= hencost;

```

Scenario 2 (ATM supertargeting, Chapter 6)

Result (HEN)

```

Global optimal solution found.
Objective value:                188943.1
Objective bound:                188942.0
Infeasibilities:                0.000000
Extended solver steps:          783
Total solver iterations:        378984282
Elapsed runtime seconds:        23.44

Model Class:                    MINLP

Total variables:                193
Nonlinear variables:            136
Integer variables:              101

Total constraints:              233
Nonlinear constraints:          107

Total nonzeros:                853
Nonlinear nonzeros:            448

Variable      Value      Reduced Cost
M             3.000000    0.000000
N             5.000000    0.000000
TOTALLOAD    18619.72    0.000000
TMIN         23.55722    0.000000
THM          130.0000    0.000000
FMIN         262.2414    0.000000
THMR         58.99779    0.000000
FCH          2.000000    0.000000
SUMAK        697.2364    0.000000
I            0.100000    0.000000
PY           25.00000    0.000000
ANL          0.1101681    0.000000
CI           4406.723    0.000000
AC           55.08404    0.000000
CEHEX        1.496099    0.000000
CHEX         77238.76    0.000000
GHI          433.0000    0.000000
GHO          150.0000    0.000000
GFCH         40.00000    0.000000
LF           50.00000    0.000000
BF           0.3000000E-03 0.000000
HF           0.1000000E-01 0.000000
X            0.7302967    0.000000
DF           40.00000    0.000000
DC           20.00000    0.000000

```

S	2.000000	0.000000
XDF	0.9074678	0.000000
NF	0.7933225	0.000000
AI	0.5000000E-01	0.000000
AO	0.5500000	0.000000
ACT	0.7000000E-01	0.000000
AF	0.4800000	0.000000
NW	0.8196269	0.000000
GFCHW	360.6358	0.000000
WFCH	2000.000	0.000000
UI	301.8521	0.000000
LGTWHRU	176.2464	0.000000
AWHRU	349.9925	0.000000
BC	156000.0	0.000000
BS	200.0000	0.000000
MTT	0.8900000	0.000000
CWHRU	42309.34	0.000000
LDS	984.1200	0.000000
LHC	4.326000	0.000000
FL	243.9281	0.000000
H	51.79084	0.000000
EFF	0.6000000	0.000000
PS	56.46485	0.000000
BCP	9840.000	0.000000
BSP	4.000000	0.000000
MP	0.5500000	0.000000
CEPUMP	2.159247	0.000000
CPUMP	10039.18	0.000000
UCPUMP	59355.85	0.000000
HENCOST	188943.1	0.000000
TOTALCOST	188943.1	0.000000
SEPCOST	0.000000	0.000000
TCIN(1)	30.00000	0.000000
TCIN(2)	39.00000	0.000000
TCIN(3)	58.40000	0.000000
TCOUT(1)	65.00000	0.000000
TCOUT(2)	65.00000	0.000000
TCOUT(3)	90.00000	0.000000
FI(1)	103.7143	0.000000
FI(2)	352.8462	0.000000
FI(3)	184.0418	0.000000
FCC(1)	2.000000	0.000000
FCC(2)	1.000000	0.000000
FCC(3)	1.500000	0.000000
TIN(1)	30.00000	0.000000
TIN(2)	39.00000	0.000000
TIN(3)	58.40000	0.000000
TIN(4)	65.00000	0.000000
TIN(5)	65.00000	0.000000
TIN(6)	90.00000	0.000000
LUKAKU(1)	0.000000	0.000000
LUKAKU(2)	1.000000	0.000000
LUKAKU(3)	2.000000	0.000000
LUKAKU(4)	4.000000	0.000000
LUKAKU(5)	3.000000	0.000000
LUKAKU(6)	5.000000	0.000000
P(1)	5.000000	0.000000
P(2)	4.000000	0.000000
P(3)	3.000000	0.000000
P(4)	2.000000	0.000000
P(5)	1.000000	0.000000

P (6)	0.000000	0.000000
T (1)	90.00000	0.000000
T (2)	65.00000	0.000000
T (3)	65.00000	0.000000
T (4)	58.40000	0.000000
T (5)	39.00000	0.000000
T (6)	30.00000	0.000000
DX (1)	4601.045	0.000000
DX (2)	0.000000	0.000000
DX (3)	4227.975	0.000000
DX (4)	8857.273	0.000000
DX (5)	933.4286	0.000000
TKIN (1)	90.00000	0.000000
TKIN (2)	65.00000	0.000000
TKIN (3)	65.00000	0.000000
TKIN (4)	58.40000	0.000000
TKIN (5)	39.00000	0.000000
TKOUT (1)	65.00000	0.000000
TKOUT (2)	65.00000	0.000000
TKOUT (3)	58.40000	0.000000
TKOUT (4)	39.00000	0.000000
TKOUT (5)	30.00000	0.000000
TH (1)	130.0000	0.000000
TH (2)	112.4549	0.000000
TH (3)	112.4549	0.000000
TH (4)	96.33248	0.000000
TH (5)	62.55722	0.000000
RDX (1)	1.234568	0.000000
RDX (2)	1.234568	0.000000
RDX (3)	1.234568	0.000000
RDX (4)	1.234568	0.000000
RDX (5)	1.234568	0.000000
LGT (1)	43.62134	0.000000
LGT (2)	70.49341	0.000000
LGT (3)	42.51612	0.000000
LGT (4)	30.17632	0.000000
LGT (5)	26.18336	0.000000
HEAT (1)	52.73846	0.000000
HEAT (2)	0.000000	0.000000
HEAT (3)	49.72202	0.000000
HEAT (4)	146.7586	0.000000
HEAT (5)	17.82484	0.000000
COLD (1)	70.31794	0.000000
COLD (2)	0.000000	0.000000
COLD (3)	81.87073	0.000000
COLD (4)	260.1789	0.000000
COLD (5)	17.82484	0.000000
ADT (1)	4601.045	0.000000
ADT (2)	4601.045	0.000000
ADT (3)	8829.020	0.000000
ADT (4)	17686.29	0.000000
ADT (5)	18619.72	0.000000
RHEAT (1)	18619.72	0.000000
RHEAT (2)	14018.68	0.000000
RHEAT (3)	14018.68	0.000000
RHEAT (4)	9790.701	0.000000
RHEAT (5)	933.4286	0.000000
MUN (1)	1.234568	0.000000
MUN (2)	1.234568	0.000000
MUN (3)	1.234568	0.000000
MUN (4)	1.234568	0.000000

MUN(5)	1.234568	0.000000
TW(1)	1.000000	7389.328
TW(2)	0.000000	-5138.122
TW(3)	1.000000	6227.882
TW(4)	1.000000	29607.19
TW(5)	1.000000	4601.812
TDIFI(1)	40.00000	0.000000
TDIFI(2)	100.0000	0.000000
TDIFI(3)	47.45493	0.000000
TDIFI(4)	37.93248	0.000000
TDIFI(5)	23.55722	0.000000
Q(1, 1)	1.234568	0.000000
Q(1, 2)	0.000000	0.000000
Q(1, 3)	0.000000	0.000000
Q(1, 4)	0.000000	0.000000
Q(1, 5)	0.000000	0.000000
Q(1, 6)	0.000000	0.000000
Q(2, 1)	1.000000	-3743.953
Q(2, 2)	0.000000	0.000000
Q(2, 3)	0.000000	0.000000
Q(2, 4)	0.000000	0.000000
Q(2, 5)	0.000000	0.000000
Q(2, 6)	0.000000	0.000000
Q(3, 1)	1.000000	-3743.950
Q(3, 2)	1.000000	-3743.950
Q(3, 3)	0.000000	0.000000
Q(3, 4)	0.000000	0.000000
Q(3, 5)	0.000000	0.000000
Q(3, 6)	0.000000	0.000000
Q(4, 1)	1.000000	-2450.514
Q(4, 2)	1.000000	-2450.514
Q(4, 3)	1.000000	-2450.514
Q(4, 4)	0.000000	0.000000
Q(4, 5)	1.000000	-2450.514
Q(4, 6)	0.000000	0.000000
Q(5, 1)	1.000000	0.000000
Q(5, 2)	1.000000	0.000000
Q(5, 3)	1.000000	0.000000
Q(5, 4)	0.000000	0.000000
Q(5, 5)	1.234568	0.000000
Q(5, 6)	0.000000	0.000000
Q(6, 1)	1.000000	0.000000
Q(6, 2)	1.000000	0.000000
Q(6, 3)	1.000000	0.000000
Q(6, 4)	1.000000	0.000000
Q(6, 5)	1.000000	0.000000
Q(6, 6)	1.234568	0.000000
L(1, 1)	0.000000	25138.33
L(1, 2)	0.000000	33070.22
L(1, 3)	0.000000	24245.76
L(1, 4)	0.000000	33070.22
L(1, 5)	0.000000	14277.69
L(1, 6)	1.000000	0.000000
L(2, 1)	0.000000	609.7152
L(2, 2)	0.000000	6501.975
L(2, 3)	0.000000	0.000000
L(2, 4)	0.000000	6501.975
L(2, 5)	1.000000	-6809.178
L(2, 6)	0.000000	0.000000
L(3, 1)	0.000000	0.000000
L(3, 2)	0.000000	1495.727

L(3, 3)	0.000000	0.000000
L(3, 4)	1.000000	1495.727
L(3, 5)	0.000000	0.000000
L(3, 6)	0.000000	0.000000
L(4, 1)	0.000000	0.000000
L(4, 2)	1.000000	0.000000
L(4, 3)	0.000000	207.4289
L(4, 4)	0.000000	0.000000
L(4, 5)	0.000000	2523.953
L(4, 6)	0.000000	0.000000
L(5, 1)	0.000000	0.000000
L(5, 2)	0.000000	0.000000
L(5, 3)	1.000000	207.4289
L(5, 4)	0.000000	0.000000
L(5, 5)	0.000000	2523.953
L(5, 6)	0.000000	0.000000
L(6, 1)	1.000000	5665.634
L(6, 2)	0.000000	0.000000
L(6, 3)	0.000000	6658.778
L(6, 4)	0.000000	0.000000
L(6, 5)	0.000000	17750.02
L(6, 6)	0.000000	0.000000
Y(1, 1)	0.000000	0.000000
Y(1, 2)	1.000000	0.000000
Y(1, 3)	1.000000	-1514.574
Y(1, 4)	1.000000	-5064.402
Y(1, 5)	1.000000	-5201.589
Y(2, 1)	0.000000	8989.057
Y(2, 2)	1.000000	0.000000
Y(2, 3)	1.000000	-2239.291
Y(2, 4)	1.000000	-7226.046
Y(2, 5)	0.000000	-3743.959
Y(3, 1)	1.000000	0.000000
Y(3, 2)	1.000000	-3743.951
Y(3, 3)	1.000000	-3743.951
Y(3, 4)	1.000000	-3743.956
Y(3, 5)	1.000000	-3743.957
PO(1, 1)	1.000000	0.000000
PO(1, 2)	1.000000	-0.1545639E-02
PO(1, 3)	1.000000	-221.1385
PO(1, 4)	1.000000	-3770.966
PO(1, 5)	1.000000	-3908.153
PO(2, 1)	1.000000	-0.2257341E-02
PO(2, 2)	1.000000	0.000000
PO(2, 3)	1.000000	1504.659
PO(2, 4)	1.000000	-3482.097
PO(2, 5)	0.000000	-0.1055569E-01
PO(3, 1)	1.000000	3743.953
PO(3, 2)	1.000000	0.000000
PO(3, 3)	1.000000	0.000000
PO(3, 4)	0.000000	-5066.482
PO(3, 5)	0.000000	-6066.149

Scenario 2 (ATM supertargeting, Chapter 6)

Mathematical formulation (Heavy oil separator)

```

!Pressure (bar);
Pressure=45;

!Ksb (m/s);
Ksb=0.24384;
trw=5;
tro=5;
G=9.81;
steelDen=7849.054;

!Density (kg/m3);
VDen=15.3567907006939;
LDen=977.541104532845;
ODen=876.945605333668;

!V=Vapour, L=liquid, O=oil, M=mixture;
!Flowrate (m3/hr);
VFlow=7.67;
LFlow=61.84;
OFlow=70.31;

!Viscosity (cP);
VVis=0.013102474412709;
LVis=0.432021164733096;
OVis=7.73801222803332;

!Dimension;
Leff/D>=3;
Leff/D<=5;
L<=13;
!Gas Capacity-Soulders-Brown K factor;
Vmax=Ksb*((ODen-VDen)/VDen)^(1/2);
Av=Vflow/(Vmax*3600);

!Liquid Phase Capacity-Retention time;
Vl=(Lflow*trw/60+OFlow*tro/60);
Al=Vl/Leff;
Amin=Av+Al;
Dmin=(Amin/(@pi()/4))^(1/2);
D>=Dmin;

Atot=(@pi()/4)*(D)^2;
ML=Al/Atot;
ML=0.5+(4/@pi())*(hdl-0.5)*(hdl-hdl^2)^0.5+(1/@pi())*@asin(2*hdl-1);
HL=hdl*D;

```

```

Vw=Lflow*trw/60;
Aw=Vw/Leff;
Mw=Aw/Atot;
Mw=0.5+(4/@pi())*(hdw-0.5)*(hdw-hdw^2)^0.5+(1/@pi())*@asin(2*hdw-1);
How=hdw*D;
Aactv=Atot-Al;
Vactg=Vflow/(3600*Aactv);

!Evans recomendation;
HL/D<=0.8;
D-HL>=0.3;

!Gas Gapacity-Oil Drop in Gas Phase;

Grt=Leff/Vactg;
Vod=(D-Hl)/Grt;
dod=(1000000)*(((Vod^2)*3*Cdg*Vden)/(4*g*(Oden-Vden)));
REg=((dod/1000000)*Vod*Vden)/(Vvis/1000);
Cdg=(24/REg)+(0.3/(REg^(1/2)))+0.34;
dod<=300;

!Liquid Capacity-Water Drop in Oil Phase;
Vwd=(Hl-How)/(tro*60);
dwd=(1000000)*(((Vwd^2)*3*Cdw*Oden)/(4*g*(Lden-Oden)));
REw=((dwd/1000000)*Vwd*Oden)/(Ovis/1000);
Cdw=(24/REw)+(0.3/(REw^(1/2)))+0.34;
dwd<=600;

!Liquid Capacity-Oil Drop in Water Phase;
Voil=how/(trw*60);
doil=(1000000)*(((Voil^2)*3*Cdo*Lden)/(4*g*(Lden-Oden)));
REo=((doil/1000000)*Voil*Lden)/(Lvis/1000);
Cdo=(24/REo)+(0.3/(REo^(1/2)))+0.34;
doil<=300;

L=(4/3)*Leff;

i=0.1;!interest;
py=25; !paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;

!pressure in psig and 110%;
Pd=Pressure*14.5038;
CA=0.125;
tp=((Pd*D*1000/25.4)/(2*15000*1-1.2*Pd))+CA;

platethickness=(tp)*25.4/1000;

!vessel weight calc;
DryVesselWeight=@pi()*(D+platethickness)*(L+0.8*D)*platethickness*steelDen;
ActualVesselWeight=DryVesselWeight*1.15;

!Costing calc;
Cv=@exp(8.9552-
0.233*@log(ActualVesselWeight*2.20462)+0.04333*(@log(ActualVesselWeight*2.2
0462)^2));
ACv=anl*CV*704.1/500;

min=acv;

```

Scenario 2 (ATM supertargeting, Chapter 6)

Result (Heavy oil separator)

Global optimal solution found.

Objective value:	12831.62
Objective bound:	12831.62
Infeasibilities:	0.1799344E-07
Extended solver steps:	255
Total solver iterations:	1744865229
Elapsed runtime seconds:	42.73

Model Class:	NLP
--------------	-----

Total variables:	35
Nonlinear variables:	27
Integer variables:	0

Total constraints:	43
Nonlinear constraints:	27

Total nonzeros:	98
Nonlinear nonzeros:	45

Variable	Value	Reduced Cost
PRESSURE	45.00000	0.000000
KSB	0.2438400	0.000000
TRW	5.000000	0.000000
TRO	5.000000	0.000000
G	9.810000	0.000000
STEELDEN	7849.054	0.000000
VDEN	15.35679	0.000000
LDEN	977.5411	0.000000
ODEN	876.9456	0.000000
VFLOW	7.670000	0.000000
LFLOW	61.84000	0.000000
OFLOW	70.31000	0.000000
VVIS	0.1310247E-01	0.000000
LVIS	0.4320212	0.000000
OVIS	7.738012	0.000000
LEFF	7.427090	0.000000
D	1.485418	0.000000
L	9.902787	0.000000
VMAX	1.826437	0.000000
AV	0.1166509E-02	0.000000
VL	11.01250	0.000000
AL	1.482748	0.000000
AMIN	1.483914	0.000000

DMIN	1.374546	0.000000
ATOT	1.732955	0.000000
ML	0.8556180	0.000000
HDL	0.7980367	0.000000
HL	1.185418	0.000000
VW	5.153333	0.000000
AW	0.6938563	0.000000
MW	0.4003891	0.000000
HDW	0.4214413	0.000000
HOW	0.6260166	0.000000
AACTV	0.2502075	0.000000
VACTG	0.8515154E-02	0.000000
GRT	872.2203	0.000000
VOD	0.3439498E-03	0.000000
DOD	3.098672	0.000000
CDG	19221.75	0.000000
REG	0.1249159E-02	0.000000
VWD	0.1864672E-02	0.000000
DWD	514.4632	0.000000
CDW	222.0052	0.000000
REW	0.1087176	0.000000
VOIL	0.2086722E-02	0.000000
DOIL	129.4083	0.000000
CDO	40.00224	0.000000
REO	0.6110220	0.000000
I	0.1000000	0.000000
PY	25.00000	0.000000
ANL	0.1101681	0.000000
PD	652.6710	0.000000
CA	0.1250000	0.000000
TP	1.431402	0.000000
PLATETHICKNESS	0.3635760E-01	0.000000
DRYVESSELWEIGHT	15131.73	0.000000
ACTUALVESSELWEIGHT	17401.49	0.000000
CV	82710.64	0.1551387
ACV	12831.62	0.000000

Scenario 2 (ATM supertargeting, Chapter 6)

Mathematical formulation (Light oil separator)

```
!Pressure (bar);
Pressure=30;

!Ksb (m/s);
Ksb=0.24384;
trw=3;
tro=3;
G=9.81;
steelDen=7849.054;

!Density (kg/m3);
VDen=17.2608777588582;
LDen=977.413932524747;
ODen=800.002949614283;

!V=Vapour, L=liquid, O=oil, M=mixture;
!Flowrate (m3/hr);
VFlow=28.57447;
LFlow=233.624;
OFlow=195.595;

!Viscosity (cP);
VVis=0.0128355540376766;
LVis=0.431445349802656;
OVis=1.39824832301802;

!Dimension;
Leff/D>=3;
Leff/D<=5;
L<=13;
!Gas Capacity-Soulders-Brown K factor;
Vmax=Ksb* ( (ODen-VDen)/VDen)^(1/2);
Av=Vflow/(Vmax*3600);

!Liquid Phase Capacity-Retention time;
Vl=(Lflow*trw/60+OFlow*tro/60);
Al=Vl/Leff;
Amin=Av+Al;
Dmin=(Amin/(@pi()/4))^(1/2);
D>=Dmin;

Atot=(@pi()/4)*(D)^2;
ML=Al/Atot;
ML=0.5+(4/@pi())*(hdl-0.5)*(hdl-hdl^2)^0.5+(1/@pi())*@asin(2*hdl-1);
```

```

HL=hdl*D;

Vw=Lflow*trw/60;
Aw=Vw/Leff;
Mw=Aw/Atot;
Mw=0.5+(4/@pi())*(hdw-0.5)*(hdw-hdw^2)^0.5+(1/@pi())*@asin(2*hdw-1);
How=hdw*D;
Aactv=Atot-Al;
Vactg=Vflow/(3600*Aactv);

!Evans recommendation;
HL/D<=0.8;
D-HL>=0.3;

!Gas Gapacity-Oil Drop in Gas Phase;

Grt=Leff/Vactg;
Vod=(D-HL)/Grt;
dod=(1000000)*(((Vod^2)*3*Cdg*Vden)/(4*g*(Oden-Vden)));
REg=((dod/1000000)*Vod*Vden)/(Vvis/1000);
Cdg=(24/REg)+(0.3/(REg^(1/2)))+0.34;
dod<=300;

!Liquid Capacity-Water Drop in Oil Phase;
Vwd=(Hl-How)/(tro*60);
dwd=(1000000)*(((Vwd^2)*3*Cdw*Oden)/(4*g*(Lden-Oden)));
REw=((dwd/1000000)*Vwd*Oden)/(Ovis/1000);
Cdw=(24/REw)+(0.3/(REw^(1/2)))+0.34;
dwd<=300;

!Liquid Capacity-Oil Drop in Water Phase;
Voil=how/(trw*60);
doil=(1000000)*(((Voil^2)*3*Cdo*Lden)/(4*g*(Lden-Oden)));
REo=((doil/1000000)*Voil*Lden)/(Lvis/1000);
Cdo=(24/REo)+(0.3/(REo^(1/2)))+0.34;
doil<=300;
L=(4/3)*Leff;

i=0.1;!interest;
py=25;!paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;

!pressure in psig and 110%;
Pd=Pressure*14.5038;
CA=0.125;
tp=((Pd*D*1000/25.4)/(2*15000*1-1.2*Pd))+CA;
platethickness=(tp)*25.4/1000;

!vessel weight calc;
DryVesselWeight=@pi()*(D+platethickness)*(L+0.8*D)*platethickness*steelDen;
ActualVesselWeight=DryVesselWeight*1.15;

!Costing calc;
Cv=@exp(8.9552-
0.233*@log(ActualVesselWeight*2.20462)+0.04333*(@log(ActualVesselWeight*2.2
0462)^2));
ACv=anl*CV*704.1/500;

min=acv;

```

Scenario 2 (ATM supertargeting, Chapter 6)

Result (Light oil separator)

```

Global optimal solution found.
Objective value:                15376.22
Objective bound:                15376.22
Infeasibilities:               0.5451147E-07
Extended solver steps:         347
Total solver iterations:       3856891
Elapsed runtime seconds:       33.29

```

```

Model Class:                    NLP

```

```

Total variables:                35
Nonlinear variables:            27
Integer variables:              0

```

```

Total constraints:              43
Nonlinear constraints:          27

```

```

Total nonzeros:                98
Nonlinear nonzeros:            45

```

Variable	Value	Reduced Cost
PRESSURE	30.00000	0.000000
KSB	0.2438400	0.000000
TRW	3.000000	0.000000
TRO	3.000000	0.000000
G	9.810000	0.000000
STEELDEN	7849.054	0.000000
VLEN	17.26088	0.000000
LDEN	977.4139	0.000000
ODEN	800.0029	0.000000
VFLOW	28.57447	0.000000
LFLOW	233.6240	0.000000
OFLOW	195.5950	0.000000
VVIS	0.1283555E-01	0.000000
LVIS	0.4314453	0.000000
OVIS	1.398248	0.000000
LEFF	9.269745	0.000000
D	1.853949	0.000000
L	12.35966	0.000000
VMAX	1.642037	0.000000
AV	0.4833846E-02	0.000000
VL	21.46095	0.000000
AL	2.315161	0.000000
AMIN	2.319995	0.000000
DMIN	1.718694	0.000000

ATOT	2.699513	0.000000
ML	0.8576215	0.000000
HDL	0.8000000	0.000000
HL	1.483159	0.000000
VW	11.68120	0.000000
AW	1.260142	0.000000
MW	0.4668036	0.000000
HDW	0.4739158	0.000000
HOW	0.8786157	0.000000
AACTV	0.3843526	0.000000
VACTG	0.2065123E-01	0.000000
GRT	448.8714	0.000000
VOD	0.8260490E-03	0.000000
DOD	4.987944	0.000000
CDG	4335.843	0.000000
REG	0.5540841E-02	0.000000
VWD	0.3358575E-02	0.000000
DWD	221.9444	0.000000
CDW	57.07298	0.000000
REW	0.4264877	0.000000
VOIL	0.4881198E-02	0.000000
DOIL	150.4844	0.000000
CDO	14.99506	0.000000
REO	1.664066	0.000000
I	0.1000000	0.000000
PY	25.00000	0.000000
ANL	0.1101681	0.000000
PD	435.1140	0.000000
CA	0.1250000	0.000000
TP	1.202386	0.000000
PLATETHICKNESS	0.3054059E-01	0.000000
DRYVESSELWEIGHT	19645.49	0.000000
ACTUALVESSELWEIGHT	22592.32	0.000000
CV	99112.72	0.1551387
ACV	15376.22	0.000000

Scenario 2 (ATM supertargeting, Chapter 6)

Mathematical formulation (Electrostatic separator)

```

!Pressure (bar);
Pressure=45;

!Ksb (m/s);

trw=5;
tro=5;
G=9.81;
steelDen=7849.054;

!Density (kg/m3);

LDen=957.49025375297;
ODen=800.110731945569;

!V=Vapour,L=liquid,O=oil,M=mixture;
!Flowrate (m3/hr);

LFlow=51.92646;
OFlow=272.1585;

!Visocity (cP);

LVis=0.311316299287234;
OVis=1.39161104714917;

!Dimension;
Leff/D>=3;
Leff/D<=5;
L<=13;

!Liquid Phase Capacity-Retention time;
Vl=(Lflow*trw/60+OFlow*tro/60);
Al=Vl/Leff;
Amin=Al;
Dmin=(Amin/(@pi()/4))^(1/2);
D>=Dmin;

Atot=(@pi()/4)*(D)^2;
ML=Al/Atot;
ML=0.5+(4/@pi())*(hdl-0.5)*(hdl-hdl^2)^0.5+(1/@pi())*@asin(2*hdl-1);
HL=hdl*D;

Vw=Lflow*trw/60;
Aw=Vw/Leff;
Mw=Aw/Atot;

```

```

Mw=0.5+(4/@pi())*(hdw-0.5)*(hdw-hdw^2)^0.5+(1/@pi())*@asin(2*hdw-1);
How=hdw*D;

!Evans recommendation;
HL/D<=0.8;
D-HL>=0.3;

!Liquid Capacity-Water Drop in Oil Phase;
Vwd=(Hl-How)/(tro*60);
dwd=(1000000)*(((Vwd^2)*3*Cdw*Oden)/(4*g*(Lden-Oden)));
REw=((dwd/1000000)*Vwd*Oden)/(Ovis/1000);
Cdw=(24/REw)+(0.3/(REw^(1/2)))+0.34;
dwd<=250;

!Liquid Capacity-Oil Drop in Water Phase;
Voil=how/(trw*60);
doil=(1000000)*(((Voil^2)*3*Cdo*Lden)/(4*g*(Lden-Oden)));
REo=((doil/1000000)*Voil*Lden)/(Lvis/1000);
Cdo=(24/REo)+(0.3/(REo^(1/2)))+0.34;
doil<=300;
L=(4/3)*Leff;

i=0.1;!interest;
py=25;!paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;

!pressure in psig and 110%;
Pd=Pressure*14.5038;
CA=0.125;
tp=((Pd*D*1000/25.4)/(2*15000*1-1.2*Pd))+CA;

platethickness=(tp)*25.4/1000;

!vessel weight calc;
DryVesselWeight=@pi()*(D+platethickness)*(L+0.8*D)*platethickness*steelDen;
ActualVesselWeight=DryVesselWeight*1.15;

!Costing calc;
Cv=@exp(8.9552-
0.233*@log(ActualVesselWeight*2.20462)+0.04333*(@log(ActualVesselWeight*2.2
0462)^2));
ACv=anl*CV*704.1/500;

min=acv;

```

Scenario 2 (ATM supertargeting, Chapter 6)

Result (Electrostatic separator)

```

Global optimal solution found.
Objective value:                26762.39
Objective bound:                26762.14
Infeasibilities:                0.3026798E-08
Extended solver steps:          239
Total solver iterations:        12078823
Elapsed runtime seconds:        8.45

Model Class:                    NLP

Total variables:                28
Nonlinear variables:            20
Integer variables:              0

Total constraints:              35
Nonlinear constraints:          21

Total nonzeros:                77
Nonlinear nonzeros:            34

```

Variable	Value	Reduced Cost
PRESSURE	45.00000	0.000000
TRW	5.000000	0.000000
TRO	5.000000	0.000000
G	9.810000	0.000000
STEELDEN	7849.054	0.000000
LDEN	957.4903	0.000000
ODEN	800.1107	0.000000
LFLOW	51.92646	0.000000
OFLOW	272.1585	0.000000
LVIS	0.3113163	0.000000
OVIS	1.391611	0.000000
LEFF	9.750000	0.000000
D	2.169452	0.000000
L	13.00000	0.000000
VL	27.00708	0.000000
AL	2.769957	0.000000
AMIN	2.769957	0.000000
DMIN	1.877983	0.000000
ATOT	3.696494	0.000000
ML	0.7493470	0.000000
HDL	0.7014259	0.000000
HL	1.521710	0.000000
VW	4.327205	0.000000

AW	0.4438159	0.000000
MW	0.1200640	0.000000
HDW	0.1776037	0.000000
HOW	0.3853027	0.000000
VWD	0.3788024E-02	0.000000
DWD	250.0000	0.000000
CDW	44.82501	0.000000
REW	0.5444838	0.000000
VOIL	0.1284342E-02	0.000000
DOIL	68.62852	0.000000
CDO	89.44678	0.000000
REO	0.2710927	0.000000
I	0.1000000	0.000000
PY	25.00000	0.000000
ANL	0.1101681	0.000000
PD	652.6710	0.000000
CA	0.1250000	0.000000
TP	2.032999	0.000000
PLATETHICKNESS	0.5163817E-01	0.000000
DRYVESSELWEIGHT	41674.55	0.000000
ACTUALVESSELWEIGHT	47925.73	0.000000
CV	172506.2	0.1551387
ACV	26762.39	0.000000

Scenario 3 (Simultaneous optimisation, Chapter 6)

Mathematical formulation (Heavy oil separator)

```
!Pressure (bar);
Pressure=45;

!Ksb (m/s);
Ksb=0.24384;
trw=5;
tro=5;
G=9.81;
steelDen=7849.054;

!Density (kg/m3);
VDen=15.55;
LDen=996.78;
ODen=895.08;

!V=Vapour,L=liquid,O=oil,M=mixture;
!Flowrate (m3/hr);
VFlow=2.66000;
LFlow=60.6600;
OFlow=68.95000;

!Visocity (cP);
VVis=0.0122;
LVis=0.6527;
OVis=12.374;

!Dimension;
Leff/D>=3;
Leff/D<=5;
L<=13;
!Gas Capacity-Soulders-Brown K factor;
Vmax=Ksb*((ODen-VDen)/VDen)^(1/2);
Av=Vflow/(Vmax*3600);

!Liquid Phase Capacity-Retention time;
Vl=(Lflow*trw/60+OFlow*tro/60);
Al=Vl/Leff;
Amin=Av+Al;
Dmin=(Amin/(@pi()/4))^(1/2);
D>=Dmin;

Atot=(@pi()/4)*(D)^2;
ML=Al/Atot;
ML=0.5+(4/@pi())*(hdl-0.5)*(hdl-hdl^2)^0.5+(1/@pi())*@asin(2*hdl-1);
HL=hdl*D;
```

```

Vw=Lflow*trw/60;
Aw=Vw/Leff;
Mw=Aw/Atot;
Mw=0.5+(4/@pi())*(hdw-0.5)*(hdw-hdw^2)^0.5+(1/@pi())*@asin(2*hdw-1);
How=hdw*D;
Aactv=Atot-Al;
Vactg=Vflow/(3600*Aactv);

!Evans recomendation;
HL/D<=0.8;
D-HL>=0.3;

!Gas Gapacity-Oil Drop in Gas Phase;

Grt=Leff/Vactg;
Vod=(D-Hl)/Grt;
dod=(1000000)*(((Vod^2)*3*Cdg*Vden)/(4*g*(Oden-Vden)));
REg=((dod/1000000)*Vod*Vden)/(Vvis/1000);
Cdg=(24/REg)+(0.3/(REg^(1/2)))+0.34;
dod<=300;

!Liquid Capacity-Water Drop in Oil Phase;
Vwd=(Hl-How)/(tro*60);
dwd=(1000000)*(((Vwd^2)*3*Cdw*Oden)/(4*g*(Lden-Oden)));
REw=((dwd/1000000)*Vwd*Oden)/(Ovis/1000);
Cdw=(24/REw)+(0.3/(REw^(1/2)))+0.34;
dwd<=600;

!Liquid Capacity-Oil Drop in Water Phase;
Voil=how/(trw*60);
doil=(1000000)*(((Voil^2)*3*Cdo*Lden)/(4*g*(Lden-Oden)));
REo=((doil/1000000)*Voil*Lden)/(Lvis/1000);
Cdo=(24/REo)+(0.3/(REo^(1/2)))+0.34;
doil<=300;
L=(4/3)*Leff;

i=0.1;!interest;
py=25; !paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;

!pressure in psig and 110%;
Pd=Pressure*14.5038;
CA=0.125;
tp=((Pd*D*1000/25.4)/(2*15000*1-1.2*Pd))+CA;

platethickness=(tp)*25.4/1000;

!vessel weight calc;
DryVesselWeight=@pi()*(D+platethickness)*(L+0.8*D)*platethickness*steelDen;
ActualVesselWeight=DryVesselWeight*1.15;

!Costing calc;
Cv=@exp(8.9552-
0.233*@log(ActualVesselWeight*2.20462)+0.04333*(@log(ActualVesselWeight*2.2
0462)^2));
ACv=anl*CV*704.1/500;

min=acv;

```

Scenario 2 (Simultaneous optimisation, Chapter 6)

Result (Heavy oil separator)

Global optimal solution found.

Objective value:	14126.64
Objective bound:	14126.61
Infeasibilities:	0.000000
Extended solver steps:	1015
Total solver iterations:	37805
Elapsed runtime seconds:	63.41

Model Class:	NLP
--------------	-----

Total variables:	35
Nonlinear variables:	27
Integer variables:	0

Total constraints:	43
Nonlinear constraints:	27

Total nonzeros:	98
Nonlinear nonzeros:	45

Variable	Value	Reduced Cost
PRESSURE	45.00000	0.000000
KSB	0.2438400	0.000000
TRW	5.000000	0.000000
TRO	5.000000	0.000000
G	9.810000	0.000000
STEELDEN	7849.054	0.000000
VDEN	15.55000	0.000000
LDEN	996.7800	0.000000
ODEN	895.0800	0.000000
VFLOW	2.660000	0.000000
LFLOW	60.66000	0.000000
OFLOW	68.95000	0.000000
VVIS	0.1220000E-01	0.000000
LVIS	0.6527000	0.000000
OVIS	12.37400	0.000000
LEFF	7.792297	0.000000
D	1.558459	0.000000
L	10.38973	0.000000
VMAX	1.833855	0.000000
AV	0.4029156E-03	0.000000
VL	10.80083	0.000000
AL	1.386091	0.000000
AMIN	1.386494	0.000000

DMIN	1.328661	0.000000
ATOT	1.907572	0.000000
ML	0.7266260	0.000000
HDL	0.6821016	0.000000
HL	1.063028	0.000000
VW	5.055000	0.000000
AW	0.6487176	0.000000
MW	0.3400751	0.000000
HDW	0.3730167	0.000000
HOW	0.5813314	0.000000
AACV	0.5214805	0.000000
VACTG	0.1416906E-02	0.000000
GRT	5499.515	0.000000
VOD	0.9008644E-04	0.000000
DOD	1.514336	0.000000
CDG	138048.4	0.000000
REG	0.1738811E-03	0.000000
VWD	0.1605654E-02	0.000000
DWD	600.0000	0.000000
CDW	345.8710	0.000000
REW	0.6968752E-01	0.000000
VOIL	0.1937771E-02	0.000000
DOIL	152.1713	0.000000
CDO	54.08253	0.000000
REO	0.4503196	0.000000
I	0.1000000	0.000000
PY	25.00000	0.000000
ANL	0.1101681	0.000000
PD	652.6710	0.000000
CA	0.1250000	0.000000
TP	1.495640	0.000000
PLATETHICKNESS	0.3798926E-01	0.000000
DRYVESSELWEIGHT	17402.24	0.000000
ACTUALVESSELWEIGHT	20012.58	0.000000
CV	91058.15	0.1551387
ACV	14126.64	0.000000

Scenario 3 (Simultaneous optimisation, Chapter 6)

Mathematical formulation (Light oil separator)

```
!Pressure (bar);
Pressure=30;

!Ksb (m/s);
Ksb=0.24384;
trw=3;
tro=3;
G=9.81;
steelDen=7849.054;

!Density (kg/m3);
VDen=16.967599124834;
LDen=989.7989647392;
ODen=812.272524586714;

!V=Vapour,L=liquid,O=oil,M=mixture;
!Flowrate (m3/hr);
VFlow=12.095;
LFlow=230.73;
OFlow=192.95;

!Visocity (cP);
VVis=0.0123823472401673;
LVis=0.554489287114886;
OVis=1.77043389159381;

!Dimension;
Leff/D>=3;
Leff/D<=5;
L<=13;
!Gas Capacity-Soulders-Brown K factor;
Vmax=Ksb*((ODen-VDen)/VDen)^(1/2);
Av=Vflow/(Vmax*3600);

!Liquid Phase Capacity-Retention time;
Vl=(Lflow*trw/60+OFlow*tro/60);
Al=Vl/Leff;
Amin=Av+Al;
Dmin=(Amin/(@pi()/4))^(1/2);
D>=Dmin;

Atot=(@pi()/4)*(D)^2;
ML=Al/Atot;
ML=0.5+(4/@pi())*(hdl-0.5)*(hdl-hdl^2)^0.5+(1/@pi())*@asin(2*hdl-1);
HL=hdl*D;
```

```

Vw=Lflow*trw/60;
Aw=Vw/Leff;
Mw=Aw/Atot;
Mw=0.5+(4/@pi())*(hdw-0.5)*(hdw-hdw^2)^0.5+(1/@pi())*@asin(2*hdw-1);
How=hdw*D;
Aactv=Atot-Al;
Vactg=Vflow/(3600*Aactv);

!Evans recomendation;
HL/D<=0.8;
D-HL>=0.3;

!Gas Gapacity-Oil Drop in Gas Phase;

Grt=Leff/Vactg;
Vod=(D-Hl)/Grt;
dod=(1000000)*(((Vod^2)*3*Cdg*Vden)/(4*g*(Oden-Vden)));
REg=((dod/1000000)*Vod*Vden)/(Vvis/1000);
Cdg=(24/REg)+(0.3/(REg^(1/2)))+0.34;
dod<=300;

!Liquid Capacity-Water Drop in Oil Phase;
Vwd=(Hl-How)/(tro*60);
dwd=(1000000)*(((Vwd^2)*3*Cdw*Oden)/(4*g*(Lden-Oden)));
REw=((dwd/1000000)*Vwd*Oden)/(Ovis/1000);
Cdw=(24/REw)+(0.3/(REw^(1/2)))+0.34;
dwd<=300;

!Liquid Capacity-Oil Drop in Water Phase;
Voil=how/(trw*60);
doil=(1000000)*(((Voil^2)*3*Cdo*Lden)/(4*g*(Lden-Oden)));
REo=((doil/1000000)*Voil*Lden)/(Lvis/1000);
Cdo=(24/REo)+(0.3/(REo^(1/2)))+0.34;
doil<=300;

L=(4/3)*Leff;

i=0.1;!interest;
py=25; !paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;

!pressure in psig and 110%;
Pd=Pressure*14.5038;
CA=0.125;
tp=((Pd*D*1000/25.4)/(2*15000*1-1.2*Pd))+CA;

platethickness=(tp)*25.4/1000;

!vessel weight calc;
DryVesselWeight=@pi()*(D+platethickness)*(L+0.8*D)*platethickness*steelDen;
ActualVesselWeight=DryVesselWeight*1.15;

!Costing calc;
Cv=@exp(8.9552-
0.233*@log(ActualVesselWeight*2.20462)+0.04333*(@log(ActualVesselWeight*2.2
0462)^2));
ACv=anl*CV*704.1/500;

min=acv;

```

Scenario 2 (Simultaneous optimisation, Chapter 6)

Result (Light oil separator)

```

Global optimal solution found.
Objective value:                15241.22
Objective bound:                15241.22
Infeasibilities:                0.1579610E-07
Extended solver steps:         187
Total solver iterations:        669645
Elapsed runtime seconds:        11.12

```

```

Model Class:                    NLP

```

```

Total variables:                35
Nonlinear variables:            27
Integer variables:              0

```

```

Total constraints:              43
Nonlinear constraints:          27

```

```

Total nonzeros:                98
Nonlinear nonzeros:            45

```

Variable	Value	Reduced Cost
PRESSURE	30.00000	0.000000
KSB	0.2438400	0.000000
TRW	3.000000	0.000000
TRO	3.000000	0.000000
G	9.810000	0.000000
STEELDEN	7849.054	0.000000
VDEN	16.96760	0.000000
LDEN	989.7990	0.000000
ODEN	812.2725	0.000000
VFLOW	12.09500	0.000000
LFLOW	230.7300	0.000000
OFLOW	192.9500	0.000000
VVIS	0.1238235E-01	0.000000
LVIS	0.5544893	0.000000
OVIS	1.770434	0.000000
LEFF	9.229698	0.000000
D	1.845940	0.000000
L	12.30626	0.000000
VMAX	1.669405	0.000000
AV	0.2012527E-02	0.000000
VL	21.18400	0.000000
AL	2.295200	0.000000
AMIN	2.297212	0.000000
DMIN	1.710234	0.000000
ATOT	2.676239	0.000000

ML	0.8576215	0.000000
HDL	0.8000000	0.000000
HL	1.476752	0.000000
VW	11.53650	0.000000
AW	1.249933	0.000000
MW	0.4670483	0.000000
HDW	0.4741082	0.000000
HOW	0.8751751	0.000000
AACTV	0.3810388	0.000000
VACTG	0.8817270E-02	0.000000
GRT	1046.775	0.000000
VOD	0.3526908E-03	0.000000
DOD	3.175006	0.000000
CDG	15648.66	0.000000
REG	0.1534462E-02	0.000000
VWD	0.3342092E-02	0.000000
DWD	248.9154	0.000000
CDW	63.70653	0.000000
REW	0.3816737	0.000000
VOIL	0.4862084E-02	0.000000
DOIL	169.9115	0.000000
CDO	16.86171	0.000000
REO	1.474684	0.000000
I	0.1000000	0.000000
PY	25.00000	0.000000
ANL	0.1101681	0.000000
PD	435.1140	0.000000
CA	0.1250000	0.000000
TP	1.197731	0.000000
PLATETHICKNESS	0.3042237E-01	0.000000
DRYVESSELWEIGHT	19400.86	0.000000
ACTUALVESSELWEIGHT	22310.99	0.000000
CV	98242.53	0.1551387
ACV	15241.22	0.000000

Scenario 2 (Simultaneous optimisation, Chapter 6)

Mathematical formulation (Electrostatic separator)

```
!Pressure (bar);
Pressure=45;

!Ksb (m/s);

trw=5;
tro=5;
G=9.81;
steelDen=7849.054;

!Density (kg/m3);

LDen=968.350336761544;
ODen=810.267118896601;

!V=Vapour,L=liquid,O=oil,M=mixture;
!Flowrate (m3/hr);

LFlow=51.01251414;
OFlow=269.111694;

!Visocity (cP);

LVis=0.366511816018134;
OVis=1.69985053731151;

!Dimension;
Leff/D>=3;
Leff/D<=5;
L<=13;

!Liquid Phase Capacity-Retention time;
Vl=(Lflow*trw/60+OFlow*tro/60);
Al=Vl/Leff;
Amin=Al;
Dmin=(Amin/(@pi()/4))^(1/2);
D>=Dmin;

Atot=(@pi()/4)*(D)^2;
ML=Al/Atot;
ML=0.5+(4/@pi())*(hdl-0.5)*(hdl-hdl^2)^0.5+(1/@pi())*@asin(2*hdl-1);
HL=hdl*D;

Vw=Lflow*trw/60;
Aw=Vw/Leff;
```

```

Mw=Aw/Atot;
Mw=0.5+(4/@pi())*(hdw-0.5)*(hdw-hdw^2)^0.5+(1/@pi())*@asin(2*hdw-1);
How=hdw*D;

!Evans recommendation;
HL/D<=0.8;
D-HL>=0.3;

!Liquid Capacity-Water Drop in Oil Phase;
Vwd=(Hl-How)/(tro*60);
dwd=(1000000)*(((Vwd^2)*3*Cdw*Oden)/(4*g*(Lden-Oden)));
REw=((dwd/1000000)*Vwd*Oden)/(Ovis/1000);
Cdw=(24/REw)+(0.3/(REw^(1/2)))+0.34;
dwd<=250;

!Liquid Capacity-Oil Drop in Water Phase;
Voil=how/(trw*60);
doil=(1000000)*(((Voil^2)*3*Cdo*Lden)/(4*g*(Lden-Oden)));
REo=((doil/1000000)*Voil*Lden)/(Lvis/1000);
Cdo=(24/REo)+(0.3/(REo^(1/2)))+0.34;
doil<=300;
L=(4/3)*Leff;

i=0.1;!interest;
py=25; !paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;

!pressure in psig and 110%;
Pd=Pressure*14.5038;
CA=0.125;
tp=((Pd*D*1000/25.4)/(2*15000*1-1.2*Pd))+CA;

platethickness=(tp)*25.4/1000;

!vessel weight calc;
DryVesselWeight=@pi()*(D+platethickness)*(L+0.8*D)*platethickness*steelDen;
ActualVesselWeight=DryVesselWeight*1.15;

!Costing calc;
Cv=@exp(8.9552-
0.233*@log(ActualVesselWeight*2.20462)+0.04333*(@log(ActualVesselWeight*2.2
0462)^2));
ACv=anl*CV*704.1/500;

min=acv;

```

Scenario 2 (Simultaneous optimisation, Chapter 6)

Result (Electrostatic separator)

```

Global optimal solution found.
Objective value:                39949.49
Objective bound:                39949.49
Infeasibilities:                0.8953270E-07
Extended solver steps:         221
Total solver iterations:        1542599005
Elapsed runtime seconds:        9.88

```

```

Model Class:                    NLP

```

```

Total variables:                28
Nonlinear variables:            20
Integer variables:              0

```

```

Total constraints:              35
Nonlinear constraints:          21

```

```

Total nonzeros:                77
Nonlinear nonzeros:            34

```

Variable	Value	Reduced Cost
PRESSURE	45.000000	0.000000
TRW	5.000000	0.000000
TRO	5.000000	0.000000
G	9.810000	0.000000
STEELDEN	7849.054	0.000000
LDEN	968.3503	0.000000
ODEN	810.2671	0.000000
LFLOW	51.01251	0.000000
OFLOW	269.1117	0.000000
LVIS	0.3665118	0.000000
OVIS	1.699851	0.000000
LEFF	9.749652	0.000000
D	2.768860	0.000000
L	12.99954	0.000000
VL	26.67702	0.000000
AL	2.736202	0.000000
AMIN	2.736202	0.000000
DMIN	1.866505	0.000000
ATOT	6.021323	0.000000
ML	0.4544187	0.000000
HDL	0.4641699	0.000000
HL	1.285221	0.000000
VW	4.251043	0.000000
AW	0.4360199	0.000000

MW	0.7241265E-01	0.000000
HDW	0.1253155	0.000000
HOW	0.3469811	0.000000
VWD	0.3127378E-02	0.000000
DWD	250.0000	0.000000
CDW	65.22956	0.000000
REW	0.3726816	0.000000
VOIL	0.1156575E-02	0.000000
DOIL	70.45358	0.000000
CDO	112.4647	0.000000
REO	0.2152888	0.000000
I	0.1000000	0.000000
PY	25.00000	0.000000
ANL	0.1101681	0.000000
PD	652.6710	0.000000
CA	0.1250000	0.000000
TP	2.560169	0.000000
PLATETHICKNESS	0.6502828E-01	0.000000
DRYVESSELWEIGHT	69137.47	0.000000
ACTUALVESSELWEIGHT	79508.09	0.000000
CV	257508.2	0.1551387
ACV	39949.49	0.000000

Scenario 2 (Simultaneous optimisation, Chapter 6)

Mathematical formulation (HEN and Separators)

```

sets:
star:Tcin,Tcout,fi,FCC;
temp/1..6/:Tin,lukaku;
Lot/1..6/:P,T;
intervals:dx,Tkin,Tkout,th,rdx,LGT,heat,cold,adt,rheat,mun,Tw,TdifI;
Connections(temp,temp):Q;
Connections2(temp,lot):L;
Connections3(star,intervals):y,po;
mour/1..49/:sepcost1,Totalload1,fio,fit,fir,tout3,hex,TotalC,minAP,area,was
teHEX,minf,Tret,bin2;

ENDSETS
DATA:
m=3;
star=1..m;
Tcin=30.00,39.00,42.61;
n=5;
intervals=1..n;
P=5,4,3,2,1,0;

FCC=2,1,1.5;

Sepcost1,Totalload1,fio,fit,fir,Tout3=@OLE('combination.XLSx',
'sepcost1','Totalload1','fio','fit','fir','tout3');
ENDDATA

fi(1)=103.28422;
fi(2)=342.65839;
Tcout(1)=40.00;
Tcout(2)=49.00;

@sum(mour(c):bin2(c)*fir(c)*(Tout3(c)-(Tcin(3))))=totalload2;

@for(mour(c):@bin(bin2(C)));
@sum(mour(c):bin2(c))=1;
@sum(mour(c):bin2(c)*Tout3(c))=Tcout(3);
totalload=4459.4261+Totalload2;
!temp sorting;

fmin>=107.5163;
fmin<=18815.377;
@for(intervals(k):dx(k)<=Totalload*Tw(k));
@for(intervals(k):@bin (Tw(k)));

@for (temp (i)| i#LE#m:Tin(i)=Tcin(i));
@for (temp (i)| i#GT#m:Tin(i)=Tcout(i-m));

```

```

@for (temp (i):
@for (temp (j) | i#NE#j:Q(i,j)*Tin(i)>=Tin(j)*Q(i,j)));
@for (temp (i):
@for (temp (j) | i#NE#j:(1-Q(i,j))*Tin(i)<=Tin(j)));

@for (temp (i):
@for (temp (j) | i#NE#j:@bin (Q(i,j))));

@sum (temp (i):
@sum (temp (j) | i#NE#j:(Q(i,j)))=15;

@for (temp (i):
@for (lot (k):
@sum (temp (j) | i#NE#j:(Q(i,j)*L(i,k))<=P(k)));

@for (temp (i):
@for (lot (k):@bin (L(i,k))));

@for (temp (i):
@sum (lot (k):(L(i,k))=1);

@for (lot (k):
@sum (temp (i):(L(i,k))=1);

@for (lot (k):
@sum (temp (i):(L(i,k)*Tin(i))=T(k));

!temp in and out sorting;

@for(intervals(k):Tkin(k)=T(k));
@for(intervals(k):Tkout(k)=T(k+1));

!Cold inlet temp. location at interval;
@for (temp (i):
@sum (temp (j) | i#NE#j:(Q(i,j)))=lukaku(i));

@for(star(o):
@for(intervals(k):Tcout(o)>=y(o,k)*Tkin(k));

@for(star(o):
@for(intervals(k):@bin (y(o,k))));

@for(star(o):
@sum(intervals(k):y(o,k)=lukaku(o+m));

!Hot inlet temp.temp at interval;

@for(star(o):
@for(intervals(k):Tcin(o)>=(1-po(o,k))*Tkin(k));

@for(star(o):
@for(intervals(k):@bin (po(o,k))));

@for(star(o):
@sum(intervals(k):po(o,k)=(5-lukaku(o)));

TMIN>=10;
TMIN<=100;

THM=130;

```

```

!heat load at each level;
@for(intervals(k):
@sum (star(o):(Tkin(k)-Tkout(k))*y(o,k)*(po(o,k))*fi(o))=dx(k));

!heat load cascade;
adt(1)=dx(1);
@for(intervals (k)| k#LT#5:adt(k+1)=adt(k)+dx(k+1));

!energy limit;
@for(intervals (k):adt(k)<=fmin*(130-tmin-Tkout(k));

!reverse heat load summation;
@for(intervals (k)| k#LT#5:rheat(k+1)=rheat(k)-dx(k));
rheat(1)=Totalload;

!reverse heat load summation;
@sum(intervals (k):dx(k))=totalload;

!to calculate return temp;
THMR=(0-Totalload)/Fmin+thm;

!source temperature calc at turning point;
@for(intervals(k):((rheat(k)-Totalload)/fmin)+thm=th(k));

!to ensure min tmin;
@smin(Th(2)-Tkin(2),Th(3)-Tkin(3),Th(4)-Tkin(4),Th(5)-Tkin(5),THMR-
Tkout(5))=tmin;

!log mean temp diff;
@for(intervals(k):
(Th(k)-Tkin(k))*TW(k)+100*(1-TW(k))=Tdifi(k));

!log mean temp diff;
@for(intervals(k)| k#LT#5:
(Tdifi(k)-(Th(k+1)-Tkin(k+1)))/@log(Tdifi(k)/(Th(k+1)-Tkin(k+1)))=lgt(k));
(Tdifi(5)-(Thmr-30))/@log(Tdifi(5)/(Thmr-30))=lgt(5);

!hot stream area requirement;
FCH=2;
@for(intervals(k)| k#LT#5:((Fmin*(Th(k)-Th(k+1)))/(FCH*lgt(k)))=heat(k));
@for(intervals(k)| k#EQ#5:((Fmin*(Th(5)-Thmr))/(FCH*lgt(k)))=heat(k));

!cold stream area requirement;
@for(intervals(k)| k#LT#5:
@sum (star(o):((Tkin(k)-
Tkout(k+1))*fi(o)*y(o,k)*(po(o,k))/(FCC(o)*lgt(k))))=cold(k));
@for(intervals(k)| k#EQ#5:
@sum (star(o):((Tkin(k)-
30)*fi(o)*y(o,k)*(po(o,k))/(FCC(o)*lgt(k))))=cold(k));

!for Hex area calc;
@sum(intervals(k):cold(k)+heat(k))=sumAk;

!for Hex cost;
i=0.1;!interest;
py=25;!paybackyear;
anl=(i*(1+i)^py)/(((1+i)^py)-1);!annualized;
CI=40000*anl;

```

```

AC=500*anl;
CeHEX=(6.52/4.358);
CHEX=(CI+(AC*sumAk)/3)*3*CeHEX;

!for WHRU area calc;
GHI=433; !Hot gas in;
GHO=150; !Hot gas out;
GFCH=40; !Gas film heat transfer coefficient;
Lf=50; !Heat conductivity of the fin material;
bf=0.0003; !fin width;
hf=0.01; !fin height;
X=@sqrt ((2*GFCH)/(bf*Lf))*hf;
df=40; !fin diameter;
dc=20; !fin core diameter;
s=2; !tube wall thickness of the core tube;
Xdf=X*(1+0.35*@log(df/dc));
nf=@tanh(xdf)/xdf; !fin efficiency;
Ai=0.05; !internal total surface area;
Ao=0.55; !total surface area;
Act=0.07; !core tube surface area;
Af=0.48; !fin surface area;
nw=((nf*Af+Act)/Ao); !weighted fin efficiency;
GFCHw= nw*GFCH*(Ao/Ai); !weighted film heat transfer coefficient;
WFCH=2000; !water film heat transfer coefficient;
Ui=1/(1/GFCHw+1/WFCH+s/(1000*50)); !overall film heat transfer coefficient;
lgtWHRU=((GHI-THM)-(GHO-THMR))/@log((GHI-THM)/(GHO-THMR));
AWHRU=(1000*Totalload)/(lgtWHRU*Ui);

!for WHRU cost;
bc=156000; !base cost;
bs=200; !base size;
mtt=0.89;
CWHRU=(bc*(AWHRU/bs)^mtt)*anl*CeHEX;

!for pump power calc;
lds=984.12; !liquid desnsity;
lhc=4.326; !liquid heat capacity;
fl=(fmin*1.1/lhc)*3600/lds; !flowrate;
h=51.7908381; !pumping height;
eff=0.6; !efficiency;
Ps=(fl*lds*9.81*h)/(3600000*0.6);

!pump cost;
bcp=9840; !base cost;
bsp=4; !base size;
mp=0.55;
CePump=(9.41/4.358);
CPump=(bcp*(Ps/bsp)^mp)*anl*CePump;

!Utility cost;
UCPump=Ps*365*24*0.12;

@sum(mour(c):bin2(c)*sepcost1)=sepcostE;
!Total cost;
HENCost=CHEx+UCPump+CWHRU+Cpump;
Totalcost=HENCost+(SepcostE)*anl;

min= totalcost;

data:

```

```
@OLE('combination.XLSx',
'hex','TotalC','minAP','area','wasteHEX','minf','tret')=hex,TotalC,minAP,ar
ea,wasteHEX,minf,Tret;
enddata
```

Scenario 2 (Simultaneous optimisation, Chapter 6)

Result (HEN and Separators)

```
Global optimal solution found.
Objective value:                182807.0
Objective bound:                182807.0
Infeasibilities:                0.000000
Extended solver steps:          3419
Total solver iterations:        1904908
Elapsed runtime seconds:        177.37
```

Export Summary Report

```
-----
Transfer Method:                OLE BASED
Workbook:                      combination.XLSx
Ranges Specified:              7
    hex
    TotalC
    minAP
    area
    wasteHEX
    minf
    tret
Ranges Found:                  7
Range Size Mismatches:         7
Values Transferred:            343
```

```
Model Class:                   MINLP
```

```
Total variables:               591
Nonlinear variables:           140
Integer variables:             150

Total constraints:              242
Nonlinear constraints:          133

Total nonzeros:                1118
Nonlinear nonzeros:            518
```

Variable	Value	Reduced Cost
M	3.000000	0.000000
N	5.000000	0.000000
TOTALLOAD2	6097.393	0.000000
TOTALLOAD	10556.82	0.000000
FMIN	140.9831	0.000000

TMIN	22.71339	0.000000
THM	130.0000	0.000000
THMR	55.11999	0.000000
FCH	2.000000	0.000000
SUMAK	358.0875	0.000000
I	0.1000000	0.000000
PY	25.00000	0.000000
ANL	0.1101681	0.000000
CI	4406.723	0.000000
AC	55.08404	0.000000
CEHEX	1.496099	0.000000
CHEX	49289.10	0.000000
GHI	433.0000	0.000000
GHO	150.0000	0.000000
GFCH	40.00000	0.000000
LF	50.00000	0.000000
BF	0.3000000E-03	0.000000
HF	0.1000000E-01	0.000000
X	0.7302967	0.000000
DF	40.00000	0.000000
DC	20.00000	0.000000
S	2.000000	0.000000
XDF	0.9074678	0.000000
NF	0.7933225	0.000000
AI	0.5000000E-01	0.000000
AO	0.5500000	0.000000
ACT	0.7000000E-01	0.000000
AF	0.4800000	0.000000
NW	0.8196269	0.000000
GFCHW	360.6358	0.000000
WFCH	2000.000	0.000000
UI	301.8521	0.000000
LGTWHRU	179.2408	0.000000
AWHRU	195.1201	0.000000
BC	156000.0	0.000000
BS	200.0000	0.000000
MTT	0.8900000	0.000000
CWHRU	25153.17	0.000000
LDS	984.1200	0.000000
LHC	4.326000	0.000000
FL	131.1378	0.000000
H	51.79084	0.000000
EFF	0.6000000	0.000000
PS	30.35596	0.000000
BCP	9840.000	0.000000
BSP	4.000000	0.000000
MP	0.5500000	0.000000
CEPUMP	2.159247	0.000000
CPUMP	7135.994	0.000000
UCPUMP	31910.19	0.000000
SEPCOSTE	629206.9	0.000000
HENCOST	113488.5	0.000000
TOTALCOST	182807.0	0.000000
TCIN(1)	30.00000	0.000000
TCIN(2)	39.00000	0.000000
TCIN(3)	42.61000	0.000000
TCOUT(1)	40.00000	0.000000
TCOUT(2)	49.00000	0.000000
TCOUT(3)	76.61000	0.000000
FI(1)	103.2842	0.000000
FI(2)	342.6584	0.000000

FI (3)	179.3351	0.000000
FCC (1)	2.000000	0.000000
FCC (2)	1.000000	0.000000
FCC (3)	1.500000	0.000000
TIN (1)	30.00000	0.000000
TIN (2)	39.00000	0.000000
TIN (3)	42.61000	0.000000
TIN (4)	40.00000	0.000000
TIN (5)	49.00000	0.000000
TIN (6)	76.61000	0.000000
LUKAKU (1)	0.000000	0.000000
LUKAKU (2)	1.000000	0.000000
LUKAKU (3)	3.000000	0.000000
LUKAKU (4)	2.000000	0.000000
LUKAKU (5)	4.000000	0.000000
LUKAKU (6)	5.000000	0.000000
P (1)	5.000000	0.000000
P (2)	4.000000	0.000000
P (3)	3.000000	0.000000
P (4)	2.000000	0.000000
P (5)	1.000000	0.000000
P (6)	0.000000	0.000000
T (1)	76.61000	0.000000
T (2)	49.00000	0.000000
T (3)	42.61000	0.000000
T (4)	40.00000	0.000000
T (5)	39.00000	0.000000
T (6)	30.00000	0.000000
DX (1)	4951.441	0.000000
DX (2)	3335.538	0.000000
DX (3)	894.3384	0.000000
DX (4)	445.9426	0.000000
DX (5)	929.5580	0.000000
TKIN (1)	76.61000	0.000000
TKIN (2)	49.00000	0.000000
TKIN (3)	42.61000	0.000000
TKIN (4)	40.00000	0.000000
TKIN (5)	39.00000	0.000000
TKOUT (1)	49.00000	0.000000
TKOUT (2)	42.61000	0.000000
TKOUT (3)	40.00000	0.000000
TKOUT (4)	39.00000	0.000000
TKOUT (5)	30.00000	0.000000
TH (1)	130.0000	0.000000
TH (2)	94.87919	0.000000
TH (3)	71.22006	0.000000
TH (4)	64.87648	0.000000
TH (5)	61.71339	0.000000
RDX (1)	1.234568	0.000000
RDX (2)	1.234568	0.000000
RDX (3)	1.234568	0.000000
RDX (4)	1.234568	0.000000
RDX (5)	1.234568	0.000000
LGT (1)	49.53974	0.000000
LGT (2)	36.56752	0.000000
LGT (3)	26.69978	0.000000
LGT (4)	23.77854	0.000000
LGT (5)	23.89649	0.000000
HEAT (1)	49.97444	0.000000
HEAT (2)	45.60793	0.000000
HEAT (3)	16.74805	0.000000

HEAT(4)	9.376999	0.000000
HEAT(5)	19.44967	0.000000
COLD(1)	66.63259	0.000000
COLD(2)	80.76988	0.000000
COLD(3)	33.49610	0.000000
COLD(4)	16.58220	0.000000
COLD(5)	19.44967	0.1495010E-08
ADT(1)	4951.441	0.000000
ADT(2)	8286.980	0.000000
ADT(3)	9181.318	0.000000
ADT(4)	9627.261	0.000000
ADT(5)	10556.82	0.000000
RHEAT(1)	10556.82	0.000000
RHEAT(2)	5605.377	0.000000
RHEAT(3)	2269.839	0.000000
RHEAT(4)	1375.501	0.000000
RHEAT(5)	929.5580	0.000000
MUN(1)	1.234568	0.000000
MUN(2)	1.234568	0.000000
MUN(3)	1.234568	0.000000
MUN(4)	1.234568	0.000000
MUN(5)	1.234568	0.000000
TW(1)	1.000000	4300.624
TW(2)	1.000000	6624.621
TW(3)	1.000000	5286.315
TW(4)	1.000000	3279.299
TW(5)	1.000000	5362.569
TDIFI(1)	53.39000	0.000000
TDIFI(2)	45.87919	0.000000
TDIFI(3)	28.61006	0.000000
TDIFI(4)	24.87648	0.000000
TDIFI(5)	22.71339	0.000000
Q(1, 1)	1.234568	0.000000
Q(1, 2)	0.000000	0.000000
Q(1, 3)	0.000000	0.000000
Q(1, 4)	0.000000	0.000000
Q(1, 5)	0.000000	0.000000
Q(1, 6)	0.000000	0.000000
Q(2, 1)	1.000000	-878.9012
Q(2, 2)	0.000000	0.000000
Q(2, 3)	0.000000	0.000000
Q(2, 4)	0.000000	0.000000
Q(2, 5)	0.000000	0.000000
Q(2, 6)	0.000000	0.000000
Q(3, 1)	1.000000	-0.4458119E-02
Q(3, 2)	1.000000	-0.4458119E-02
Q(3, 3)	0.000000	0.000000
Q(3, 4)	1.000000	-0.4458119E-02
Q(3, 5)	0.000000	0.000000
Q(3, 6)	0.000000	0.000000
Q(4, 1)	1.000000	-1377.963
Q(4, 2)	1.000000	-1377.963
Q(4, 3)	0.000000	0.000000
Q(4, 4)	1.234568	0.000000
Q(4, 5)	0.000000	0.000000
Q(4, 6)	0.000000	0.000000
Q(5, 1)	1.000000	-716.3235
Q(5, 2)	1.000000	-716.3235
Q(5, 3)	1.000000	-716.3235
Q(5, 4)	1.000000	-716.3235
Q(5, 5)	0.000000	0.000000

Q(5, 6)	0.000000	0.000000
Q(6, 1)	1.000000	0.000000
Q(6, 2)	1.000000	0.000000
Q(6, 3)	1.000000	0.000000
Q(6, 4)	1.000000	0.000000
Q(6, 5)	1.000000	0.000000
Q(6, 6)	1.234568	0.000000
L(1, 1)	0.000000	1926.398
L(1, 2)	0.000000	2612.489
L(1, 3)	0.000000	777.4088
L(1, 4)	0.000000	8854.236
L(1, 5)	0.000000	0.000000
L(1, 6)	1.000000	-18568.34
L(2, 1)	0.000000	406.6744
L(2, 2)	0.000000	603.0890
L(2, 3)	0.000000	77.74088
L(2, 4)	0.000000	4052.905
L(2, 5)	1.000000	0.000000
L(2, 6)	0.000000	0.000000
L(3, 1)	0.000000	0.000000
L(3, 2)	0.000000	0.000000
L(3, 3)	1.000000	0.000000
L(3, 4)	0.000000	2329.941
L(3, 5)	0.000000	202.9037
L(3, 6)	0.000000	0.000000
L(4, 1)	0.000000	237.8162
L(4, 2)	0.000000	379.8223
L(4, 3)	0.000000	0.000000
L(4, 4)	1.000000	3519.424
L(4, 5)	0.000000	0.000000
L(4, 6)	0.000000	0.000000
L(5, 1)	0.000000	0.000000
L(5, 2)	1.000000	0.000000
L(5, 3)	0.000000	582.2396
L(5, 4)	0.000000	0.000000
L(5, 5)	0.000000	1281.908
L(5, 6)	0.000000	0.000000
L(6, 1)	1.000000	1849.889
L(6, 2)	0.000000	0.000000
L(6, 3)	0.000000	4947.877
L(6, 4)	0.000000	-8217.354
L(6, 5)	0.000000	7793.971
L(6, 6)	0.000000	0.000000
Y(1, 1)	0.000000	230.6462
Y(1, 2)	0.000000	-195.4003
Y(1, 3)	0.000000	0.000000
Y(1, 4)	1.000000	278.8102
Y(1, 5)	1.000000	-2564.955
Y(2, 1)	0.000000	6815.335
Y(2, 2)	1.000000	0.000000
Y(2, 3)	1.000000	-438.8356
Y(2, 4)	1.000000	-300.2401
Y(2, 5)	1.000000	-37.83504
Y(3, 1)	1.000000	0.000000
Y(3, 2)	1.000000	-1682.138
Y(3, 3)	1.000000	-841.0711
Y(3, 4)	1.000000	-841.0711
Y(3, 5)	1.000000	-841.0756
PO(1, 1)	1.000000	0.000000
PO(1, 2)	1.000000	-0.7322926E-03
PO(1, 3)	1.000000	0.000000

PO(1, 4)	1.000000	-258.0818
PO(1, 5)	1.000000	-3101.847
PO(2, 1)	1.000000	-37.83010
PO(2, 2)	1.000000	0.000000
PO(2, 3)	1.000000	-438.8357
PO(2, 4)	1.000000	-300.2401
PO(2, 5)	0.000000	-5010.875
PO(3, 1)	1.000000	1682.139
PO(3, 2)	1.000000	0.000000
PO(3, 3)	0.000000	149.6290
PO(3, 4)	0.000000	496.5448
PO(3, 5)	0.000000	-3617.052
SEPCOST1(1)	734333.2	0.000000
SEPCOST1(2)	709055.2	0.000000
SEPCOST1(3)	686143.8	0.000000
SEPCOST1(4)	665336.3	0.000000
SEPCOST1(5)	646430.1	0.000000
SEPCOST1(6)	629206.9	0.000000
SEPCOST1(7)	613595.1	0.000000
SEPCOST1(8)	599376.6	0.000000
SEPCOST1(9)	586368.0	0.000000
SEPCOST1(10)	574493.4	0.000000
SEPCOST1(11)	563613.9	0.000000
SEPCOST1(12)	553647.8	0.000000
SEPCOST1(13)	544513.0	0.000000
SEPCOST1(14)	536129.5	0.000000
SEPCOST1(15)	528443.5	0.000000
SEPCOST1(16)	521386.7	0.000000
SEPCOST1(17)	514917.0	0.000000
SEPCOST1(18)	508971.9	0.000000
SEPCOST1(19)	503514.3	0.000000
SEPCOST1(20)	498500.1	0.000000
SEPCOST1(21)	493899.6	0.000000
SEPCOST1(22)	489681.7	0.000000
SEPCOST1(23)	486074.5	0.000000
SEPCOST1(24)	486256.6	0.000000
SEPCOST1(25)	486439.4	0.000000
SEPCOST1(26)	486623.0	0.000000
SEPCOST1(27)	486807.4	0.000000
SEPCOST1(28)	486992.7	0.000000
SEPCOST1(29)	487178.8	0.000000
SEPCOST1(30)	487365.7	0.000000
SEPCOST1(31)	487553.5	0.000000
SEPCOST1(32)	487740.9	0.000000
SEPCOST1(33)	487928.2	0.000000
SEPCOST1(34)	488116.3	0.000000
SEPCOST1(35)	488305.1	0.000000
SEPCOST1(36)	488494.9	0.000000
SEPCOST1(37)	488685.4	0.000000
SEPCOST1(38)	488876.9	0.000000
SEPCOST1(39)	489069.2	0.000000
SEPCOST1(40)	489262.3	0.000000
SEPCOST1(41)	489434.6	0.000000
SEPCOST1(42)	489589.1	0.000000
SEPCOST1(43)	489743.2	0.000000
SEPCOST1(44)	489897.1	0.000000
SEPCOST1(45)	490050.9	0.000000
SEPCOST1(46)	490204.5	0.000000
SEPCOST1(47)	490358.0	0.000000
SEPCOST1(48)	490511.2	0.000000
SEPCOST1(49)	490664.2	0.000000

TOTALLOAD1 (1)	9638.381	0.000000
TOTALLOAD1 (2)	9821.447	0.000000
TOTALLOAD1 (3)	10004.82	0.000000
TOTALLOAD1 (4)	10188.51	0.000000
TOTALLOAD1 (5)	10372.51	0.000000
TOTALLOAD1 (6)	10556.82	0.000000
TOTALLOAD1 (7)	10741.44	0.000000
TOTALLOAD1 (8)	10926.38	0.000000
TOTALLOAD1 (9)	11111.64	0.000000
TOTALLOAD1 (10)	11297.22	0.000000
TOTALLOAD1 (11)	11483.11	0.000000
TOTALLOAD1 (12)	11669.33	0.000000
TOTALLOAD1 (13)	11855.86	0.000000
TOTALLOAD1 (14)	12042.72	0.000000
TOTALLOAD1 (15)	12229.91	0.000000
TOTALLOAD1 (16)	12417.42	0.000000
TOTALLOAD1 (17)	12605.25	0.000000
TOTALLOAD1 (18)	12793.42	0.000000
TOTALLOAD1 (19)	12981.91	0.000000
TOTALLOAD1 (20)	13170.74	0.000000
TOTALLOAD1 (21)	13359.90	0.000000
TOTALLOAD1 (22)	13549.39	0.000000
TOTALLOAD1 (23)	13739.22	0.000000
TOTALLOAD1 (24)	13929.39	0.000000
TOTALLOAD1 (25)	14119.89	0.000000
TOTALLOAD1 (26)	14310.74	0.000000
TOTALLOAD1 (27)	14501.93	0.000000
TOTALLOAD1 (28)	14693.46	0.000000
TOTALLOAD1 (29)	14885.33	0.000000
TOTALLOAD1 (30)	15077.55	0.000000
TOTALLOAD1 (31)	15270.12	0.000000
TOTALLOAD1 (32)	15463.04	0.000000
TOTALLOAD1 (33)	15656.31	0.000000
TOTALLOAD1 (34)	15849.93	0.000000
TOTALLOAD1 (35)	16043.91	0.000000
TOTALLOAD1 (36)	16238.24	0.000000
TOTALLOAD1 (37)	16432.93	0.000000
TOTALLOAD1 (38)	16627.99	0.000000
TOTALLOAD1 (39)	16823.40	0.000000
TOTALLOAD1 (40)	17019.18	0.000000
TOTALLOAD1 (41)	17216.42	0.000000
TOTALLOAD1 (42)	17414.76	0.000000
TOTALLOAD1 (43)	17613.52	0.000000
TOTALLOAD1 (44)	17812.72	0.000000
TOTALLOAD1 (45)	18012.35	0.000000
TOTALLOAD1 (46)	18212.43	0.000000
TOTALLOAD1 (47)	18412.95	0.000000
TOTALLOAD1 (48)	4459.426	0.000000
TOTALLOAD1 (49)	4459.426	0.000000
FIO (1)	103.2842	0.000000
FIO (2)	103.2842	0.000000
FIO (3)	103.2842	0.000000
FIO (4)	103.2842	0.000000
FIO (5)	103.2842	0.000000
FIO (6)	103.2842	0.000000
FIO (7)	103.2842	0.000000
FIO (8)	103.2842	0.000000
FIO (9)	103.2842	0.000000
FIO (10)	103.2842	0.000000
FIO (11)	103.2842	0.000000
FIO (12)	103.2842	0.000000

FIO(13)	103.2842	0.000000
FIO(14)	103.2842	0.000000
FIO(15)	103.2842	0.000000
FIO(16)	103.2842	0.000000
FIO(17)	103.2842	0.000000
FIO(18)	103.2842	0.000000
FIO(19)	103.2842	0.000000
FIO(20)	103.2842	0.000000
FIO(21)	103.2842	0.000000
FIO(22)	103.2842	0.000000
FIO(23)	103.2842	0.000000
FIO(24)	103.2842	0.000000
FIO(25)	103.2842	0.000000
FIO(26)	103.2842	0.000000
FIO(27)	103.2842	0.000000
FIO(28)	103.2842	0.000000
FIO(29)	103.2842	0.000000
FIO(30)	103.2842	0.000000
FIO(31)	103.2842	0.000000
FIO(32)	103.2842	0.000000
FIO(33)	103.2842	0.000000
FIO(34)	103.2842	0.000000
FIO(35)	103.2842	0.000000
FIO(36)	103.2842	0.000000
FIO(37)	103.2842	0.000000
FIO(38)	103.2842	0.000000
FIO(39)	103.2842	0.000000
FIO(40)	103.2842	0.000000
FIO(41)	103.2842	0.000000
FIO(42)	103.2842	0.000000
FIO(43)	103.2842	0.000000
FIO(44)	103.2842	0.000000
FIO(45)	103.2842	0.000000
FIO(46)	103.2842	0.000000
FIO(47)	103.2842	0.000000
FIO(48)	103.2842	0.000000
FIO(49)	103.2842	0.000000
FIT(1)	342.6584	0.000000
FIT(2)	342.6584	0.000000
FIT(3)	342.6584	0.000000
FIT(4)	342.6584	0.000000
FIT(5)	342.6584	0.000000
FIT(6)	342.6584	0.000000
FIT(7)	342.6584	0.000000
FIT(8)	342.6584	0.000000
FIT(9)	342.6584	0.000000
FIT(10)	342.6584	0.000000
FIT(11)	342.6584	0.000000
FIT(12)	342.6584	0.000000
FIT(13)	342.6584	0.000000
FIT(14)	342.6584	0.000000
FIT(15)	342.6584	0.000000
FIT(16)	342.6584	0.000000
FIT(17)	342.6584	0.000000
FIT(18)	342.6584	0.000000
FIT(19)	342.6584	0.000000
FIT(20)	342.6584	0.000000
FIT(21)	342.6584	0.000000
FIT(22)	342.6584	0.000000
FIT(23)	342.6584	0.000000
FIT(24)	342.6584	0.000000

FIT(25)	342.6584	0.000000
FIT(26)	342.6584	0.000000
FIT(27)	342.6584	0.000000
FIT(28)	342.6584	0.000000
FIT(29)	342.6584	0.000000
FIT(30)	342.6584	0.000000
FIT(31)	342.6584	0.000000
FIT(32)	342.6584	0.000000
FIT(33)	342.6584	0.000000
FIT(34)	342.6584	0.000000
FIT(35)	342.6584	0.000000
FIT(36)	342.6584	0.000000
FIT(37)	342.6584	0.000000
FIT(38)	342.6584	0.000000
FIT(39)	342.6584	0.000000
FIT(40)	342.6584	0.000000
FIT(41)	342.6584	0.000000
FIT(42)	342.6584	0.000000
FIT(43)	342.6584	0.000000
FIT(44)	342.6584	0.000000
FIT(45)	342.6584	0.000000
FIT(46)	342.6584	0.000000
FIT(47)	342.6584	0.000000
FIT(48)	342.6584	0.000000
FIT(49)	342.6584	0.000000
FIR(1)	178.5846	0.000000
FIR(2)	178.7340	0.000000
FIR(3)	178.8838	0.000000
FIR(4)	179.0339	0.000000
FIR(5)	179.1843	0.000000
FIR(6)	179.3351	0.000000
FIR(7)	179.4862	0.000000
FIR(8)	179.6377	0.000000
FIR(9)	179.7896	0.000000
FIR(10)	179.9419	0.000000
FIR(11)	180.0945	0.000000
FIR(12)	180.2475	0.000000
FIR(13)	180.4009	0.000000
FIR(14)	180.5547	0.000000
FIR(15)	180.7088	0.000000
FIR(16)	180.8634	0.000000
FIR(17)	181.0184	0.000000
FIR(18)	181.1737	0.000000
FIR(19)	181.3295	0.000000
FIR(20)	181.4857	0.000000
FIR(21)	181.6423	0.000000
FIR(22)	181.7994	0.000000
FIR(23)	181.9568	0.000000
FIR(24)	182.1147	0.000000
FIR(25)	182.2730	0.000000
FIR(26)	182.4317	0.000000
FIR(27)	182.5909	0.000000
FIR(28)	182.7505	0.000000
FIR(29)	182.9106	0.000000
FIR(30)	183.0711	0.000000
FIR(31)	183.2321	0.000000
FIR(32)	183.3935	0.000000
FIR(33)	183.5555	0.000000
FIR(34)	183.7178	0.000000
FIR(35)	183.8807	0.000000
FIR(36)	184.0440	0.000000

FIR(37)	184.2078	0.000000
FIR(38)	184.3721	0.000000
FIR(39)	184.5369	0.000000
FIR(40)	184.7022	0.000000
FIR(41)	184.8840	0.000000
FIR(42)	185.0761	0.000000
FIR(43)	185.2689	0.000000
FIR(44)	185.4624	0.000000
FIR(45)	185.6565	0.000000
FIR(46)	185.8514	0.000000
FIR(47)	186.0470	0.000000
FIR(48)	186.2435	0.000000
FIR(49)	186.4409	0.000000
TOUT3(1)	71.61000	0.000000
TOUT3(2)	72.61000	0.000000
TOUT3(3)	73.61000	0.000000
TOUT3(4)	74.61000	0.000000
TOUT3(5)	75.61000	0.000000
TOUT3(6)	76.61000	0.000000
TOUT3(7)	77.61000	0.000000
TOUT3(8)	78.61000	0.000000
TOUT3(9)	79.61000	0.000000
TOUT3(10)	80.61000	0.000000
TOUT3(11)	81.61000	0.000000
TOUT3(12)	82.61000	0.000000
TOUT3(13)	83.61000	0.000000
TOUT3(14)	84.61000	0.000000
TOUT3(15)	85.61000	0.000000
TOUT3(16)	86.61000	0.000000
TOUT3(17)	87.61000	0.000000
TOUT3(18)	88.61000	0.000000
TOUT3(19)	89.61000	0.000000
TOUT3(20)	90.61000	0.000000
TOUT3(21)	91.61000	0.000000
TOUT3(22)	92.61000	0.000000
TOUT3(23)	93.61000	0.000000
TOUT3(24)	94.61000	0.000000
TOUT3(25)	95.61000	0.000000
TOUT3(26)	96.61000	0.000000
TOUT3(27)	97.61000	0.000000
TOUT3(28)	98.61000	0.000000
TOUT3(29)	99.61000	0.000000
TOUT3(30)	100.6100	0.000000
TOUT3(31)	101.6100	0.000000
TOUT3(32)	102.6100	0.000000
TOUT3(33)	103.6100	0.000000
TOUT3(34)	104.6100	0.000000
TOUT3(35)	105.6100	0.000000
TOUT3(36)	106.6100	0.000000
TOUT3(37)	107.6100	0.000000
TOUT3(38)	108.6100	0.000000
TOUT3(39)	109.6100	0.000000
TOUT3(40)	110.6100	0.000000
TOUT3(41)	111.6100	0.000000
TOUT3(42)	112.6100	0.000000
TOUT3(43)	113.6100	0.000000
TOUT3(44)	114.6100	0.000000
TOUT3(45)	115.6100	0.000000
TOUT3(46)	116.6100	0.000000
TOUT3(47)	117.6100	0.000000
TOUT3(48)	118.6100	0.000000

TOUT3 (49)	119.6100	0.000000
BIN2 (1)	0.000000	2988.563
BIN2 (2)	0.000000	1916.945
BIN2 (3)	0.000000	1108.727
BIN2 (4)	0.000000	534.9937
BIN2 (5)	0.000000	173.4308
BIN2 (6)	1.000000	0.000000
BIN2 (7)	0.000000	6.824259
BIN2 (8)	0.000000	169.8952
BIN2 (9)	0.000000	468.9909
BIN2 (10)	0.000000	895.7886
BIN2 (11)	0.000000	1434.988
BIN2 (12)	0.000000	2077.608
BIN2 (13)	0.000000	2814.595
BIN2 (14)	0.000000	3637.156
BIN2 (15)	0.000000	4539.384
BIN2 (16)	0.000000	5513.751
BIN2 (17)	0.000000	6555.656
BIN2 (18)	0.000000	7658.200
BIN2 (19)	0.000000	8817.315
BIN2 (20)	0.000000	10028.16
BIN2 (21)	0.000000	11287.48
BIN2 (22)	0.000000	12591.84
BIN2 (23)	0.000000	13966.40
BIN2 (24)	0.000000	15761.35
BIN2 (25)	0.000000	17559.32
BIN2 (26)	0.000000	19360.33
BIN2 (27)	0.000000	21164.40
BIN2 (28)	0.000000	22971.55
BIN2 (29)	0.000000	24781.78
BIN2 (30)	0.000000	26595.11
BIN2 (31)	0.000000	28411.55
BIN2 (32)	0.000000	30230.99
BIN2 (33)	0.000000	32053.48
BIN2 (34)	0.000000	33879.12
BIN2 (35)	0.000000	35707.93
BIN2 (36)	0.000000	37539.92
BIN2 (37)	0.000000	39375.12
BIN2 (38)	0.000000	41213.54
BIN2 (39)	0.000000	43055.20
BIN2 (40)	0.000000	44900.11
BIN2 (41)	0.000000	46755.46
BIN2 (42)	0.000000	48618.35
BIN2 (43)	0.000000	50484.90
BIN2 (44)	0.000000	52355.21
BIN2 (45)	0.000000	54229.29
BIN2 (46)	0.000000	56107.22
BIN2 (47)	0.000000	57989.02
BIN2 (48)	0.000000	59874.74
BIN2 (49)	0.000000	61764.22