Report of Cooperative Education

Design the Level Control System of BFW
and Study of the Vacuum System of a Hydrodesulfurization Unit

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ABSTRACT

The objectives of this project are to design level control system of a boiler feed water (BFW) tank and study the vacuum system of gas oil hydrodesulfurization unit in a refinery plant. The BFW system consists of the BFW tank to receive the condensate in a refinery plant, and the deaerator tank to remove the oxygen gas by using steam. So far, the BFW overflows because the amount of inlet condensate is higher than the design condition. Furthermore, the company is planning to start a new cogeneration plant which will increase the condensate in the future. To resolve the BFW overflow, the level control system of the BFW tank was studied. Split range control is selected because the output of the controller, that occur simultaneously, can be split and sent to two control valves. The level control system of the BFW was modified without the investment cost by using one controller (211LIC004) to split the signals to the control valve (211PV002) for make up mode, and to the control valve (211FV016) for excess mode. In case of the vacuum system of the gas oil hydrodesulfurization, it is responsible for water removal in the diesel gas oil. The vacuum system consists of three main components which are steam ejector to generate the vacuum condition, heat exchanger to cool the stream to liquid phase (3E-3711), and liquid ring vacuum pump. The operating pressure of the vacuum system is currently higher than the design pressure due to the back pressure in the vacuum system. The cause of back pressure is the phase of the outlet stream before sending to liquid ring vacuum pump that is not liquid phase. To resolve the high pressure problem, the installation of a new heat exchanger, that parallel with heat exchanger

(3E-3711), is an appropriate method by cooling the outlet temperature of the stream to saturated temperature (41.4 °C) for changing to liquid phase. The new type of heat exchanger is shell and tube type which having heat transfer area of 30.2 m^2. After installation of a new heat exchanger, the outlet temperature of the stream is 34.6 °C that the stream before sending to liquid ring vacuum pump are absolutely liquid phase. Therefore, the operating pressure will be decreased or closed to design pressure because the back pressure does not occur.

Keywords: Boiler feed water tank, Split Range Control, Vacuum System, Back pressure

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Kanokpon Maungthong

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CHAPTER I

INTRODUCTION

1.1 Background

Bangchak Corporation Public Company Limited was established to provide mainly energy and complex refinery business since procurement of crude oil from domestic and overseas sources. They were refined into various standard products and sold through Bangchak gas station. The company has expanded the business to produce electricity from solar cells which is clean energy from environment, biological energy and innovation-related business.

Control systems are very significant for industrial processes which are interested by Bangchak Corporation Public Company Limited. They are responsible for monitoring the production and investigating performance to follow in the objectives of process. It can verify weakness, and then the process will be improved and developed for following to the code and standard or according to the law. Control systems are applied to control physical quality currently for example temperature, pressure, flow rate, level and so on. For this reason, the control systems need to design seriously for appropriate process.

Boiler system in plant 3 has been used level control system for several tanks such as boiler feed water tank which is used for keeping the condensate from other plants. Nowadays Bangchak Corporation Public Company Limited is planning to install a new cogeneration plant which the amount of inlet condensate is higher than design condition. The boiler feed water tank cannot control the level of condensate that occurs overflow. It must drain some condensate for solving overflow problem. Therefore, the boiler feed water tank must have level control system for collecting the amount of condensate and solving overflow problems.

The vacuum system is frequently required in many industrial manufacturing process as well as in research fields such as vacuum dryer because the vacuum system can be reduce the boiling point of water that is separated easily. In the gas oil hydrodesulfurization unit (DGO-HDSU) in plant 3 have been used vacuum dryer for removing water from diesel gas oil until the water content is lower than 120 mg/L. Vacuum dryer can be operated at vacuum condition by using 3 main components which are steam ejector to generate the vacuum condition, heat exchanger to cool the stream to liquid phase (3E-3711), and liquid ring vacuum pump. Nowadays the operating pressure of the vacuum dryer is higher than the design pressure approximately 0.10 barg that effects the water content in diesel oil. After study of root cause analysis, the cause of high pressure in the vacuum dryer is that the outlet stream is not liquid phase because the outlet temperature of the stream from heat exchanger (3E-3711) is higher than the saturated temperature (41.4 °C) at pressure 0.14 barg. After that the outlet stream will be sent to the vacuum pump that is liquid ring. The back pressure is occurred in the vacuum pump, and then the back pressure is blocked the steam that is used in the steam ejectors until the operating pressure is high. Therefore, the objective of this work is to modify the vacuum system to solve the high pressure problem by decreasing the outlet temperature of the stream until it is lower than 41.4 °C.

1.2 Objectives

1.2.1 To design level control system of the boiler feed water tank

1.2.2 To modify the gas oil hydrodesulfurization unit for decreasing the outlet temperature of the stream

1.3 Scope of Work

1.3.1 Study the boiler feed water system and principles of control system for level control system design at the boiler feed water tank for instance, cascade control, split range control and override control

1.3.2 Select type of level control system for the boiler feed water tank

1.3.3 Study the process of gas oil hydrodesulfurization unit and the principles of heat transfer

1.3.4 Modify the gas oil hydrodesulfurization unit by installation of a new heat exchanger to cool down the outlet stream from the heat exchanger (3E-3711)

1.4 Expected Outputs

1.4.1 Understand the principles of control system including application of cascade control, split range control, and override control

1.4.2 Provide details of level control design for collecting the amount of condensate from a new cogeneration plant

1.4.3 Understand the principles of heat transfer and vacuum system in gas oil hydrodesulfurization unit including modification of the vacuum system in the vacuum dryer

CHAPTER II

LITERATURE REVIEW

PART I Design the Level Control System of BFW

2.1 Boiler System [Bangchak Petroleum Public Co., LTD, 2017]

A boiler system is the central heating up section in the industrial manufacturing by using the boiling water and steam that are used to heat in several units such as distillation column.

2.1.1 Process Description of the Boiler System

The process of the boiler system consists of 4 parts as follows

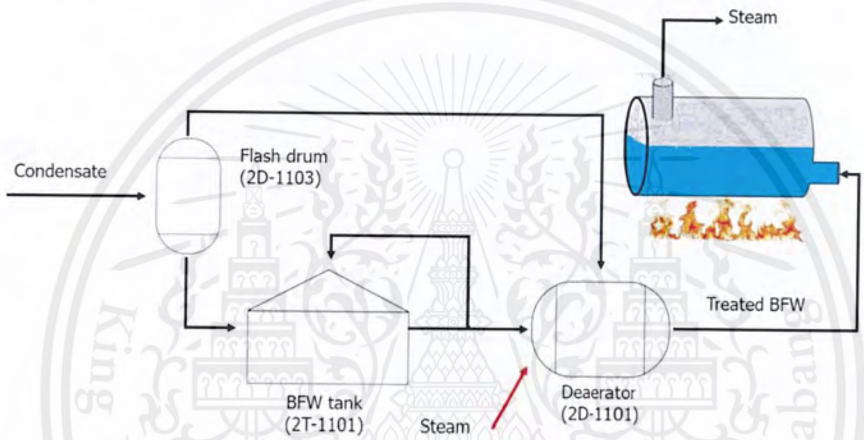


Figure 2.1 Boiler System Diagram

2.1.1.1 Flash drum is the equipment that separates liquid and vapor phases of the condensate from several units in the refinery plant. The liquid phase will be sent to the boiler feed water tank for storaging and the vapor phase will be sent to the deaerator tank. Nowadays the amount of inlet condensate is approximately 65,654 kg/h

2.1.1.2 Boiler feed water tank is the equipment that stroages the liquid phase of condensate from flash drum, and then the condensate will be sent to the deaerator tank. The characteristics of boiler feed water tank is the cone roof tank and the capacity is approximately 100 cubic meter.

2.1.1.3 Deaerator tank is the equipment that removes some gas dissolved in the condensate before sending to the steam production in the boiler. Those gas are oxygen, carbon dioxide and hydrogen chloride. Because the oxygen can be formed the oxide compound with the metal surface, and then the pipelines and the equipment can be deteriorated.

2.1.1.4 Boiler is the equipment that heat the condensate to form the steam by using natural gas. That is the energy source for several units in the refinery plant.

2.2 Advance Control Systems [Process Control Systems Application Design and Tuning Third Edition, F.G. Shinskey, 1988]

Although many companies around the world are implementing the advanced control techniques, the technology of each company is still misunderstood. For many industry, the advanced control means a complete changing of field instruments, wall-to-wall computing, pages of complicated equations and algorithms, and certainly substantial cost. As a result there are still managers in the industry who will find it easier to approve major redesign projects than investing in the process control improvements even though the control changes would cost far less and lead to greater improvements in productivity. The advance control systems used in the present day are three control systems as follows:

2.2.1 Cascade Control

Cascade control is technique which contains two closed loop controls and the loop controls are cascaded to each other. The output of the first loop controller is a set point for the second loop controller. The advantages and disadvatages as shown in Table 2.1

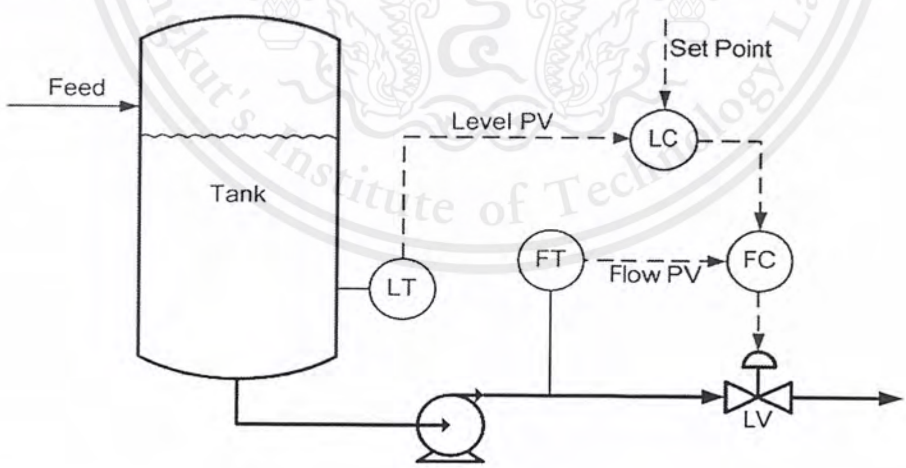


Figure 2.2 Cascade Control System

Reference: Process Control Systems Application Design and Tuning, F.G. Shinskey, 1988

Table 2.1 Advantages and Disadvantages of Cascade Control

Advantages	Disadvantages
1. Cascade control reduces the dead time and phase lag time in the control system.	1. Cascade control makes the system more complex.
2. Cascade control can be combined with feed forward and other type of controllers.	2. Cascade control requires more instruments and equipment.
3. Cascade control improved dynamic response and performance.	3. Tuning of cascade controllers is more difficult than close loop control.

2.2.1 Split Range Control

Split range control is technique which the output of the controller is split and sent to 2 or more control valves. The splitter defines how each valve is sequenced as the controller output changes from 0 – 100%. The split range control has one measured and more than one manipulated variables. The advantages and disadvatages are shown in Table 2.2

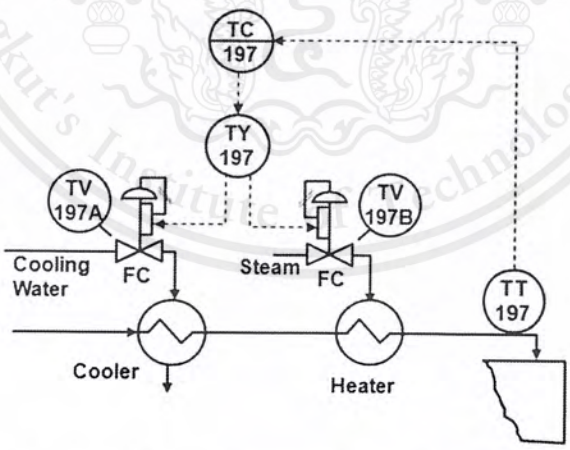


Figure 2.3 Split Range Control System

Refference: Process Control Systems Application Design and Tuning, F.G. Shinskey, 1988

Table 2.2 Advantages and Disadvantages of Split Range Control

Advantages	Disadvantages
1. Split range control can be controlled the process effectively.	1. Split range control requires physical quantity sources more than one
2. Split range control uses single controller for many final control element, so the control system is cheap and loading is educed.	2. If there will be some fault in the controller, all the system will be unbalance and device will be damaged.
3. Split range control uses many sources of one physical quantity to control as single output.	

2.2.3 Override Control

Override control is used to regulate input of process to maintain the process output at the target without violating a constraint on another output. Two controllers and a control selector (Low or High) are required for override control. The selector output is fed back to both controllers for the purpose of preventing controller reset windup when the controller is not selected.The advantages and disadvatages are shown in Table 2.3

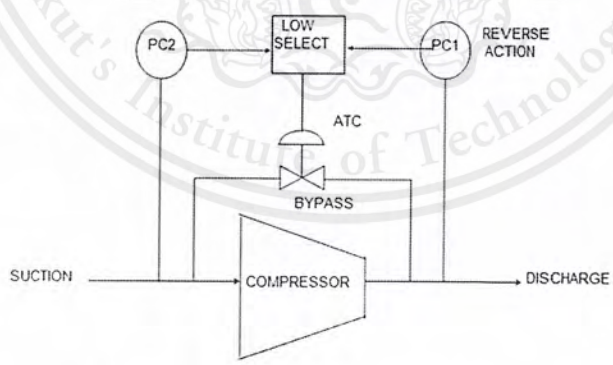


Figure 2.4 Override Control System

Reference: Process Control Systems Application Design and Tuning, F.G. Shinskey, 1988

Table 2.3 Advantages and Disadvantages of Override Control

Advantages	Disadvantages
1. Different switchover criteria can be used for an override control scheme. It depends on the given application which switch over logic is favorable.	1. This scheme should be applied only if the secondary regulator is rarely needed.
2. Using the limiting control with selection via controlled variables, a secondary controlled variable can be limited in both directions, preventing violations of upper and lower limits.	2. It is used only when you required certain level of protection for device.
3. Using an override control scheme with max/min selection of the manipulated variables, the secondary controlled variable can be limited only in one direction.	3. Not using it for continuous measurement.

2.3 Flow Coefficient

Control valves are specified by considering both properties of the process fluid and the desired flow characteristics in order to choose the valve body material and type. Then the desired characteristics for the actuator are considered. The choice of construction material depends on the corrosive properties of the process fluid at operating conditions. Commercial valves made of brass, carbon steel, and stainless steel can be ordered off the shelf, at least in smaller sizes. For large valves and more exotic materials of construction, special orders usually are required.

The sizing control valve equation as equation 2.1

$$\dot{m} = \frac{nC_v\rho}{1.17}\sqrt{\frac{\Delta P}{SG}} \qquad (2\text{-}1)$$

Where m = mass flow rate of fluid (kg/h)

n = percentage of valve opening

C_v = flow coefficient

ρ = density of fluid which is passing through CV (kg/m^3)

ΔP = pressure drop between inlet and outlet of CV (kg/cm^2)

SG = ratio of density of fluid to the density of water

PART II Study of the Vacuum System of a Hydrodesulfurization Unit

2.1 Crude Oil Refining [Bangchak Petroleum Public Co.,LTD, 2017]

Bangchak Petroleum Public Company is a complex refinery which have 4 main steps for refining the crude oil as follows

- Fractionation or Distillation Units
- Treating Units
- Conversion Units
- Blending

In each step consists of different units as shown in Figure 2.4

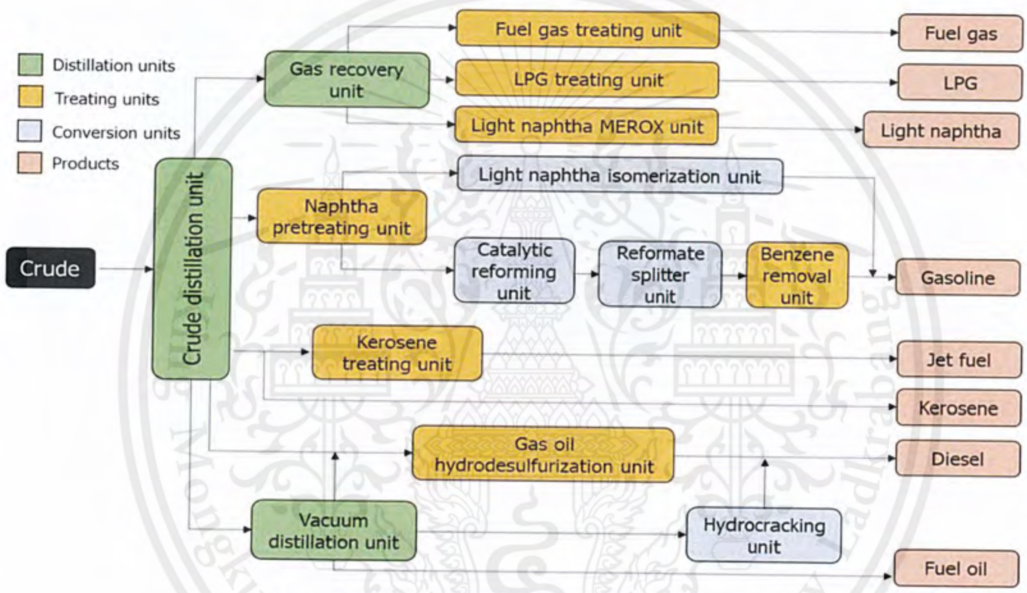


Figure 2.5 Complex Refinery Diagram

2.1.1 Fractionation or distillation unit is the separated method for the crude oil by the different boiling points of each composition in the crude oil.

2.1.1.1 Topping unit or crude distillation unit is the unit that fractionates crude oil by the different boiling points of each composition in the crude oil. Firstly, the crude oil from the storage tank will be sent to heating up section for example furnace, heat exchanger and boiler to increase the temperature before sending to factionator or distillation column. Secondaly, the factionator will separate the crude oil by different boiling points of each composition in the crude oil. The oil which has the lowest boiling point will leave out from the top of column and the others which have higher boiling point will leave out from the lower hight of column until the bottom of column. The products from topping unit are fuel gas, whole naphtha, kerosene, diesel, gas oil, and atmospherinc residue. These products will be sent to the treating unit and conversion.

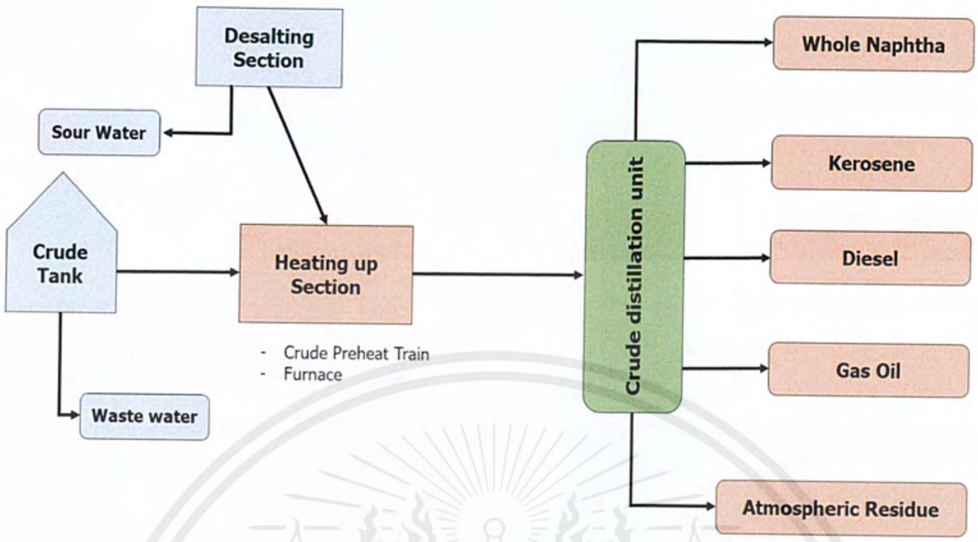


Figure 2.6 Topping Unit Diagram

2.1.1.2 Gas recovery unit is the unit that accumulates gas from the distillation unit especially light naphtha, after that light naphtha will be separated by deethanizer and debutanizer. The products from gas recovery unit are light naphtha, fuel oil and LPG (Liquified petroleum gas). These products will be sent to the treating unit and conversion.

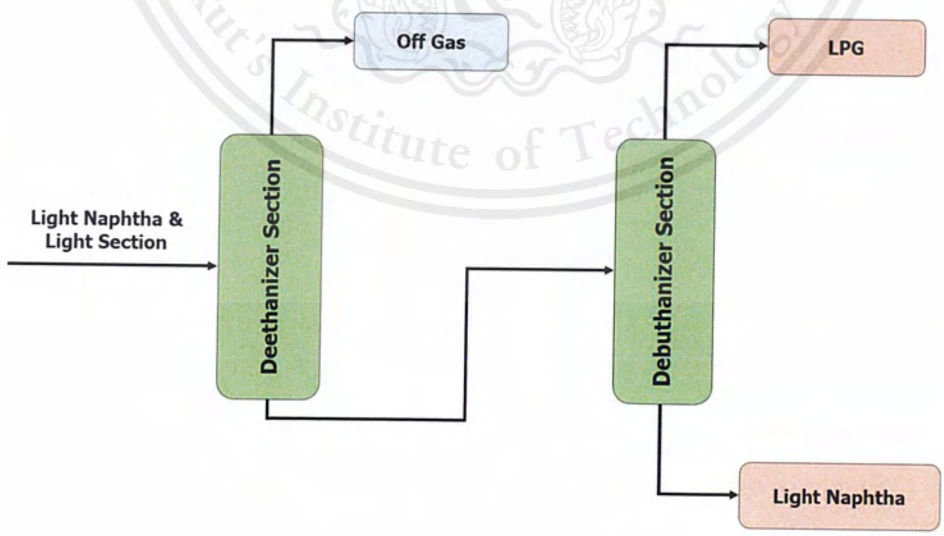


Figure 2.7 Gas Recovery Unit Diagram

2.1.1.3 Vacuum distillation unit is the unit that factionates the residue crude or atmospheric residue (AR) from the topping unit. The residue crude will be sent to the furnace for preheating before sending to the vacuum distillation column by different boiling points. The products from the vacuum distillation unit are light vacuum gas oil (LVGO) which will be sent to the gas oil hydrodesulfurization unit, vacuum gas oil which will be sent to the hydrocracking unit and vacuum residue (VR).

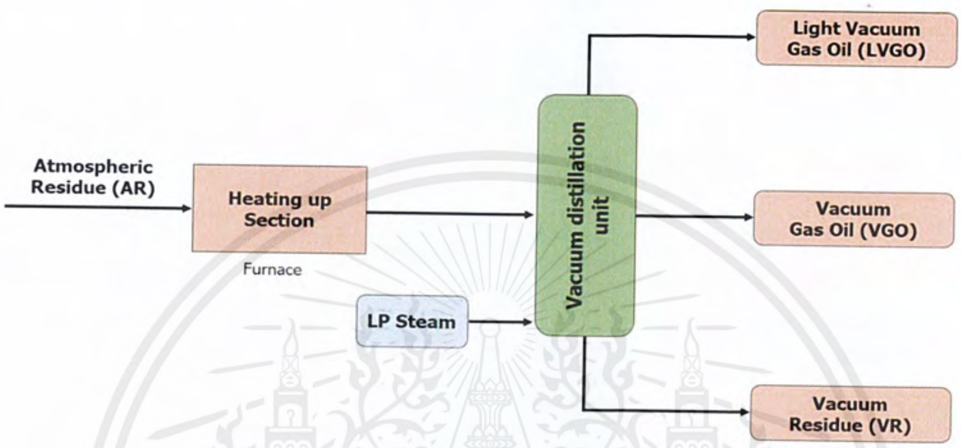


Figure 2.8 Vacuum Distillation Unit Diagram

2.1.1.4 Reformate splitter unit is the unit that distillates reformate from the catalytic reforming unit. The products are carbon five atoms which will be sent to the hydrogen plant and used to gasoline blending, Unsaturated light reformate which will be sent to the benzene removal unit and heavy reformate which will be sent to the gasoline blending.

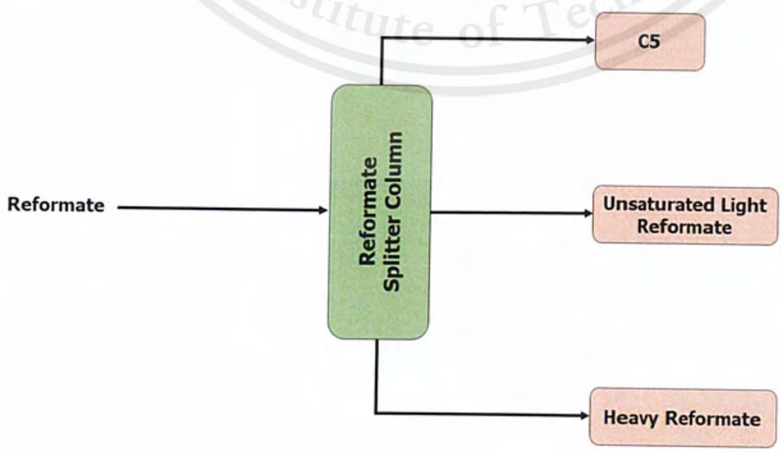


Figure 2.9 Reformate Splitter Unit Diagram

2.1.2 Treating unit is the unit that product quality improvement by removing contaminate from the crude oil such as sulfur, water, oxygen, nitrogen and mercury.

2.1.2.1 Fuel gas treating unit is the unit that removes the sulfur from the fuel gas by using amine treating. The amine solution can trap the sulfur in the fuel gas as the absorbent in the absorbtion section. The low sulfur fuel gas so called "Sweet fuel gas" will be used in distillation processes and the sulfur received will be sent to the sulfur recovery unit.

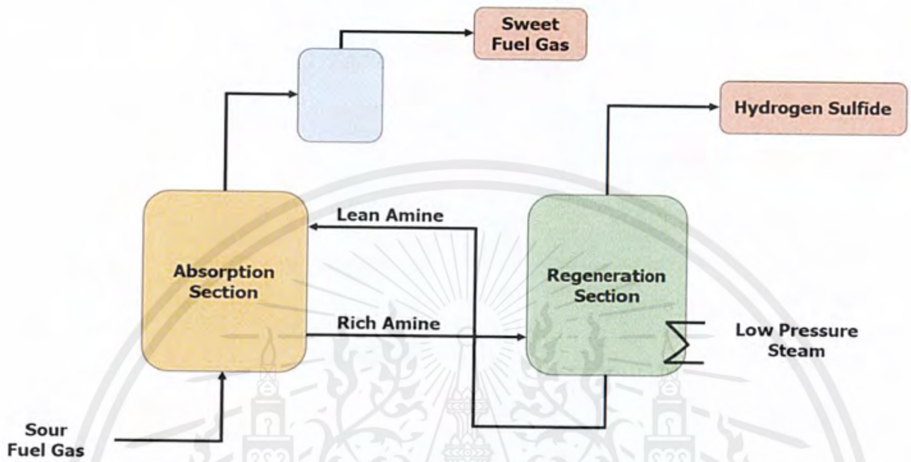


Figure 2.10 Fuel Gas Treating Unit Diagram

2.1.2.2 LPG treating unit is the unit that converts the mercaptan in the LPG to disulfide and separates these sulfur by using catalyst in regeneration section. LPG from gas recovery unit will be sent to the sulfur removal unit for forming sulfur to hydrogen sulfide, after that the hydrogen sulfide will be sent to the LPG treating unit to convert mercaptan to disulfide.The low sulfur LPG will be dealed along specification of customers.

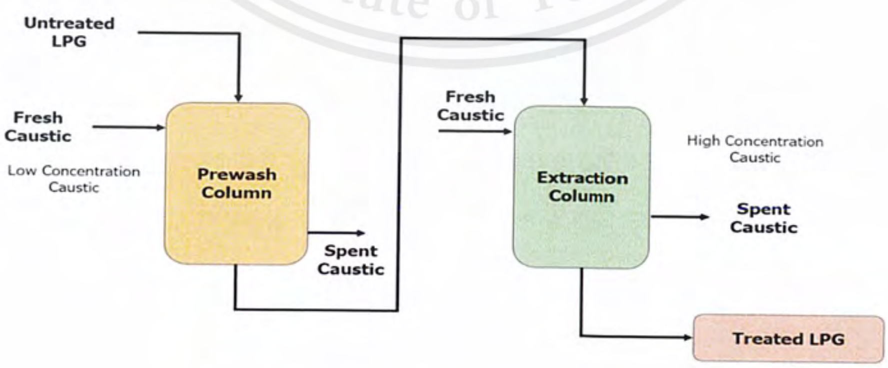


Figure 2.11 LPG Treating Unit Diagram

2.1.2.3 Naphtha pretreating unit is the unit that removes the sulfur from naphtha by using thermal energy and catalyst. Naphtha from the topping unit will be sent to the preheating section for increasing temperature. After that naphtha will be sent to the reactor for converting the sulfur to hydrogen sulfide gas. The off gas will be separated by stripper including hydrogen sulfide gas and the others will be sent to splitter to factionate light naphtha and heavy naphtha.

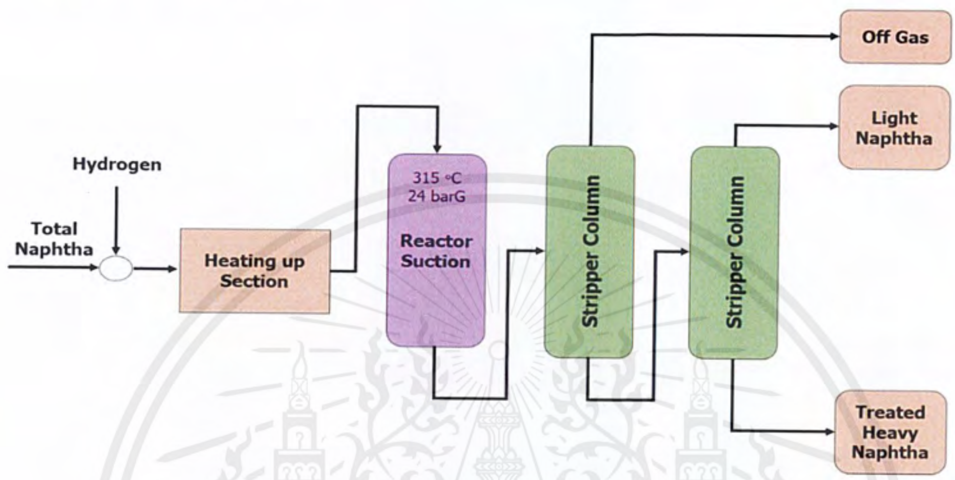


Figure 2.12 Naphtha Pretreating Unit Diagram

2.1.2.4 Kerosene treating unit is the unit that converts the mercaptan in the kerosene to disulfide by using catalyst and air including the quality improvement to jet fuel. Gas oil from the topping unit will be sent to equipment improved the suitable condition for the reactor, after that it will be sent to the reactor and equipment removed contaminate.

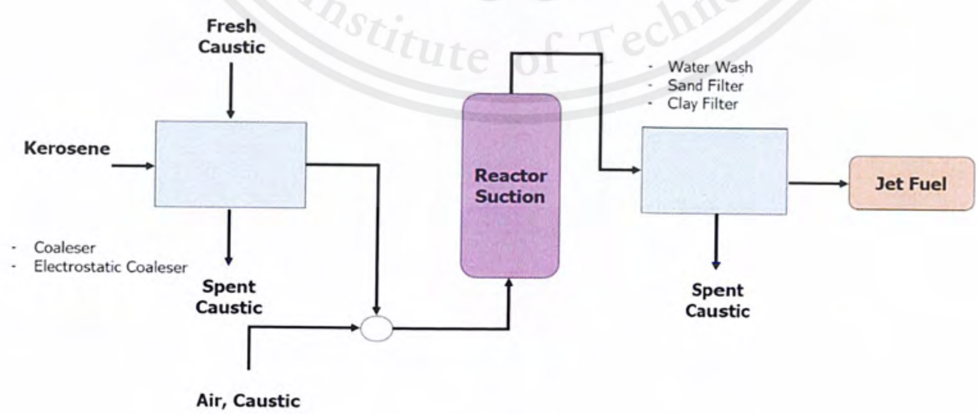


Figure 2.13 Kerosene Treating Unit Diagram

2.1.2.5 Gas oil hydrodesulfurization unit is the unit that removes the sulfur from high speed diesel by using hydrogen, thermal energy and catalyst. Diesel oil and gas oil from the topping unit will be sent to the preheating section and the reactor to convert the sulfur to hydrogen sulfide gas. The received off gas will be togethered with fuel gas and they will be sent to the sulfur removal unit.The low sulfur high speed diesel will be dealed along specification of customers.

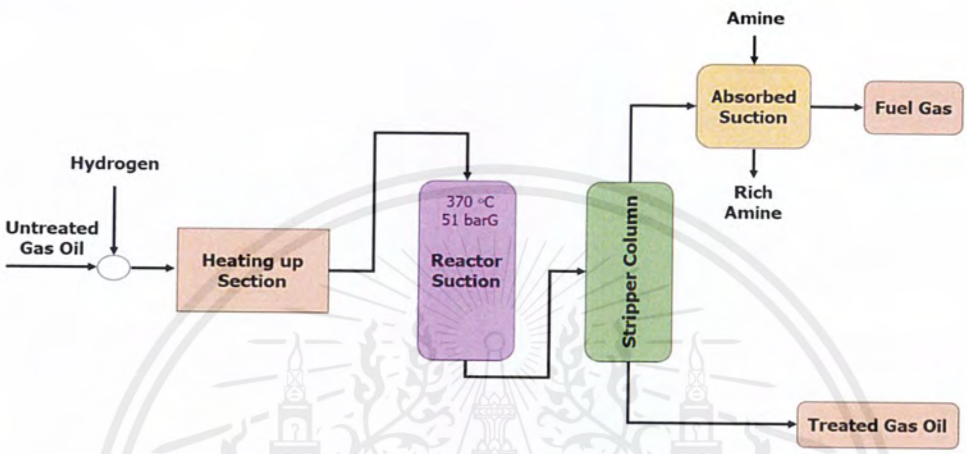


Figure 2.14 Gas Oil Hydrodesulfurization Unit Diagram

2.1.2.6 Benzene removal unit is the unit that removes benzene from unsaturated light reformate by using hydrogen gas, thermal energy and catalyst. Light reformate oil will be sent to the preheating unit and the reactor to convert benzene to saturated cyclohexane. After that the saturated cyclohexance will be blended to gasoline.

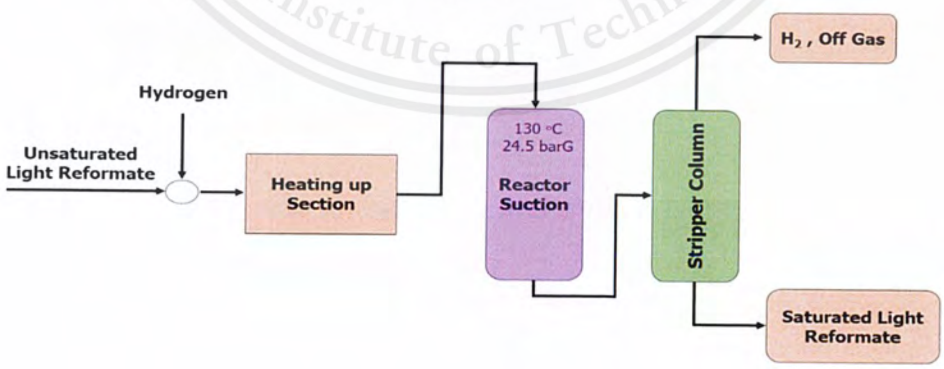


Figure 2.15 Benzene Removal Unit Diagram

2.1.2.7 Sour water stripper unit is the unit that eliminates hydrogen sulfide gas and ammonia gas from sour water which receives from distillation unit and treating unit. The off gas will be sent to burning with fuel gas in distillation unit or sulfur recovery unit.The treated water which have low sulfur content will be sent to wastewater treatment unit or recycle in the system.

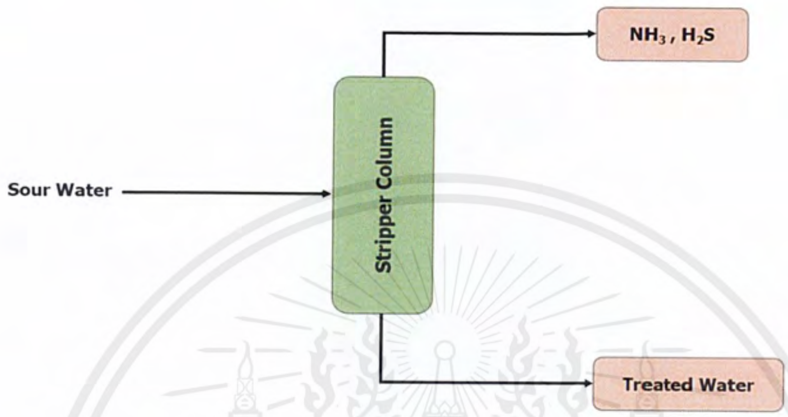


Figure 2.16 Sour Water Stripper Unit Diagram

2.1.2.8 Spent caustic treating unit is the unit that removes the hydrogen sulfide gas from spent caustic receiving LPG treating unit and kerosene treating unit.The spent caustic will be reconditioned with hydrochloric acid and sent to stripper column to remove the hydrogen sulfide gas. The off gas will be sent to burning with fuel gas in distillation unit and the treated spent caustic will be adjusted pH for suitable condition in the system.

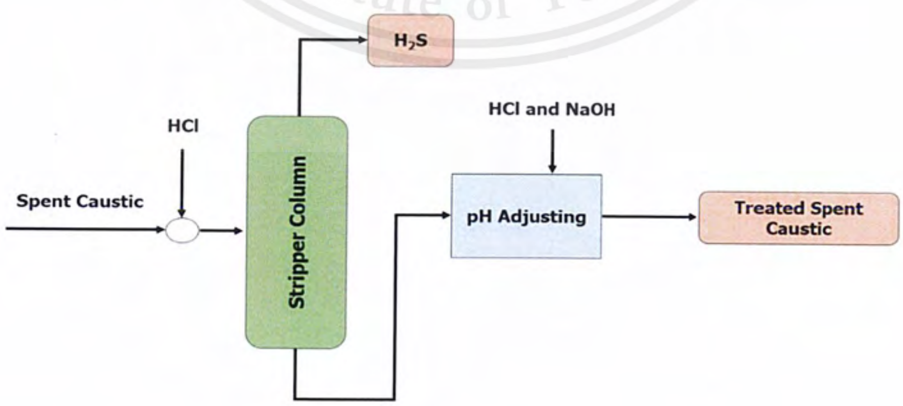


Figure 2.17 Spent Caustic Treating Unit Diagram

2.1.3 Conversion is the unit that obtains high quality products and value by changing molecular structure of oil.

2.1.3.1 Light naphtha isomerization unit is the unit that changes the molecular structure of light naphtha oil from normal to isomerate which octane number will be increased. The reaction is occurred by using hydrogen gas, thermal energy and catalyst so called "Isomerization reaction"

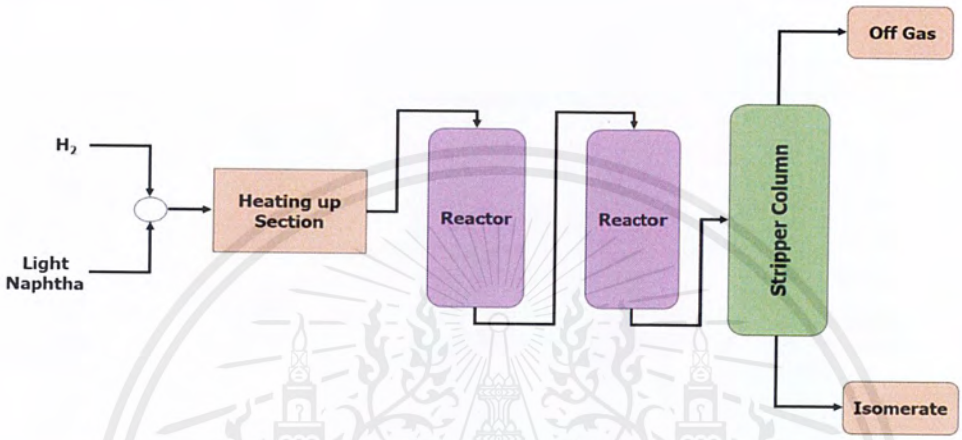


Figure 2.18 Light Naphtha Isomerization Unit Diagram

2.1.3.2 Catalytic reforming unit is the unit that changes the molecular structure of heavy naphtha from normal to ring which octane number will be increased. The reaction is occurred by using thermal energy and catalyst so called "Reforming reaction".

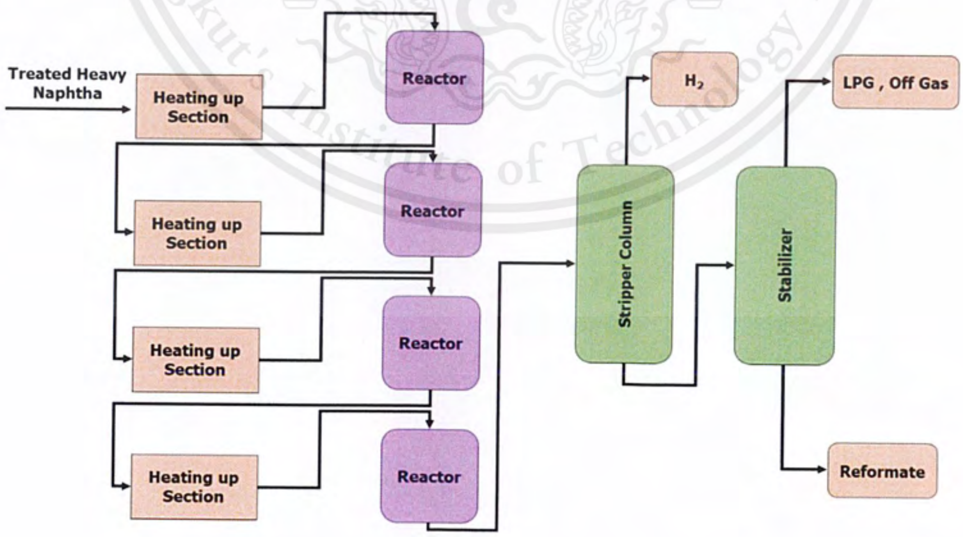


Figure 2.19 Catalytic Reforming Unit Diagram

2.1.3.3 Hydrocracking unit is the unit that crecks the heavy naphtha or vacuum gas oil (VGO) from the vacuum distillation unit to light naphtha by using hydrogen gas and thermal energy. The products are fuel gas, LPG, light naphtha, heavy naphtha, jet fuel, diesel oil and fuel oil.

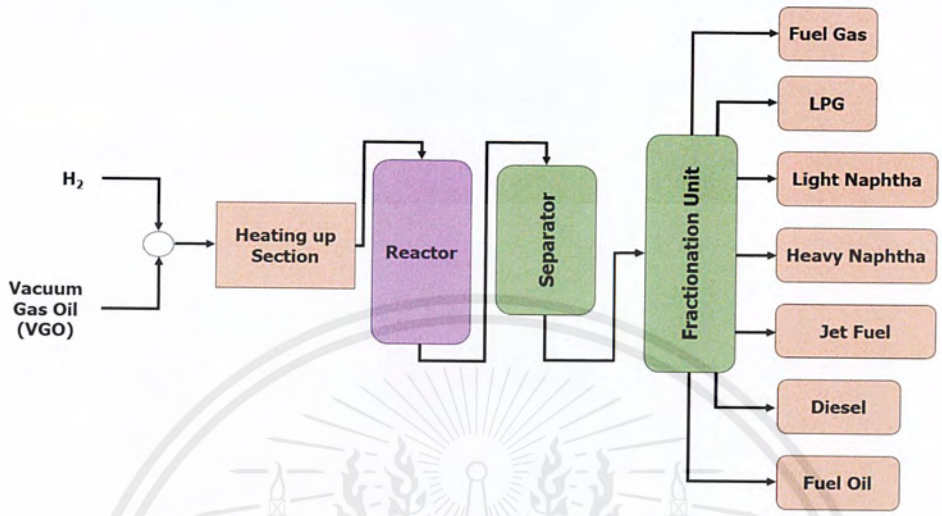


Figure 2.20 Hydrocracking Unit Diagram

2.1.4 Supporter unit or Pollution controller unit is the special unit that is responsible for supporting the production processes and controlling the pollution in the system.

2.1.4.1 Hydrogen plant is the unit that produces hydrogen gas for many units such as hydrocracking unit and isomerization unit. The reactant may be LPG, fuel gas or light naphtha which have low sulfur, chloride and metal contents.

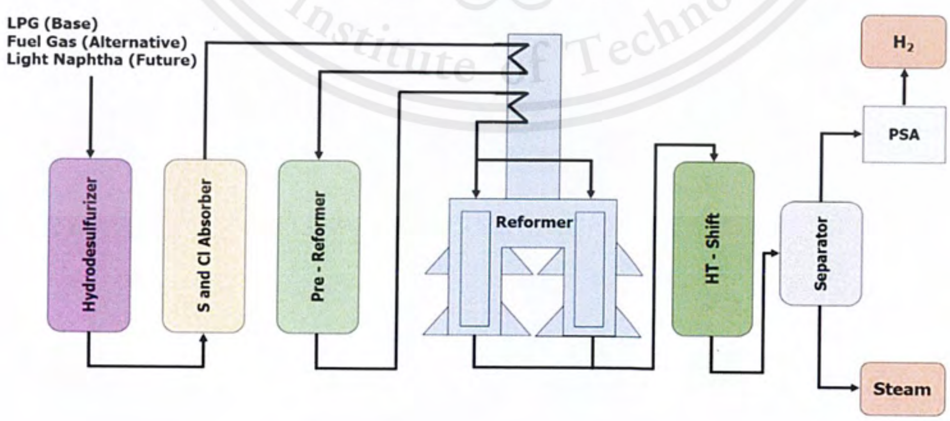


Figure 2.21 Hydrogen Plant Diagram

2.1.4.2 Sulfur recovery unit is the unit that converts the separated sulfur from many types of oil to liquid sulfur by using air, thermal energy and catalyst.

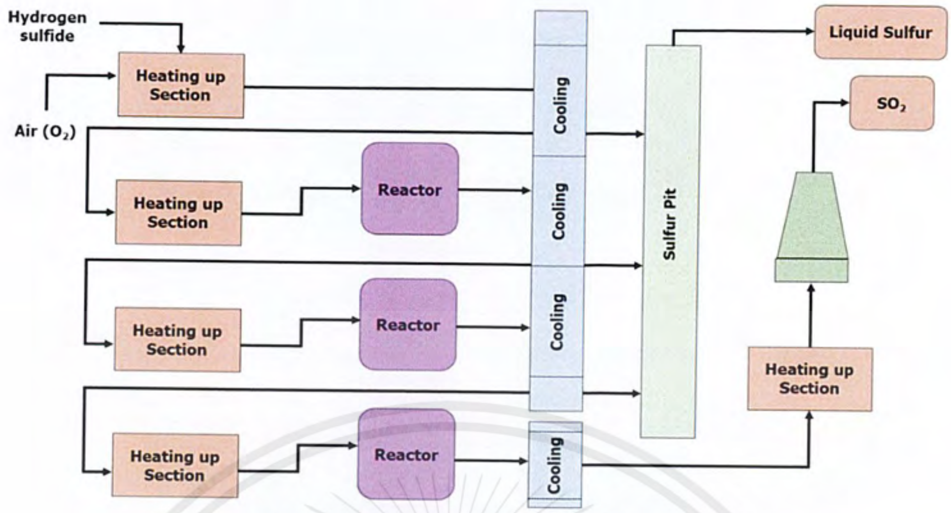


Figure 2.22 Sulfur Recovery Unit Diagram

2.1.4.3 Wastewater treatment unit is the unit that treats wastewater from several units in the refinery plant. Non-contaminated wastewater such as wastewater from cooling tower, power plant, boiler blowdown and demineralizer will be treated to neutral and drained to the pond. On the other hand, contaminated wastewater from production processes must be treated by physical treatment process that separates the oil from the water surface by using a wier, API separator in CPI or TPI plate form and Dissolved air floatation system.

2.1.5 Blending is the combination method of oil since two or more products that may be added to additive for upgrating some quality and covering to standard of Ministry of Energy and customers.

2.2 Gas Oil Hydrodesulfurization Unit [Bangchak Petroleum Public Co.,LTD, 2017]

2.2.1 Process Principle of Sulfur Removal in Hydrodesulfurization

Sulfur compounds in the gas oil include mercaptans, sulfides, disulfides, cyclic sulfides and thiophenes. The reaction is to change the sulfur compound to hydrogen sulfide gas which is exothermic reaction. The typical hydrodesulfurization reactions are shown as follows:

- Mercaptans

$$CH_3 - CH_2\text{-}CH_2\text{-}SH + H_2 \rightarrow CH_3\text{-}CH_2\text{-}CH_3 + H_2S \qquad (2\text{-}2)$$

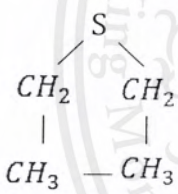
- Sulfides

$$CH_3 - CH_2\text{-}S\text{-}CH_2 - CH_3 + 2H_2 \rightarrow 2CH_3 - CH_3 + H_2S \qquad (2\text{-}3)$$

- Disilfides

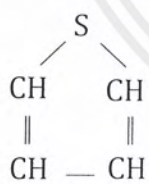
$$CH_3 - CH_2\text{-}S\text{-}S\text{-}CH_2 - CH_3 + 3H_2 \rightarrow 2CH_3 - CH_3 + 2H_2S \qquad (2\text{-}4)$$

- Cyclic Sulfides



$$+ 2H_2 \rightarrow CH_3 - CH_2 - CH_2 - CH_3 + H_2S \qquad (2\text{-}5)$$

- Thiophene



$$+ 4H_2 \rightarrow CH_3 - CH_2 - CH_2 - CH_3 + H_2S \qquad (2\text{-}6)$$

2.2.2 Process Description of Gas Oil Hydrodesulfurization in Plant 3

The process of the HDS unit under normal operation consists of 7 main parts (see Figure 2.23) as follows

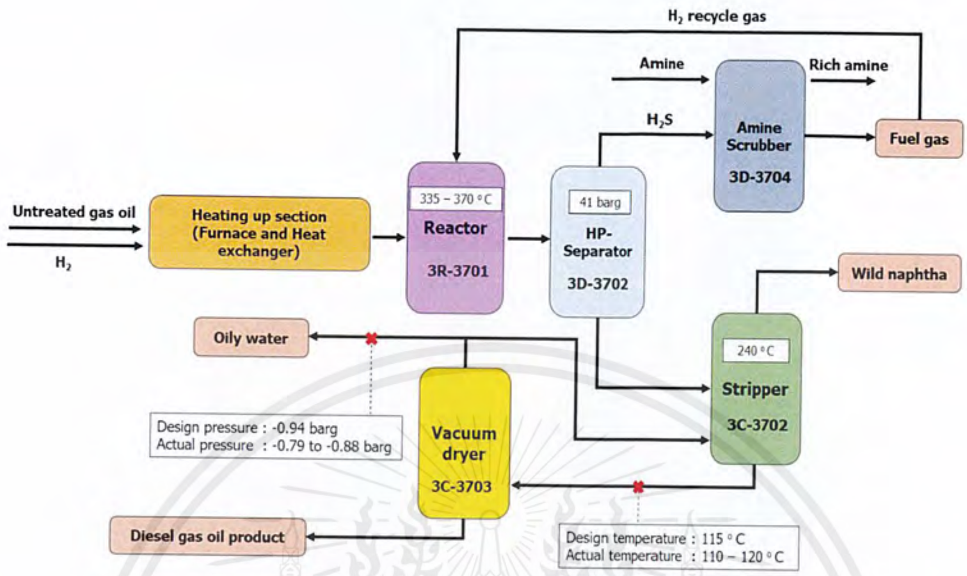


Figure 2.23 Gas Oil Hydrodesulfurization Diagram in Plant 3

2.2.2.1 Heating up Section

In plant 3 the heating section are furnance and heat exchanger for heating the feed stream to 335 – 370 $^{\circ}$ C that is suitable condition for gas oil hydrodesulfurization reaction. This heater is equipped with decoking facilities and is used for situ regeneration of the reactor catalyst.

2.2.2.2 Reactor Section

After passing through the heating up section the stream is brought up to the required temperature. The preheated stream is fed to the top of the reactor and passed across two catalyst beds which the desulfurization reaction takes place. The catalyst consists of a very high purity alumina, loaded with cobalt and molybdenum oxides. A recycle gas quench is needed between the two beds because the reaction is exothermic. The temperature increase across the reactor should be limited to 15 $^{\circ}$ C and the pressure drop across the reactor in normal operation is 5 bar maximum. The product of reaction will be cooled down from the feed steam by using heat exchanger and then sent to the High pressure separator section.

2.2.2.3 High Pressure Separator Section

The air cooler effluent from compressor passes to the high pressure separator drum where the liquid and vapor phases are separated. The vapor phase is hydrogen sulfide gas and leave off to the Amine scrubber section. The liquid consists of hydrocarbon and water phases. The liquid hydrocarbon from the high pressure separator is directed to the stripper. Separation of the HC/H_2O phases is poor in this drum because the aim of this part is hydrogen sulfide removal and the decantation is aided equipped by the installation of a coalescer.

2.2.2.4 Amine Scrubber Section

The amine scrubber is a column equipped with 18 one pass valve trays, The H_2S content is reduced to acceptable levels (< 1.1 wt %) to prevent catalyst poisoning, by contacting with an amine stream (Diethamolamine compound, DEA). The lean amine stream which originates from the Fuel Gas Treating Unit and it's sent back after passage in the scrubber for DEA regeneration.

2.2.2.5 Stripper Section

The liquid hydrocarbon from the high pressure separator is dropped down in pressure across the control valve and preheated to the required stripper feed temperature 240 $^{\circ}$C. The feed enters the stripper on the tray No.6 (numbered from top to bottom). The stripper consists of 24 valve trays, the top 5 being one pass trays, the lower 19 being two passing trays.

The purpose of the column is to reduce the gas oil H_2S content and to stabilize the gas oil by removal of the light end components. Live steam, either from the steam generator or from the HP steam header, are introduced below the bottom tray and are contacted with the gas oil descending the column. The steam provides a partial pressure which helps the stripping of light components from the gas oil.

The air cooler outlet stream is of mixed phase and is collected in the reflux drum. The operating conditions being 6.3 barg and 52 $^{\circ}$C. The reflux drum separates the vapor phase (which is routed to the FGTU) from the liquid phases. The decanted water is sent to the Sour water treating unit (SWTU) and the hydrocarbon phase serves in part as a reflux to the stripper and the rest is sent to the topping unit in plant3 downstream of the valve.

The vapor leaving the top of the column consist principally of water vapor and the light end components. These gases are partially condensed in the air cooler.On the other hand, the liquid hydrocarbon phase is sent to the vacuum dryer section.

2.2.2.6 Vacuum Dryer Section

The purpose of the vacuum drier is to remove free and dissolved water in order to meet the required product specification for the gasoil about less than 120 mg/L. The vacuum drier is equipped with 8 trays. The four top trays are discs and donuts (4 discs, 4 donuts) and the four lower are perforated trays. These trays increase the efficiency of separation between the vaporized phase and the liquid phase. The feed enters the column, which is at a low pressure of -0.88 barg, between the second and third pair of disc and donuts. The vacuum is assured by the ejector package.

The resulting flash in the drier column is enhanced by the installation of the disc and donuts. The vapor phase rises in the column and is washed with a recycle gasoil stream so as to minimize the gasoil losses in the overhead gas stream. The liquid phase falls in the column, the bottom of the column having a larger diameter compatible with a residence time of 8 minutes so as to facilitate the downstream blending operations.

2.2.2.7 Vacuum Package

The vacuum package in the vacuum dryer unit consists of 3 main components are steam ejectors (3J-3717), heat exchanger (3E-3717) and vacuum pump (3P-3708) which each component is colllaborated to generate vacuum condition as shown in Figure 2.24

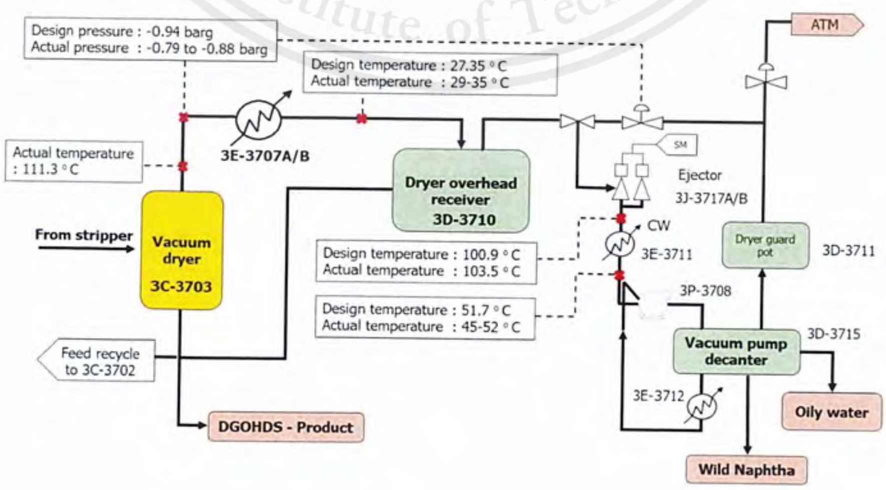


Figure 2.24 Vacuum Package in the Vacuum Dryer

The steam ejectors are the vacuum generator by using the principle of fluid mechanic that is changing the velocity of fluid immediately from reducing surface area in the steam ejectors. It effects turbulent flow at the throat and the pressure will be decreased from different velocity. In normal operation the system consists of 2 ejectors, one of which is in operation and one in standby.The steam which is used in steam ejector is being 10 barg and 210 °C

The type of heat exchanger (3E-3711) is shell and tube that is one shell and four pass. The function of this heat exchanger is cool down the mixing steam from Dryer overhead receiver and Steam ejector until the stream is all liquid phase or 41.4 °C for protecting back pressure problem in vacuum pump and then the vacuum pressure in the vacuum dryer will be decreased as well. The capacity of heat exchanger is 51.86 kW.

The type of vacuum pump is liquid ring that the feed stream must be all liquid phase. If the inlet stream is vapor and liquid phases, the part of vapor will block the outlet stream and accumulate in the vacuum pump until the outlet pressure is higher than the inlet pressure. The back pressure will be occurred from the vacuum pump to steam ejector.Therefore, the vacuum pressure in the vacuum dryer will be increased, and then the water content in the product will be increased.

2.3 Heat Exchanger [Heat and Mass Transfer Fundamentals & Applications Yunus A. Cengel and Afshin J. Ghajar, 2015]

The heat exchanger is the equipment that used to transfer heat between fluids by the fluid does not mix. The heat exchangers are the important tools for several production processes because its can increase or descrease temperature to the suitable conditions. The design of heat exchanger must have basic knowleage in the convention, conduction and fluid mechanic.

Nowadays the most of industrial processes have related the heat exchangers such as the oil refinery. It is used for increasing the temperature of crude oil to fractionate the components of crude oil by using different boiling points of each component in the distillation unit. In the same way the fertilizer production industry, polyester industry and the other industries use the heat exchanger to increase, decrease temperature or recycle heat from the fluid.

The main function of the heat exchanger is using thermal energy to the highest efficiency, so the optional heat exchanger and the application are related to the cost of the process and the price of the product. The key criteria for choosing a type of heat exchanger depends on efficiency of heat exchanger at that condition and cost.

There are many kinds of heat exchangers and classified by the direction of the fluid movement and characteristic of the heat exchanger. The basic types of heat exchangers are shown as follows

2.3.1 Concentric Tube or Double Pipe

The simplest type of heat exchanger consists of two concentric pipes of different diameters called the double-pipe heat exchanger. One fluid in a double-pipe heat exchanger flows through the smaller pipe while the other fluid flows through the annular space between the two pipes. The flow arrangements are possible two types that are parallel flow, counter flow and cross flow as shown in Figure 2.25 and Figure 2.26

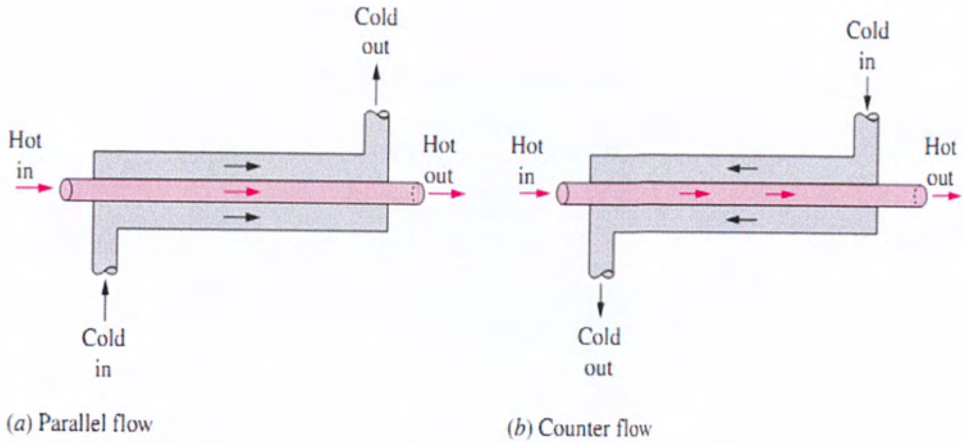


Figure 2.25 Different Flow in a Double-pipe Heat Exchanger.

Reference: Heat and Mass Transfer Fundamentals & Applications, Yunus A. Cengel and Afshin J. Ghajar, 2015

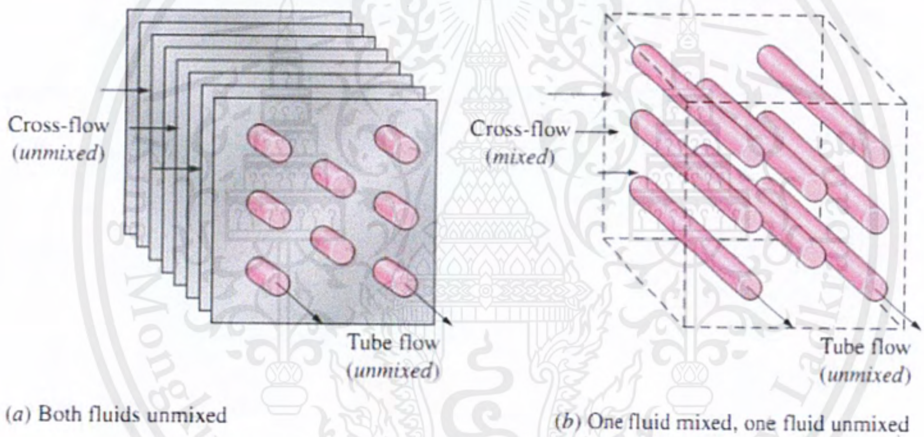


Figure 2.26 Different Flow Configurations in Cross-Flow Heat Exchangers.

Reference: Heat and Mass Transfer Fundamentals & Applications, Yunus A. Cengel and Afshin J. Ghajar, 2015

2.3.2 Shell and Tube Heat Exchanger

The most common type of heat exchanger in industrial applications is the shell and tube heat exchanger. One fluid is in the shell and another fluid flows through tubes that heat transfer occurs. Shell and Tube heat exchangers contain several number of tubes packed in a shell with their axes parallel to that of the shell.

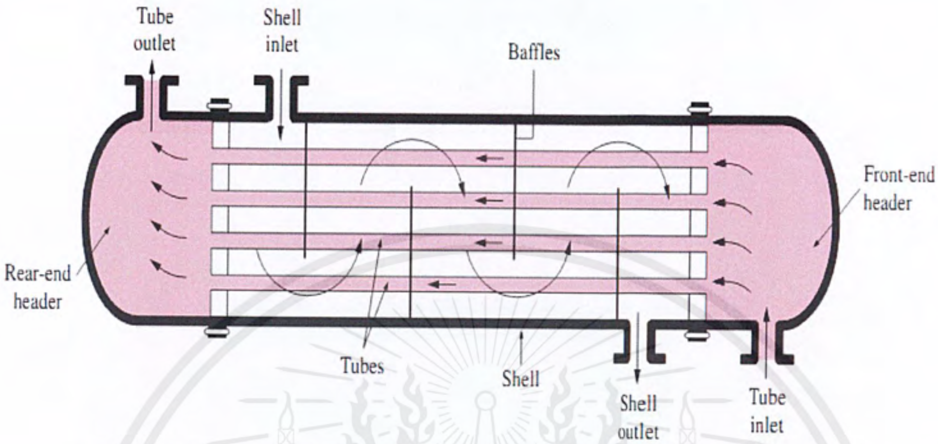


Figure 2.27 The Schematic of a Shell and Tube Heat Exchanger (One-shell Pass and One-tube Pass).

Determining the type of fluid in the tube or shell based on the properties of the fluid is shown in Table 2.4

Table 2.4 Determining the Type of Fluid in Shell and Tube Heat Exchanger [Coulson, 1999]

Fluid in the Tube	Fluid in the Shell
Fouling and Corrosion	Pressure Drop Limitng
High Pressure	High Flow Rate
High Temperature	May be Condensation and Boiling
Low Viscosity	Do not have Suspensions

2.3.3 Air Cooled Heat Exchanger

This type of heat exchanger uses air as a coolant instead of cold water. The structure of the exchanger is composed of 3 main components that are fin, steel structure for pipe tie and air blower. The principle of air cooled heat exchanger is air blowing through induced draft or fin which heat is transferred from the fluid in pipe to the air. The special features of this type are 3 parts as follows. First, it's using air that is no limit in the atmosphere instead of water. Second, it can eliminate rust or dirt problems from using water in the cooling. Finally, the maintenance costs are cheaper than the other types. The disadvantage of this type are loud noise and high cost for installation.

2.3.4 Plate Type Heat Exchanger

The special features of this type is that removes the multiple heat transfer plates to arrange in the same distance and then the fulid flows through the gap between the plates. This type of heat exchanger is often used to increase the temperature of liquid that the structure is made of thin stainless steel or titanium sheet because it can resist to the rust. The advantages of this type are that the heat transfer coefficient of the plate type is high, the maintenance cost is low because it is easy to clean up and it can adjust the amount of heat transfer by controlling the number of plates. The disadvantage of this type is not suitable for high pressure and temperature condition because the gaskets are made of rubber or synthetic rubber.

2.4 The Overall Heat Transfer Coefficient [Heat and Mass Transfer Fundamentals & Applications Yunus A. Cengel and Afshin J. Ghajar, 2015]

The overall heat transfer coefficient is important to design the heat exchange because it is combination of conduction, convection and radiation in the system that is performance of each heat exchanger. The definition of overall heat transfer coefficient is described by the equation 2-7

$$\dot{Q} = \frac{\Delta T}{R} = UA\Delta T = U_i A_i \Delta T = U_o A_o \Delta T \quad (2\text{-}7)$$

$$R = R_{total} = R_i + R_{wall} + R_o \quad (2\text{-}8)$$

When

$$R_i = \frac{1}{h_i A_i} \quad (2\text{-}9)$$

$$R_o = \frac{1}{h_o A_o} \quad (2\text{-}10)$$

And

$$R_{wall} = \frac{\ln(r_2/r_1)}{2\pi k L} \quad (2\text{-}11)$$

Where A = heat transfer area (m^2)

A_i and A_o = heat transfer area inlet and outlet (m^2)

h = the convective heat transfer coefficient (W/m^2 °C)

k = thermal conductivity (W/m °C)

L = length of pipeline (m)

Q = heat transfer rate (W)

R = total thermal resistant of the system (°C /W)

R_i and R_o = thermal resistant of fluid inside and outside (°C /W)

r_i and r_o = inner radius and the outer radius of piper (m)

ΔT = temperature difference (°C)

U = the overall heat transfer coefficient (W/m^2 °C)

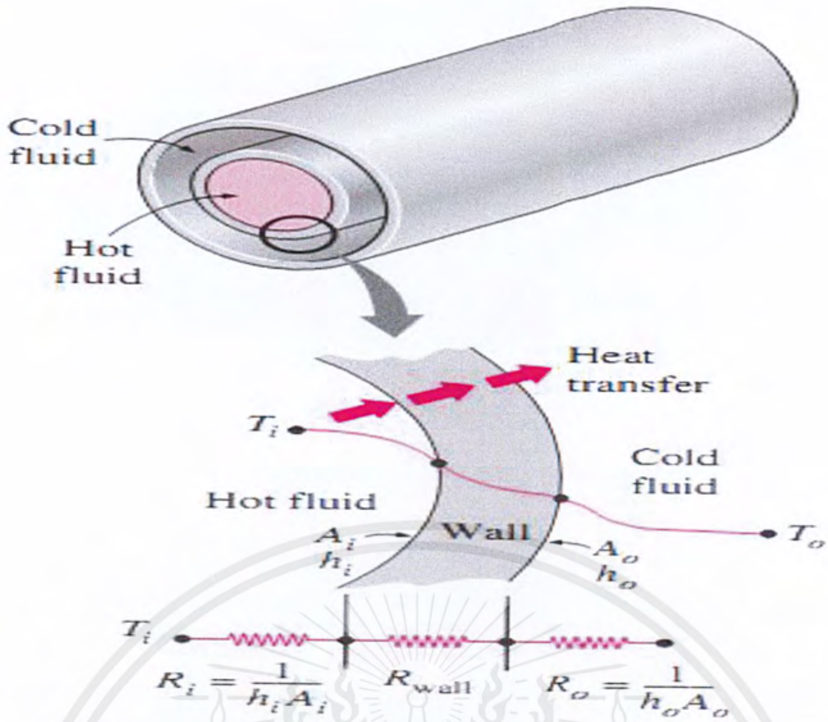


Figure 2.28 Thermal Resistance Network Associated with Heat Transfer in a Double-pipe Heat Exchanger.

2.5 Foiling Factor [Heat and Mass Transfer Fundamentals & Applications Yunus A. Cengel and Afshin J. Ghajar, 2015]

When the heat exchanger have been operated for a long time, the dirt from the fluid will attach to the inner surface of heat exchanger which is effect decreasing heat transfer rate at this surface and deteriorating of heat exchanger. If the design of heat exchanger does not consider with the resistant of dirt or foiling factor, the heat exchanger cannot operate to follow by design. Therefore, the foiling factor is the important thing in the design of heat exchanger.

The amount of dirt in the heat exchanger is foiling factor (R_f) that is the dirt resistant at the wall of heat exchanger (m^2 K/W). It can be found from the back part of dirt coefficient (h_d)

Table 2.5 Dirt Coefficient [Coulson, 1999]

	h_d (W/m^2 K)	h_d (btu/h ft^2 °F)
Distilled and Seawater	11,350	2,000
City Water	5,680	1,000
Muddy Water	1,990 - 2,840	350 - 500
Gases	2,840	500
Vaporizing Liquids	2,840	500
Vegetable	1,990	350

2.6 Analysis of Heat Exchangers [Heat and Mass Transfer Fundamentals & Applications Yunus A. Cengel and Afshin J. Ghajar, 2015]

The design and option of heat exchanger for each application must have the outlet tempertature of hot and cold fluid prediction ability and the total heat transfer rate calculation when knowing the mass flow rate of the fluids. For total heat transfer calculation can be found two methods which are Log Mean Temperature Difference (LMTD) and NTU Effectiveness.

Because the heat transfer is usually used for a long time, it assumes that the flow of fluid is constant. The fluid properties such as temperature, exist velocity, kinetic emergy, protentail energy and heat capacity are not change. The conduction heat transfer along the axis of pipe is not important and not considered. It assumes that the outlet surface is good insulated and it does not have loss of energy to the environment. Consequently, there are only heat transfer rate between fluids. From the conservation of energy, the heat transfer rate from the hot fluid is equal to the heat transfer rate from cold fluid that follows by equation 2-12

$$\dot{Q} = \dot{m}_c Cp_c\left(T_{c,out} - T_{c,in}\right) = \dot{m}_h Cp_h\left(T_{h,in} - T_{h,out}\right) \quad (2\text{-}12)$$

Where Cp_h = specific heats of hot fluid (kJ/kg °C)

Cp_c = specific heats of cold fluid (kJ/kg °C)

m_h = mass flow rate of hot fluid (kg/h)

m_c = mass flow rate of cold fluid (kg/h)

Q = the heat transfer rate (W)

$T_{h,in}$ = hot inlet temperature (°C)

$T_{c,in}$ = cold inlet temperature (°C)

$T_{h,out}$ = hot outlet temperature (°C)

$T_{c,out}$ = cold outlet temperature (°C)

The rate of heat transfer in a heat exchanger can also be expressed in an analogous manner to Newton's law of cooling as equation 2-13

$$\dot{Q} = UA_s\Delta T_m \quad (2\text{-}13)$$

Where A_s = heat transfer area (m^2)

ΔT_m = the average temperature difference (°C)

U = the overall heat transfer coefficient (W/m^2 °C)

2.6.1 Log Mean Temperature Difference (LMTD)

The temperature of fluid in the heat exchanger is not normally constant. Its change from one point to the next point when the heat is transferred from the hot fluid to the cold fluid. Although the heat resistant is constant, the heat transfer rate is changed by the flow movement in the heat exchanger because the heat transfer rate varies with the average difference temperature as shown in Figure 2.29

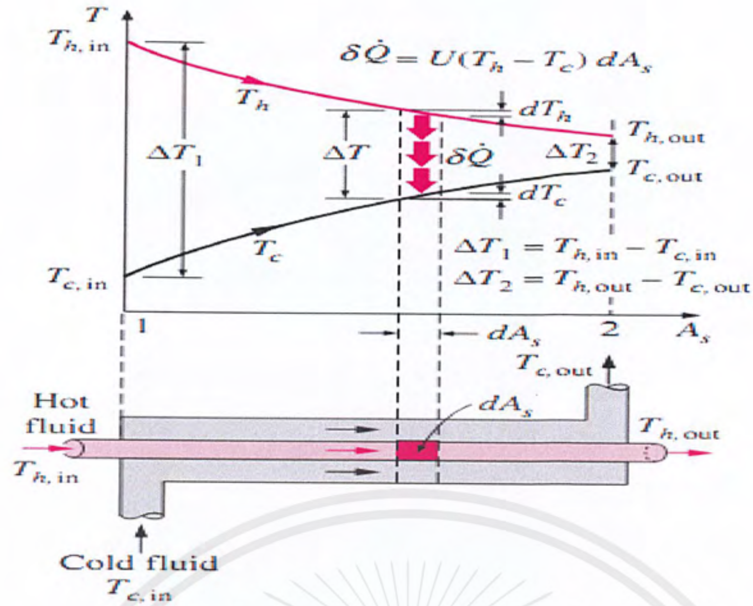


Figure 2.29 Variation of the Fluid Temperatures in a Parallel-flow Double-pipe

Figure 2.29 shows the variation of temperature in the parallel flow. The different temperature between hot and cold fluids at the entrance is high, but it will be reduced exponentially to the exit. The temperature of hot fluid is decreased while another one is increased. However, the temperature of cold fluid cann't exceed the temperature of hot fluid in the heat exchanger.

In part of heat transfer rate calculation, it can find from integration of heat transfer rate in each dA which is infinitesimal heat transfer area. From the equation 2-12 ΔT_m will be changed to ΔT_{lm} as equation 2-14

$$\Delta T_{lm} = \frac{\Delta T_1 - \Delta T_2}{\ln(\Delta T_1/\Delta T_2)} \tag{2-14}$$

Where ΔT_{lm} = log mean temperature difference (°C)

ΔT_m = average temperature difference (°C)

The log mean temperature different is used for heat transfer rate calculation roughly in each application because the overall heat transfer coefficient doesn't constant. However, in the design of heat exchanger often uses the overall heat transfer coefficient for surface area where is the center of heat exchanger. If the overall heat transfer is high deviation, it will be calculated the overall heat transfer in each steps and integrated to find the heat transfer rate.

For the complex heat exchanger such as multipass and cross flow heat exchanger will be calculated the average temperature different by using correction factor multiply with the log mean temperature different as equation 2-15.

$$\Delta T_{lm,CF} = F\Delta T_{lm} \quad (2\text{-}15)$$

Where F = correction factor

$\Delta T_{lm,CF}$ = log mean temperature difference (°C)

ΔT_{lm} = log mean temperature difference (°C)

The correction factor can be found by

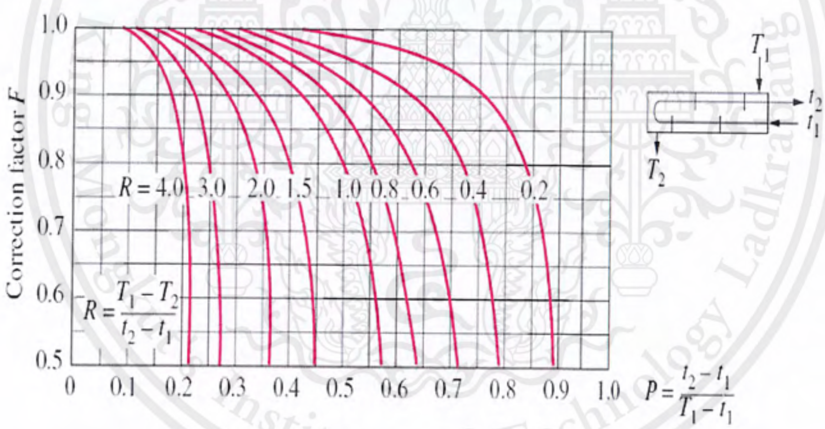


Figure 2.30 One-shell Pass and 2, 4, 6, etc. (any multiple of 2), Tube Passes

Reference: Heat and Mass Transfer Fundamentals & Applications, Yunus A. Cengel and Afshin J. Ghajar, 2015

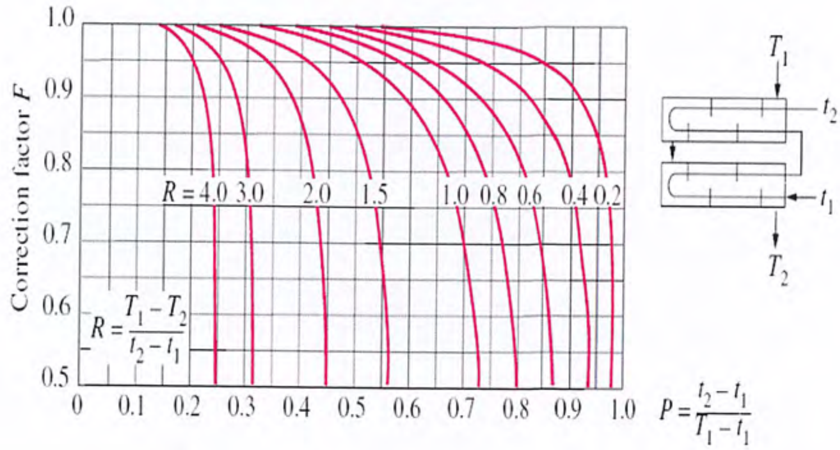


Figure 2.31 Two-shell Passes and 4, 8, 12, etc. (any multiple of 4), Tube Passes

Reference: Heat and Mass Transfer Fundamentals & Applications, Yunus A. Cengel and Afshin J. Ghajar, 2015

The value at the x-axis is the ratio of different temperature as equation 2-16

$$P = \frac{t_2 - t_1}{T_1 - t_1} \tag{2-16}$$

Where T_1 = inlet temperature of shell side (°C)

t_1 = inlet temperature of tube side (°C)

t_2 = outlet temperature of tube side (°C)

P value indicates to increase or decrease of fluid and is between 0 – 1. In case of the inlet temperature of hot fluid is equal to the outlet temperature of cold fluid, so the P value is equal to 1. For the R value is the ratio between both the mass flow rates multiply with specific heat capacity. The R ratio is equal to the temperature different of fluid in shell side devied by temperature different of fluid in tube side as equation 2-17.

$$R = \frac{T_1 - T_2}{t_2 - t_1} = \frac{(mc_p)_{tube\ side}}{(mc_p)_{shell\ side}} \tag{2-17}$$

Where T_1 = inlet temperature of shell side (°C)

T_2 = outlet temperature of shell side (°C)

t_1 = inlet temperature of tube side (°C)

t_2 = outlet temperature of tube side (°C)

The R value does not consider with the hot fluid or cold fluid which flows either shell or tube. If some fluid have constant temperature, it does not consider with flow direction of that fluid because the correction factor is equal to 1 and can be used for LMTD.

2.6.2 NTU Effectiveness

The LMTD method is used when knowing the inlet and outlet temperature of heat exchanger , and then it can calculate to LMTD easily. However, The LMTD method is not the best method to estimate the average temperature different. The NTU Effectiveness method is the estimation of average temperature different better than the LMTD method because the NTU method is more advantage in the problem analysis and comparision between types of heat exchangers. The most of them are used as the reason for choosing the suitable type in each application. The efficiency of the heat exchanger is determined as equation 2-18

$$Effectveness\ (\eta) = \frac{q}{q_{max}} = \frac{Actual\ heat\ transfer\ rate}{Maximum\ heat\ transfer\ rate} \qquad (2\text{-}18)$$

Where q = actual heat transfer rate (kJ/h)

q_{max} = maximum possible heat transfer rate (kJ/h)

η = effectiveness

The actual heat transfer rate can be calculated from loss energy of hot fluid or receiving energy of cold fluid. From the conservation of energy law the heat transfer rate of hot fluid is equal to the heat transfer rate of cold fluid as equation 2-19

$$q = m_c Cp_c(T_{c.out} - T_{c.in}) = m_h Cp_h(T_{h.in} - T_{h,out}) \qquad (2\text{-}19)$$

The NTU effectiveness method need to use maximum possible heat transfer rate. Because the heat transfer rate is direct variation to the different temperature, the mamximum different temperature effect the maximum possible heat transfer rate which the definition as equation 2-20

$$q_{max} = (\dot{m}Cp)_{min}(T_{h.in} - T_{c.in}) \qquad (2\text{-}20)$$

Where C_p = specific heats capacity (kJ/kg °C)

m = mass flow rate of fluid (kg/h)

$T_{h,in}$ = inlet temperature of hot fluid (°C)

$T_{c,in}$ = outlet temperature of cold fluid (°C)

The number of transfer units (NTU) can be calculated by equation 2-21

$$NTU = \frac{UA_s}{c_{min}} = \frac{UA_s}{(\dot{m}Cp)_{min}} \quad (2\text{-}21)$$

Where C_p = specific heats capacity (kJ/kg °C)

m = mass flow rate of fluid (kg/h)

A_s = heat transfer area (m^2)

U = the overall heat transfer coefficient (W/m^2 °C)

NTU = number of transfer units

The UA_s/c_{min} is called "the number of transfer unit" that indicates to size of heat exchanger. Kays and London show the ratio of effectiveness for several types of heat exchanger and some result in the NTU relation table as Table 2.6 and Table 2.7. The NTU relation table can be calculated heat transfer rate accurately that is better than calculation by using graph. If the steps of design are complicate, it may be used computer programming to calculate and analyze them.

Table 2.6 NTU Relations for Heat Exchangers I (Kays and London) [Heat and Mass Transfer Fundamentals & Applications Yunus A. Cengel and Afshin J. Ghajar, 2015]

Heat Exchanger Type	Effectiveness Relation
Double Pipe Parallel-flow Counter-flow	$\eta = \frac{1 - exp[-NTU(1+c)]}{1+c}$ $\eta = \frac{1 - exp[-NTU(1-c)]}{1-(c)exp[-NTU(1-c)]}$
Shell-and-tube One Shell Pass 2, 4, . . . Tube Passes	$\eta = 2[1 + c + \sqrt{1+c^2}\frac{1+\exp(-NTU\sqrt{1+c^2})}{1-\exp(-NTU\sqrt{1+c^2})}]^{-1}$
Cross-flow (Single Pass) Both Fluids Unmixed C_{max} Mixed, C_{min} Unmixed C_{min} Mixed, C_{max} Unmixed	$\eta = 1 - exp\left\{\frac{NTU^{0.22}}{c}[\exp(-cNTU^{0.78}) - 1]\right\}$ $\eta = \frac{1}{c}(1 - exp\{1 - c[\exp(-NTU)]\})$ $\eta = 1 - exp\left\{-\frac{1}{c}[1 - \exp(-cNTU)]\right\}$
All Heat Exchangers with c = 0	$\eta = 1 - \exp(-NTU)$

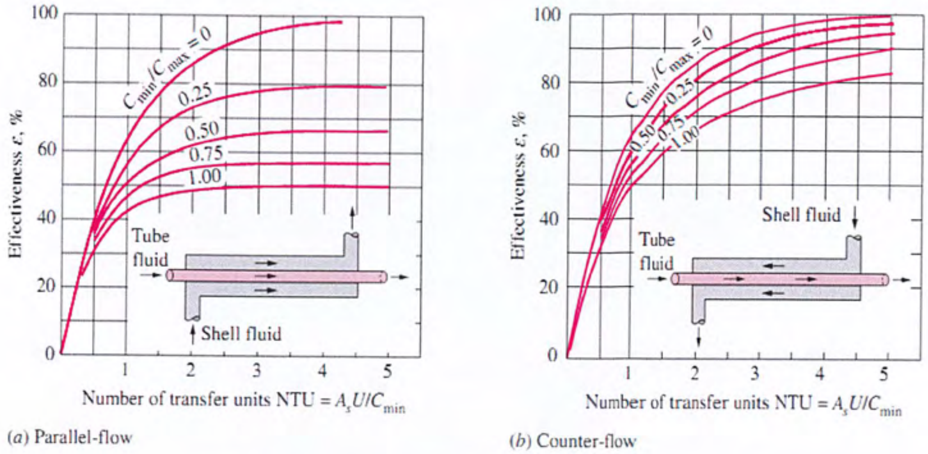


Figure 2.32 The Effectiveness for Double Pipe Heat Exchanger

Refference: Heat and Mass Transfer Fundamentals & Applications, Yunus A. Cengel and Afshin J. Ghajar, 2015

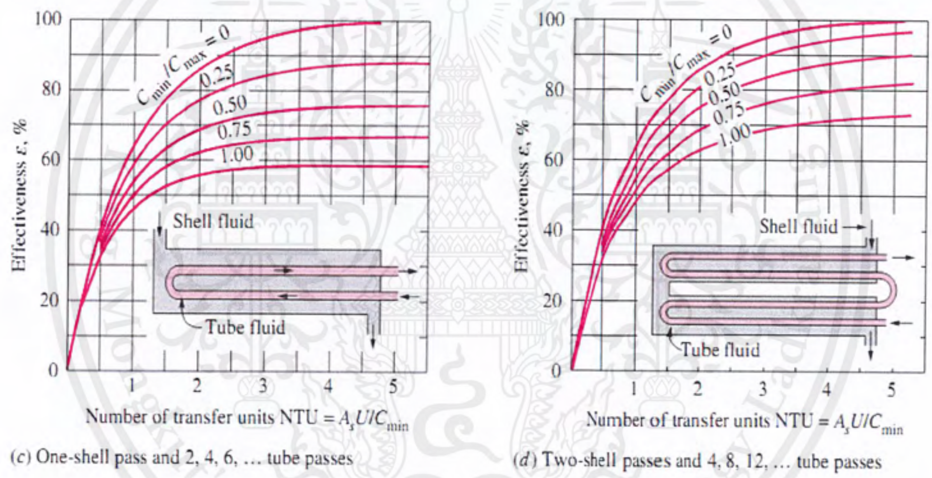


Figure 2.33 The Effectiveness for Shell and Tube Heat Exchanger

Refference: Heat and Mass Transfer Fundamentals & Applications, Yunus A. Cengel and Afshin J. Ghajar, 2015

Table 2.7 NTU Relations for Heat Exchangers II (Kays and London) [Heat and Mass Transfer Fundamentals & Applications Yunus A. Cengel and Afshin J. Ghajar, 2015]

Heat Exchanger Type	Effectiveness Relation
Double Pipe	
Parallel-flow	$\mathrm{NTU} = -\frac{\ln[1-\eta(1+c)]}{1+c}$
Counter-flow	$NTU = \frac{1}{c-1}\ln(\frac{\eta-1}{\eta c-1})$
Shell-and-tube One Shell Pass 2, 4, . . . Tube Passes	$NTU = -\frac{1}{\sqrt{1+c^2}}\ln(\frac{\frac{2}{\eta}-1-c-\sqrt{1+c^2}}{\frac{2}{\eta}-1-c+\sqrt{1+c^2}})$
Cross-flow (Single Pass)	
C_{max} Mixed, C_{min} Unmixed	$NTU = -\ln(1+\frac{\ln(1-\eta c)}{c})$
C_{min} Mixed, C_{max} Unmixed	$NTU = -\frac{\ln[c\ln(1-\eta)+1]}{c}$
All Heat Exchangers with c = 0	$NTU = -\ln(1-\eta)$

CHEPTER III

RESERCH AND METHODOLOGY

PART I Design the Level Control System of BFW

The aim of this part is to design level control system of the boiler feed water tank for collecting an amount of condensate from a new cogeneration plant. The phases of this work have 3 steps as follows

3.1 Balance Inlet and Outlet Condensate at the BFW Tank

The problem can be found from mass balance at the boiler feed water tank by calculating the inlet and outlet mass flow rate as shown in Figure 3.1

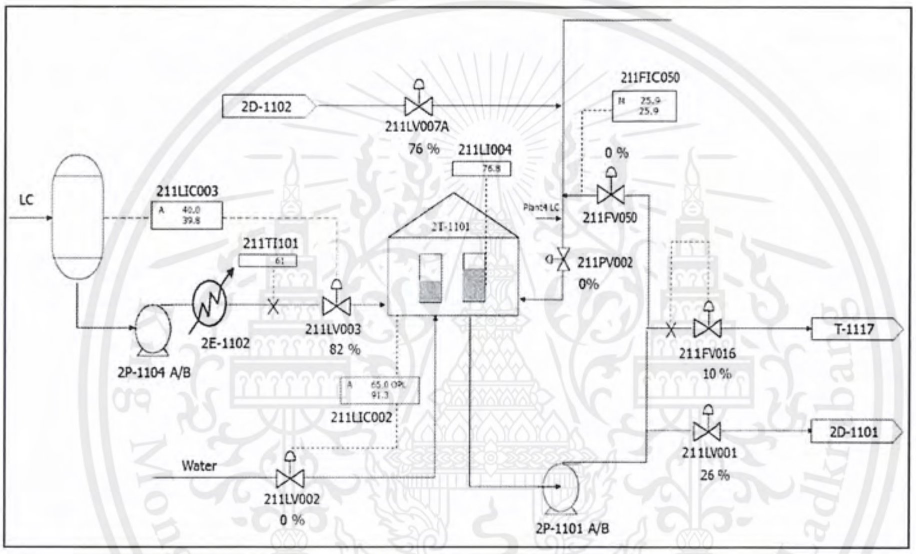


Figure 3.1 Simplified Diagram of the Boiler Feed Water

[Bangchak Petroleum Public Company Limited, 2017]

The inlet and outlet mass flow rate of condensate are calculated by equation 3-1 as follows

$$\dot{m} = \frac{nC_v\rho}{1.17}\sqrt{\frac{\Delta P}{SG}} \qquad (3\text{-}1)$$

Where m = mass flow rate of condensate (kg/h)

n = percentage of valve opening

C_v = flow coefficient of control valve

ρ = density of fluid (kg/m^3)

ΔP = pressure drop of control valve (kg/cm^3)

SG = ratio of density of fluid to the density of water

3.2 Study and Determine the Suitable Level Control System

Learning involve with types of control system for finding suitable type with this problem by condition of improvement is depend on the suitable type.

Budgets in this work are evaluated from instruments, equipment and computer programming. Benefits are evaluated in each design for calculation payback period and considering the best design.

PART II Study of the Vacuum System of a Hydrodesulfurization Unit

The aim of this part is to modify the gas oil hydrodesulfurization unit (DGO-HDSU) for decreasing the outlet temperature of the stream until the stream is condensed to all liquid phase which it can solve the high pressure in the vacuum dryer. The phases of this work have 3 steps as follows

3.1 Study the Possible Methods to Decrease the Outlet Stream Temperature

Learning design heat exchanger for decreasing outlet steam temperature and configuration of each exchanger in order to finding the best configuration and design in solving this problem

3.2 Calculate the Overall Heat Transfer Coefficient and Heat Transfer Rate

3.2.1 Calculate the overall heat transfer coefficient, which is ability of wall, is transferred heat including conduction, convection and radiation between hot and cold fluid by equation 3-2 as follows

$$U = \frac{Q}{A_s \Delta T_{lm}} \qquad (3\text{-}2)$$

Where U = the overall heat transfer coefficient ($kW/m^2\,^{\circ}C$)

Q = heat transfer rate (kW)

A_s = heat transfer area (m^2)

ΔT_{lm} = log mean temperature difference

Note: ΔT_{lm} is the log mean temperature difference which can be calculated by

$$\Delta T_{lm} = \frac{\Delta T_1 - \Delta T_2}{\ln(\Delta T_1/\Delta T_2)}$$

Where $\Delta T_1 = T_{hot,\,in} - T_{cold,\,out}$

$\Delta T_2 = T_{hot,\,out} - T_{cold,\,in}$

3.2.2 Calculate the number of transfer unit (NTU) that is the method for actual heat transfer rate calculation, and then calculate the outlet temperature by equation 3-3 (using for shell and tube 1 shell and 2n passed) and equation 3-4 as follows

$$NTU = -\frac{1}{\sqrt{1+c^2}} ln\left(\frac{2/\eta - 1 - c - \sqrt{1+c^2}}{2/\eta - 1 - c + \sqrt{1+c^2}}\right) \quad (3\text{-}3)$$

$$NTU = \frac{UA_s}{(\dot{m}Cp)_{min}} \quad (3\text{-}4)$$

Where NTU = number of transfer unit

c = $(mC_p)_{min} / (mC_p)_{max}$

η = effectiveness

U = the overall heat transfer coefficient (kW/m^2 °C)

A_s = heat transfer area (m^2)

$(mC_p)_{min}$ = the minimum heat capacity rate (kW/°C)

3.2.3 Calculate the maximum heat transfer rate (Q_{max}) by equation 3-5 as follows

$$Q_{max} = (\dot{m}Cp)_{min}(T_{h,in} - T_{c,in}) \quad (3\text{-}5)$$

Where Q_{max} = the maximum heat transfer rate (kW)

$(mC_p)_{min}$ = the minimum heat capacity rate (kW/°C)

$T_{h,in}$ = hot inlet temperature (°C)

$T_{c,out}$ = cold inlet temperature (°C)

3.2.4 Calculate the actual heat transfer rate (Q_{actual}) by using energy conservation and assume that the outlet surface of heat exchanger is insulated perfectly by equation 3-6 as follows

$$\dot{Q}_{actual} = \dot{m}_c Cp_c\left(T_{c,out} - T_{c,in}\right) = \dot{m}_h Cp_h\left(T_{h,in} - T_{h,out}\right) \qquad (3\text{-}6)$$

Where Q_{actual} = the actual heat transfer rate (kW)

m_h = mass flow rate of hot fluid (kg/h)

m_c = mass flow rate of cold fluid (kg/h)

Cp_h = specific heats of hot fluid (kJ/kg °C)

Cp_c = specific heats of cold fluid (kJ/kg °C)

$T_{h,in}$ = hot inlet temperature (° C)

$T_{c,in}$ = cold inlet temperature (° C)

$T_{h,out}$ = hot outlet temperature (° C)

$T_{c,out}$ = cold outlet temperature (° C)

3.2.5 Calculate the effectiveness (η) which is the ability to produce desired output of heat exchanger by equation 3-7 and equation 3-8 as follows

$$\eta = 2\left\{1 + c + \sqrt{1+c^2}\left(\frac{1 + exp\left(-NTU\sqrt{1+c^2}\right)}{1 - exp\left(-NTU\sqrt{1+c^2}\right)}\right)\right\}^{-1} \qquad (3\text{-}7)$$

$$\eta = \frac{Q}{Q_{max}} = \frac{Actual\ heat\ transfer\ rate}{Maximum\ possible\ heat\ transfer\ rate} \qquad (3\text{-}8)$$

Where η = effectiveness

c = $(mC_p)_{min} / (mC_p)_{max}$

NTU = number of transfer unit

Q = actual heat transfer rate (kW)

Q_{max} = the maximum heat transfer rate (kW)

3.3 Determine the Suitable Method to Decrease the Outlet Stream Temperature

Learning involve with heat exchanger for finding suitable design with this problem by condition of improvement is depend on the suitable design

Budgets in this work are evaluated from a new exchanger and valves. Benefits are evaluated in each design for calculation payback period and considering the best design.

CHAPTER IV

RESULTS AND DISCUSSION

PART I Design the Level Control System of BFW

4.1 Calculation Results from Mass Balance at the BFW Tank

Nowadays the most of inlet condensate come from the steam which have been already exchanged heat in many plants such as heat exchangers, distillation columns and so on. The mass flow rate of each line can be calculated by specification and percentage of valve opening of each control valve as shown in Figure 4.1

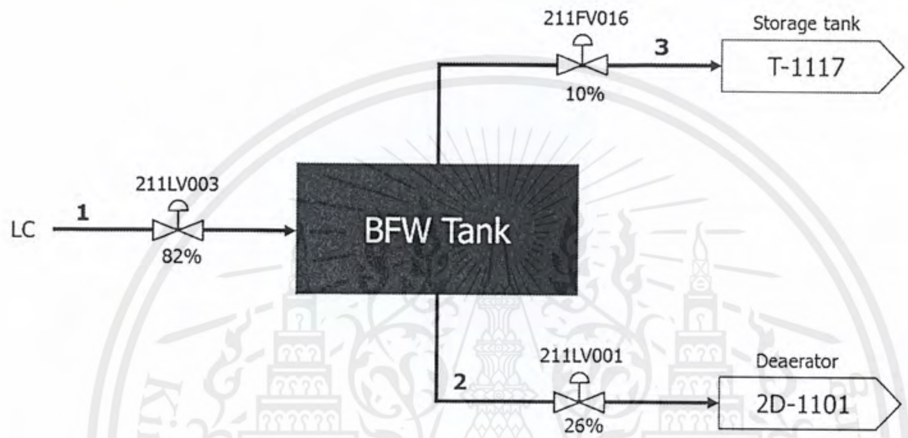


Figure 4.1 Mass Balance at the Boiler Feed Water Tank

Table 4.1 Mass Flow Rate of Each Line (Data for 21 August 2017 at 10:27:31)

Line No.	Valve Discharge	% Valve Opening	Mass Flow Rate (kg/h)
1	211LV003	82	65,654.29
2	211LV001	26	32,899.45
3	211FV016	10	11,442.12

Table 4.1 shows the calculation results from 3 lines which are inlet and outlet condensate flow rate. The mass flow rate calculation of each lines are shown in Appendix B. From mass balance at the boiler feed water tank, the inlet condensate flow rate is higher than the outlet condensate flow rate. Therefore, the problem which occurs overflow or excess mode.

4.2 Design and Selection Control System to Control Level of the Condensate

From the study of the principles of control system and boiler system, the analysis of level control system are divided by 2 parts as follows

4.2.1 Types of Control Systems

The control systems have 3 types for using in general industry

4.2.1.1 Cascade Control

4.2.1.2 Split Range Control

4.2.1.3 Override Control

After the appropriate type have been already chosen, it lead to the analysis of level control system which adapts this type with the boiler feed water tank. The split range control is the most suitable system for collecting the condensate from a new cogeneration plant.

4.2.2 Designs of Split Range Control

The designs of split ranger control are divided by 2 types

4.2.2.1 The first type is adjust a split range control from manual to automatic system by using controller (211LIC002) splits the signals to control valves (211LV002) and (211FV016) which define control valve (211LV002) is make up water and control valve (211FV016) is excess condensate as shown in Figure 4.2

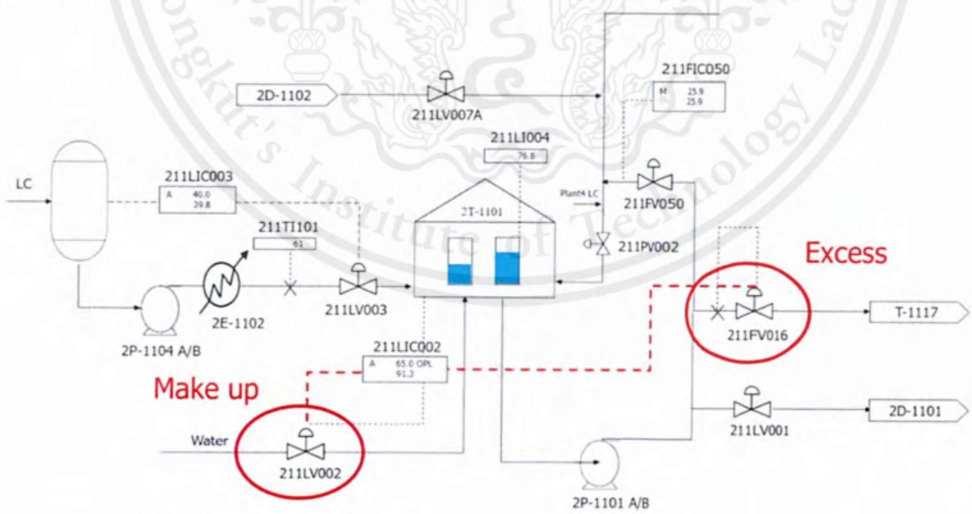


Figure 4.2 Simplified Diagram after Modification Type I

4.2.2.2 The second type is adjust a split range control from manual to automatic system by using indicator (211LI004) is converted to controller (211LIC004), then it splits the signals to control valves (211PV002) and (211FV016) which define control valve (211PV002) is make up condensate from the boiler feed water tank and control valve (211FV016) is excess condensate as shown in Figure 4.3

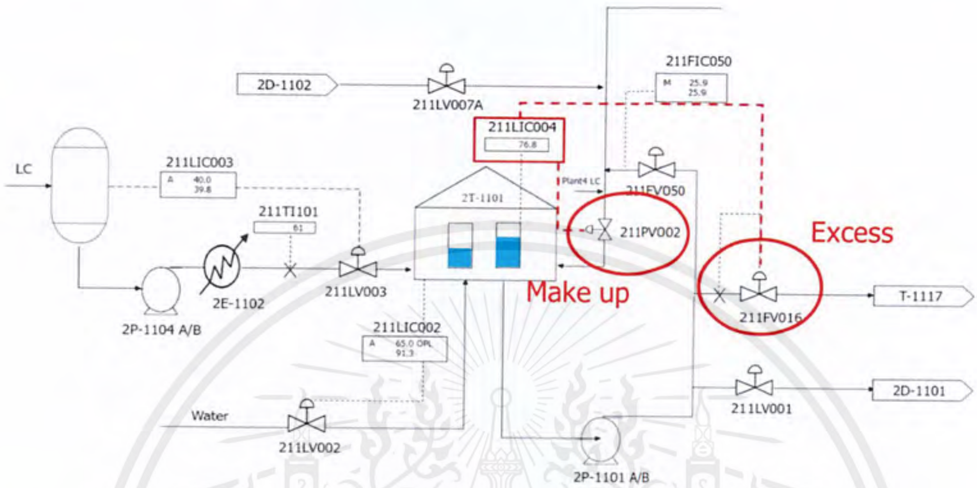


Figure 4.3 Simplified Diagram after Modification Type II

4.2.3 Selection Control System to Control Level of the Condensate

4.2.3.1 The types of control system by estimation from the advantages and disadvantages of each type as shown in Table 4.2, 4.3 and 4.4

Table 4.2 Advantages and Disadvantages of Cascade Control after Modification

Cascade Control	
Modification	Improved the level of condensate in the boiler feed water tank by controlling the inlet or outlet condensate flow rate
Advantages	1. Reduces the response time and phase lag time 2. Improved dynamic response and performance 3. Make the system control easily
Disadvantages	1. The system cannot control the inlet condensate flow rate because It effect another control system. 2. Cascade control makes the system more complex. 3. Cascade control requires more instruments and equipment. 4. Tuning of cascade controller is difficult.
Note	-

Table 4.3 Advantages and Disadvantages of Split Range Control after Modification

Split Range Control	
Modification	Improved the level of condensate in boiler feed water tank by controlling inlet and outlet condensate flow rate simultaneously
Advantages	1. Controlled the system effectively 2. Control the several control valve although it use one controller 3. Make the system good reliability 4. Require less instruments and equipment 5. Can be used for continuous measurement
Disadvantages	1. If the controller in spilt range control is fail, the system will be unbalance and some device may be damaged. 2. Split range control use the control valve more than one.
Note	-

Table 4.4 Advantages and Disadvantages of Override Control after Modification

Override Control	
Modification	Improved the level of condensate in boiler feed water tank by controlling inlet and outlet condensate flow rate but not occur simultaneously
Advantages	1. Used for control valve selection to control each line in that suitable condition such as emergency case 2. Improved dynamic response, performance and safety
Disadvantages	1. The control system can't use for continuous measurement. 2. The control system isn't suitable for level control. 3. Override control makes the system more complex. 4. Override control makes the system control difficultly.
Note	-

From the advantages and disadvantages of 3 types, the suitable type for controlling an amount of condensate is split range control because this problem requires to control both inlet and outlet condensate flow rate that occur simultaneously.

4.2.3.2 The design of split range control will be considered from the advantages and disadvantage of each type as shown in Table 4.5 and 4.6

Table 4.5 Advantages and Disadvantages of the First Type

The First Type	
Modification	Designed split range control from manual to automatic system by using controller (211LIC002) splits the signals to control valves (211LV002) and (211FV016) which define control valve (211LV002) is make up water and control valve (211FV016) is excess condensate.
Advantages	1. The system uses the same controller (211LIC002) which has redundant card for protecting a hardware failure. 2. This control system is not complicated.
Disadvantages	1. The control system uses make up water to maintain level of condensate in boiler feed water tank. It effect high cost. 2. The outlet temperature of condensate is lower than the outlet temperature of condensate from the second type.
Note	-

Table 4.6 Advantages and Disadvantages of the Second Type

The Second Type	
Modification	Designed split range control from manual to automatic system by using indicator (211LI004) is converted to controller (211LIC004) ,then it splits the signals to control valves (211PV002) and (211FV016) which define control valve (211PV002) is make up condensate and control valve (211FV016) is excess condensate.
Advantages	1. The control system uses recirculating condensate to maintain level of condensate which effects low cost. 2. This control system can use make up water to maintain level in tank if the recirculating condensate is not enough.
Disadvantages	This control system does not have the redundant card in a new controller (211LIC004) which effect protecting a hardware failure.
Note	The redundant card is equipment in controller which can protect some fault in process.

From the advantages and disadvantages of 2 type, the suitable type for controlling an amount of condensate is the second type because second type have good management at the boiler feed water tank and the cost is low.

Modification of level control system of the boiler feed water tank can reduce amount of steam in the deaerator tank because the deaerator tank will heat condensate which leave from the boiler feed water tank by using steam until oxygen and carbon dioxide in those condensate can be separated. From this study, the outlet temperature of condensate after modification is higher than the outlet temperature of condensate before modification equal to 10.8 $^{\circ}$C. It will reduce amount of the steam and operating cost approximately 2.54 kg/h and 15,296 baht/year respectively.

PART II Study of the Vacuum System of a Hydrodesulfurization Unit

4.1 The Possible Methods to Decrease the Outlet Stream Temperature

From the study of process description of gas oil hydrodesulfurization unit and principles of heat transfer, the suitable method for decreasing the outlet temperature of stream are 2 methods as follow

4.1.1 The first method is to install a new heat exchanger in series for decreasing the outlet temperature of the stream after passing through the heat exchanger (3E-3711) from 49.6 °C to 41.4 °C as shown in Figure 4.4

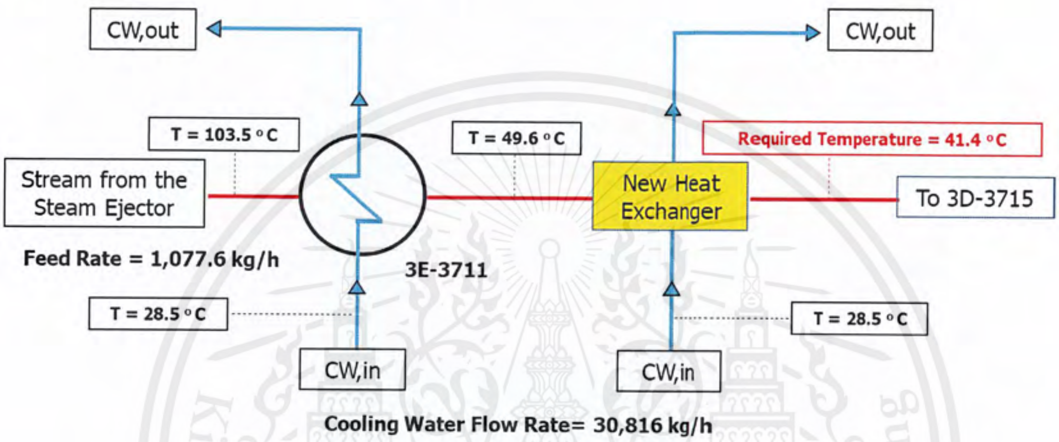


Figure 4.4 Simplified Diagram after Modification in Series

4.1.2 The second method is to design a new heat exchanger which the specification of new heat exchanger is the same as heat exchanger 3E-3711 and install it parallel with heat exchanger 3E-3711 for decreasing hot steam temperature to 41.4 °C as shown in Figure 4.5

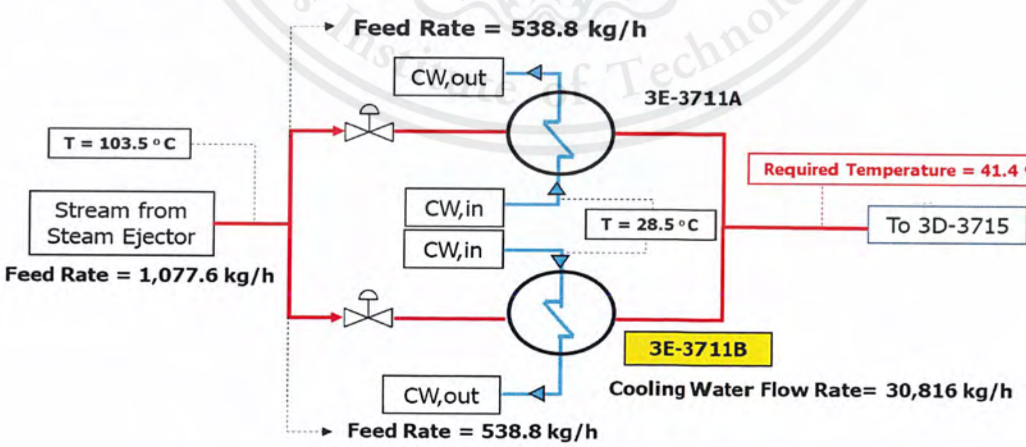


Figure 4.5 Simplified Diagram after Modification in Parallel

4.2 Calculation Results after Installation of a New Heat Exchanger

4.2.1 Calculation results of the first method by installed a new heat exchanger in series with the heat exchanger (3E-3711) for decreasing the outlet temperature of the stream as shown in Table 4.7

Table 4.7 Calculation Results for the First Method

Parameter	
The Outlet Temperature of the Steam from a New Heat Exchanger (°C)	41.4
The Outlet Temperature of the Cooling Water from a New Heat Exchanger (°C)	30.5
Heat Transfer Area of a New Heat Exchanger (m^2)	1.04

Where	Mass Flow Rate of Hot Stream	1,170.6	kg/h
	Specific Heats of Hot Stream at 50 °C	4.182	kJ/kg °C
	Inlet Temperature of Hot Stream	49.6	°C
	Mass Flow Rate of Cooling Water	30,816	kg/h
	Specific Heats of Cooling Water at 29 °C	4.178	kJ/kg °C
	Inlet Temperature of Cooling Water	28.5	°C

Note: Inlet temperature of hot stream and cooling water are average temperature from recording data during June to October 2017

4.2.2 Calculation results of the second method by installed a new heat exchanger which is the same specification of heat exchanger (3E-3711) in parallel for decreasing the outlet temperature of the stream as shown in Table 4.8

Table 4.8 Calculation Results for the Second Method

Parameter	
The Outlet Temperature of Stream from a New Heat Exchanger (°C)	34.6
The Outlet Temperature of Cooling Water from a New Heat Exchanger (°C)	29.6
Heat Transfer Area of a New Heat Exchanger (m^2)	30.2

Where			
	Mass Flow Rate of Hot Stream	538.8	kg/h
	Specific Heats of Hot Stream at 103.5 °C	4.226	kJ/kg °C
	Inlet Temperature of Hot Stream	103.5	°C
	Mass Flow Rate of Cooling Water	30,816	kg/h
	Specific Heats of Cooling Water at 29 °C	4.178	kJ/kg °C
	Inlet Temperature of Cooling Water	28.5	°C
	The Overall Heat Transfer Coefficient	0.054	$kW/m^2 °C$

Note: Inlet temperature of hot stream and cooling water are average temperature from recording data during June to October 2017

4.3 Selection the Suitable Method to Decrease the Outlet Stream Temperature

4.3.1 The first method will be considered from the advantages and the disadvantages as shown in Table 4.9

Table 4.9 Advantages and Disadvantages of the First Method

The First Method	
Modification	Installed a new heat exchanger in series for decreasing hot stream temperature after passing through the heat exchanger (3E-3711) by using cooling water
Advantages	1. Installation area is less than the second method. 2. Operating expense is lower than the second method. 3. The flow rate of hot stream is controlled easily.
Disadvantages	1. If one heat exchanger fails, the temperature of fluid is uncontrolled except the system has bypass. 2. Lifetime of a new heat exchanger is shorter than the second method because the flow rate of hot steam is higher than the second method.
Note	-

4.3.2 The second method will be considered from advantages and disadvantages as shown in Table 4.10

Table 4.10 Advantages and Disadvantages of the Second Method

The Second Method	
Modification	Installed a new heat exchanger that is the same as specification of heat exchanger 3E-3711 in parallel with heat exchanger (3E-3711) by using cooling water
Advantages	1. The outlet temperature of the stream is less than the first method. 2. The parallel type is better reliability than the series.
Disadvantages	1. The system is using more instruments to control flow rate in each line 2. The operating expense is high.
Note	-

From the advantages and disadvantages of 2 methods, the suitable method for decreasing the outlet temperature of condensate from heat exchanger (3E-3711) is the second method because the second method can reduce the outlet temperature of the stream better than the first method which the outlet stream is in liquid phase absolutely and the parallel is good reliability and performance as well.

CHAPTER V

CONCLUSIONS

PART I Design the Level Control System of BFW

5.1 Conclusion

The reliable method for collecting the amount of condensate and solving the overflow problem is the modification of level control system to become automatic split range control type II. This method can reduce the amount of the steam used in the deaerator tank which cost saving is approximately 15,300 Baht during operating time 360 days excluding the cost of condensate drained from the boiler feed water tank.

5.2 Suggestions

This project is modification of level control of the boiler feed water tank to resolve overflow and low level problems that the boiler feed water tank can be collected the amount of condensate.

5.2.1 This project does not consider with temperature of condensate after mixing which will be used in cooling system of compressor.

5.2.2 Tuning of controller (211LIC004) should be followed after modification for finding the set point to control level of condensate in the boiler feed water tank.

PART II Study of the Vacuum System of a Hydrodesulfurization Unit

5.1 Conclusion

From study the vacuum system in gas oil hydrodesulfurization unit, the method for solving high pressure problem in vacuum dryer is installation of a new shell and tube heat exchanger which having heat transfer area of 30.2 m^2. A new heat exchanger is installed to parallel with heat exchanger (3E-3711). The outlet temperature of the stream is 34.6 $^{\circ}$C that all of the streams before sending to vacuum pump are absolutely liquid phase. Therefore, the operating pressure in the vacuum dryer will be decreased and closed to the design pressure because the back pressure does not occur. Water content in diesel gas oil is lower than 120 mg/L.

5.2 Suggestions

5.2.1 Decreasing the outlet temperature of the stream from heat exchanger (3E-3711) is the one option for modification the vacuum system in gas oil hydrodesulfurization unit however, the other options can solve high pressure problem that are better or cheaper than this method.

5.2.2 The NTU method is not the best method to solve the heat exchanger problem. It may be used log mean temperature different (LMTD) to solve this problem.

5.2.3 In case of installation of a new heat exchanger in parallel with heat exchanger (3E-3711), the hot stream flow rate in each line must be higher than the design velocity of fluid that is passing through the heat exchanger because the fouling is not occurred.

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Appendix

APPENDIX A

Raw Data

Appendix A shows the raw data to refer the basic calculation of both parts

Table A.1 Specification of Each Control Valve at the Boiler Feed Water Tank.

[Bangchak Petroleum Public Company Limited, 2017]

Name	Specific Gravity	Mass Flow Rate (kg/h)		CV	Pressure (bar)		T (°C)
		normal	max		P_1	ΔP	
211LV001	0.983	10,425.0	112,590.0	58.0	7.20	6.50	60.0
211LV002	0.992	10,000.0	70,000.0	40.5	7.00	6.00	40.0
211LV003	0.947	5,090.0	49,882.0	72.0	2.70	1.70	115.0
211PV002	0.983	5,000.0	47,500.0	41.6	4.70	3.70	60.0

Table A.2 Before and After Modification of Level Control System at BFW Tank.

[Bangchak Petroleum Public Company Limited, 2017]

Parameters	Unit	Before Modification	After Modification
m_1	ton/h	10.23	10.23
m_2	ton/h	7.93	2.69
T_1	°C	60	60
T_2	°C	35	60

Where m_1 = mass flow rate of the make up line (ton/h)

m_2 = mass flow rate of the excess line (ton/h)

T_1 = temperature of the make up line (°C)

T_2 = temperature of the excess line (°C)

Note: The price of steam used in the deaerator is approximately 695.7 Baht/ton

Table A.3 Design of Shell and Tube Heat Exchanger 3E-3711

[Bangchak Petroleum Public Company Limited, 2017]

3E-3711		Design Condition			
		Shell Side		Tube Side	
		Inlet	Outlet	Inlet	Outlet
Fluid Name		Steam		Cooling Water	
Mass flow rate	kg/h	898		25,680	
Temperature	°C	100.9	51.7	35	48.9
Specific Heat	kJ/kg K	4.2175	4.1825	4.1799	4.1805
Inlet Pressure	bar	1.136		5.116	
Heat transfer area	m^2	30.2			

Table A.4 Design of Shell and Tube Heat Exchanger 3E-3712

[Bangchak Petroleum Public Company Limited, 2017]

3E-3712		Design Condition			
		Shell Side		Tube Side	
		Inlet	Outlet	Inlet	Outlet
Fluid Name		Seal Liquid		Cooling Water	
Mass flow rate	kg/h	2,268		8,546	
Temperature	°C	67.1	46.1	35	40.6
Specific Heat	kJ/kg K	4.186	4.187	4.189	4.188
Inlet Pressure	bar	1.724		5.116	
Heat transfer area	m^2	5.2			

Table A.5 Design Condition of Vacuum Pump Decanter 3D-3715

[Bangchak Petroleum Public Company Limited, 2017]

3D-3715	Design Condition
Fluid Name	Seal Liquid
Phase	All Liquid
Mass flow Rate (kg/hr)	889
Temperature (°C)	41.4
Specific Heats (kJ/kg K)	4.180
Inlet Pressure (bar)	1.01

Table A.6 Saturation Properties - Pressure Table for 1 kPa - 1 MPa

[Thermodynamics An Engineering Approach Eighth Edition, Yunus A. Cengel and Michael A. Boles. 2015.]

Pressure	Temp	Energy (kJ/kg)		Enthalpy (kJ/kg)			Entropy (kJ/kg.K)		
kPa	°C	u_f	u_g	h_f	h_{fg}	h_g	s_f	s_{fg}	s_g
1	6.97	29.3	2384.5	29.3	2484.4	2513.7	0.1059	8.869	8.9749
1.2	9.65	40.6	2388.2	40.6	2478	2518.6	0.146	8.7622	8.9082
1.4	11.97	50.3	2391.3	50.3	2472.5	2522.8	0.1802	8.672	8.8522
1.6	14.01	58.8	2394.1	58.8	2467.7	2526.5	0.21	8.5935	8.8035
1.8	15.84	66.5	2396.6	66.5	2463.4	2529.9	0.2366	8.5242	8.7608
2	17.5	73.4	2398.9	73.4	2459.5	2532.9	0.2606	8.462	8.7226
3	24.08	101	2407.9	101	2443.8	2544.8	0.3543	8.2221	8.5764
4	28.96	121.4	2414.5	121.4	2432.3	2553.7	0.4224	8.051	8.4734
6	36.16	151.5	2424.2	151.5	2415.1	2566.6	0.5208	7.8082	8.329
8	41.51	173.8	2431.4	173.8	2402.4	2576.2	0.5925	7.6348	8.2273
10	45.81	191.8	2437.2	191.8	2392.1	2583.9	0.6492	7.4996	8.1488
12	49.42	206.9	2442	206.9	2383.4	2590.3	0.6963	7.3886	8.0849
14	52.55	220	2446.1	220	2375.8	2595.8	0.7366	7.2945	8.0311
16	55.31	231.6	2449.8	231.6	2369	2600.6	0.772	7.2126	7.9846
18	57.8	242	2453	242	2363	2605	0.8036	7.1401	7.9437
20	60.06	251.4	2456	251.4	2357.5	2608.9	0.832	7.0752	7.9072
30	69.1	289.2	2467.7	289.3	2335.2	2624.5	0.9441	6.8234	7.7675
40	75.86	317.6	2476.3	317.6	2318.5	2636.1	1.0261	6.6429	7.669
60	85.93	360	2489	359.9	2293	2652.9	1.1454	6.3857	7.5311
80	93.49	391.6	2498.2	391.7	2273.5	2665.2	1.233	6.2009	7.4339
100	99.61	417.4	2505.6	417.5	2257.4	2674.9	1.3028	6.056	7.3588
120	104.78	439.2	2511.7	439.4	2243.7	2683.1	1.3609	5.9368	7.2977
140	109.29	458.3	2516.9	458.4	2231.6	2690	1.411	5.8351	7.2461
160	113.3	475.2	2521.4	475.4	2220.6	2696	1.4551	5.7463	7.2014
180	116.91	490.5	2525.5	490.7	2210.7	2701.4	1.4945	5.6676	7.1621
200	120.21	504.5	2529.1	504.7	2201.5	2706.2	1.5302	5.5967	7.1269
300	133.52	561.1	2543.2	561.4	2163.5	2724.9	1.6717	5.3199	6.9916
400	143.61	604.2	2553.1	604.7	2133.4	2738.1	1.7765	5.119	6.8955
600	158.83	669.7	2566.8	670.4	2085.7	2756.1	1.9308	4.8284	6.7592
800	170.41	720	2576	720.9	2047.4	2768.3	2.0457	4.6159	6.6616
1000	179.88	761.4	2582.7	762.5	2014.6	2777.1	2.1381	4.4469	6.585

APPENDIX B

MASS BALANCE AT BOILER FEED WATER TANK

Appendix B shows mass balance at boiler feed water tank where is in the utility plant for finding and solving the problems. The calculation data as shown in the Table B.1

Table A.1 Actual Percentage of Valve Opening of Each Control Valve

[Bangchak Petroleum Public Company Limited, 2017]

No.	Control Valve	% Valve Opening
1	211LV002	0
2	211LV003	82
3	211FV050	0
4	211LV007A	76
5	211FV009	0
6	211PV002	0
7	211LV001	26
8	211FV016	10

Note: The data are recorded at 21 August 2017

1. Mass Balance at the Boiler Feed Water Tank Calculation

2.1 Calculate mass flow rate of fluid that is passing through the control valve 211LV003 as follows

$$\dot{m} = \frac{nC_v\rho}{1.17}\sqrt{\frac{\Delta P}{SG}} \qquad (3\text{-}1)$$

Where

n	=	0.82	
C_v	=	72.0	kJ/kg °C
ρ	=	0.947	kg/m^3
ΔP	=	1.7335	kg/cm^3
SG	=	0.947	

Therefore, the mass flow rate is approximately

$$m = \frac{0.82 \times 72.0 \times 0.947}{1.17}\sqrt{\frac{1.7335}{0.947}} \text{ kg/h}$$

$$= 64{,}654.2987 \text{ kg/h}$$

1.2 Balance the inlet and outlet mass flow rate of fluid at the boiler feed water by using the law of conservation of mass in the close system as equation

$$In = Out$$

From Figure 4.1, the equation is

$$\dot{m}_1 = \dot{m}_2 + \dot{m}_3$$

Where	$\dot{m}_1$	=	64,654	kg/h
	$\dot{m}_2$	=	32,899	kg/h
	$\dot{m}_3$	=	11,442	kg/h

Therefore

64,654 = 32,899 + 11,442

64,654 < 44,341

APPENDIX C

HEAT BALANCE AT BOILER FEED WATER TANK

Appendix C shows heat balance at boiler feed water tank for calculation the amount of reducing steam used in the deaerator tank and economic evaluation

1. Calculation Data [Bangchak Petroleum Public Co., LTD, 2017]

For Case I

Mass Flow Rate of Inlet Condensate	=	10.23	ton/h
Mass Flow Rate of Make up Line	=	7.93	ton/h
Inlet Temperature of Condensate	=	60	°C
Inlet Temperature of Make up Line	=	35	°C
Specific Heats of Inlet Condensate at 60 °C	=	4.185	kJ/kg °C
Specific Heats of Make up Line at 35 °C	=	4.178	kJ/kg °C

For Case II

Mass Flow Rate of Inlet Condensate	=	10.23	ton/h
Mass Flow Rate of Make up Line	=	2.69	ton/h
Inlet Temperature of Condensate	=	60	°C
Inlet Temperature of Make up Line	=	60	°C
Specific Heats of Inlet Condensate at 60 °C	=	4.185	kJ/kg °C
Specific Heats of Make up Line at 60 °C	=	4.185	kJ/kg °C

2. An Amount of Reducing Steam Calculation

2.1 For case I, the total heat transfer rate is calculated by heat balance at the boiler feed water as follows

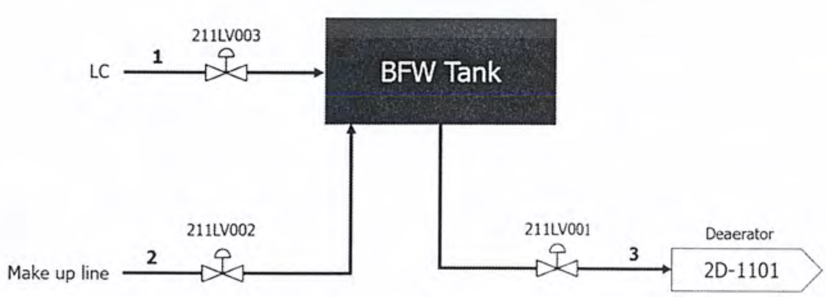

Figure C.1 Heat Balance at the Boiler Feed Water Tank Diagram in Case I

From Figure C.1, the equation is

$$Q_1 + Q_2 = Q_3$$

And then

$$m_1 c_{p_1}(T_1 - T_3) + m_2 c_{p_2}(T_2 - T_3) = Q_3$$

Where	$\dot{m}_1$	=	10.23	ton/h
	$\dot{m}_2$	=	7.93	ton/h
	c_{p_1}	=	4.185	kJ/kg °C
	c_{p_2}	=	4.178	kJ/kg °C
	T_1	=	60	°C
	T_2	=	35	°C
	T_3	=	49.189	°C

Therefore, the total heat transfer rate is approximately

Q_3 = 10.23 (1,000) (4.185) (60-49.189)

+7.93(1,000) (4.178) (35-49.189) kJ/h

= -7256.943 kJ/h

Or = -2.0158 kW

2.2 For case II, the total heat transfer rate is calculated by heat balance at the boiler feed water as shown in Figure C.2

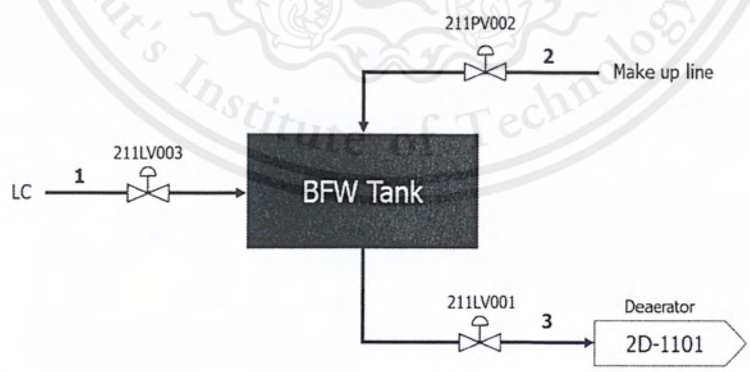

Figure C.2 Heat Balance at the Boiler Feed Water Tank Diagram in Case II

From Figure C.2, the equation is

$$Q_1 + Q_2 = Q_3$$

And then

$$m_1 c_{p_1}(T_1 - T_3) + m_2 c_{p_2}(T_2 - T_3) \quad = \quad Q_3$$

Where	$\dot{m}_1$	=	10.23	ton/h
	$\dot{m}_2$	=	2.69	ton/h
	c_{p_1}	=	4.185	kJ/kg °C
	c_{p_2}	=	4.185	kJ/kg °C
	T_1	=	60	°C
	T_2	=	60	°C
	T_3	=	60	°C

Therefore, the total heat transfer rate is approximately

Q_3 = 10.23 (1,000) (4.185) (60-60)

+2.69 (1,000) (4.185) (60-60) kJ/h

= 0 kJ/h

2.3 Calculate an amount of reducing steam used in the deaerator tank

In case II, it does not have the amount of loss energy which it can increase the outlet temperature of condensate after passing through the boiler feed water and reduce the amount of steam in deaerator for heating condensate to separate oxygen and carbon dioxide gas. After modification, the outlet temperature of condensate is increased to 60 ° °C and the amount of loss energy is 2.0158 kW

Therefore, the amount of reducing steam is approximately

$$\dot{Q}_3 = \dot{m}_{SL} \Delta H_{sat}$$

Where ΔH_{sat} = enthalpy of steam at 3.4 barg, 185 °C = 2,826.9 kJ/kg

$\dot{Q}_3$ = the amount of loss energy = 7,256.9 kJ

And $\dot{m}_{SL}$ = 7,256.9 / 2,826.9

= 2.567 kg/h

The saving cost of reducing steam is approximately

$$\text{Saving cost} = \frac{2.567\ kg}{1\ hr} \times \frac{1\ ton}{1{,}000\ kg} \times \frac{695.7\ baht}{1\ ton}$$

$$= 1.7859 \text{ baht/h}$$

The operating time is approximately 360 day/year

$$\text{Saving cost} = \frac{1.7859\ baht}{1\ hr} \times \frac{24\ hr}{1\ day} \times \frac{360\ day}{1\ year}$$

$$= 15{,}430.176 \text{ baht/year}$$

APPENDIX D

EXAMPLE OF CALCULATION FOR SERIES TYPE

Appendix D shows the example calculation for the first method which is to design a new heat exchanger and install it next to heat exchanger (3E-3711) in series for decreasing the outlet temperature of stream.

1. **Calculation Data [Bangchak Petroleum Public Co.,LTD, 2017]**

Mass Flow Rate of Inlet Stream	=	1,077.6	kg/h
Specific Heats of Stream at 49.6 °C	=	4.182	kJ/kg °C
Actual Inlet Temperature of Stream	=	49.6	°C
Require Outlet Temperature of Stream	=	41.4	°C
Mass Flow Rate of Cooling Water	=	30,816	kg/h
Specific Heats of Cooling Water at 28.5 °C	=	4.178	kJ/kg °C
Actual Inlet Temperature of Cooling Water	=	28.5	°C

2. **Heat Transfer Area of a New Heat Exchanger Calculation**

2.1 Calculate the Maximum Heat Transfer Rate (Q_{max})

$$\dot{Q} = \dot{m}_h c_{ph}(T_{h,in} - T_{c,in}) \quad (3\text{-}5)$$

Where

m_h	=	1,077.6	kg/h
$C_{p,h}$	=	4.182	kJ/kg °C
$T_{h,in}$	=	49.6	°C
$T_{c,in}$	=	28.5	°C

Therefore, the maximum heat transfer rate is approximately

Q_{max}	=	(1,077.6) (4.182) (49.6-28.5)	kJ/h
	=	95,087.6395	kJ/h
Or	=	26.4130	kW

2.2 Calculate Actual Heat Transfer Rate (Q_{Actual})

$$\dot{Q} = \dot{m}_h c_{ph} \left(T_{h,in} - T_{h,out}\right) \tag{3-6}$$

Where	m_h	=	1,077.6 kg/h	
	$C_{p,h}$	=	4.182	kJ/kg °C
	$T_{h,in}$	=	49.6	°C
	$T_{c,in}$	=	41.4	°C (Assume)

Therefore, the actual heat transfer rate is approximately

Q_{max}	=	(1,077.6) (4.182) (49.6-41.4)	kJ/h
	=	37,133.7512	kJ/h
Or	=	10.3148	kW

2.3 Calculate Effectiveness (η)

$$\eta = \frac{Q}{Q_{max}} = \frac{Actual\ heat\ transfer\ rate}{Maximum\ possible\ heat\ transfer\ rate} \tag{3-7}$$

Where	Q_{Actual}	=	10.3148	kW
	Q_{max}	=	26.4130	kW

Therefore, the effectiveness is approximately

η	=	10.3148 / 26.4130
	=	0.6867

2.4 Calculate Number of Transfer Unit (NTU)

$$NTU = -\frac{1}{\sqrt{1+c^2}} \, ln\left(\frac{2/\eta - 1 - c - \sqrt{1+c^2}}{2/\eta - 1 - c + \sqrt{1+c^2}}\right) \tag{3-3}$$

Where	c	=	0.035
	η	=	0.6867

Therefore, the number of transfer unit is approximately

$$NTU = -\frac{1}{\sqrt{1+(0.035)^2}} \, ln\left(\frac{2/0.6867 - 1 - 0.035 - \sqrt{1+(0.035)^2}}{2/0.6867 - 1 - 0.035 + \sqrt{1+(0.035)^2}}\right)$$

$$= 0.4995$$

2.3 Calculate Heat Transfer Area of a New Heat Exchanger (A_s)

$$A_s = \frac{NTU \times (\dot{m}Cp)_{min}}{U} \qquad (3\text{-}4)$$

Where	NTU	=	0.4995		
	m_h	=	1,077.6	kg/h	
	$C_{p,h}$	=	4.182	kJ/kg °C	
	$U_{service}$	=	0.6	kW/m^2 °C	[Coulson,1999]

Therefore, the heat transfer area is approximately

$$A_s = \frac{0.4995 \times 1{,}077.6 \times 4.182}{0.6} \quad m^2$$

$$= 1.0421 \quad m^2$$

APPENDIX E

EXAMPLE OF CALCULATION FOR THE PARALLEL TYPE

Appendix E shows the example calculation for the second method which is to design a heat exchanger as a (3E-3711) specification and install it next to heat exchanger (3E-3711) in parallel for decreasing the outlet temperature of the steam.

1. Calculation Data [Bangchak Petroleum Public Co.,LTD, 2017]

Mass Flow Rate of Inlet Stream	=	538.8	kg/h
Specific Heats of Stream at 49.6 °C	=	4.182	kJ/kg °C
Actual Inlet Temperature of Stream	=	49.6	°C
Require Outlet Temperature of Stream	=	41.4	°C
Mass Flow Rate of Cooling Water	=	30,816	kg/h
Specific Heats of Cooling Water at 28.5 °C	=	4.178	kJ/kg °C
Actual Inlet Temperature of Cooling Water	=	28.5	°C

2. The Actual Overall Heat Transfer Coefficient of Heat Exchanger Calculation

2.1 Calculate the Maximum Heat Transfer Rate (Q_{max})

$$\dot{Q} = \dot{m}_h c_{ph}\left(T_{h.in} - T_{c.in}\right) \tag{3-5}$$

Where

m_h	=	538.8	kg/h
$C_{p,h}$	=	4.260	kJ/kg °C
$T_{h,in}$	=	103.5	°C
$T_{c,in}$	=	28.5	°C

Therefore, the maximum heat transfer rate is approximately

Q_{max}	=	(1077.6) (4.182) (103.5-28.5)	kJ/h
	=	341,545.32	kJ/h
Or	=	94.8737	kW

2.2 Calculate Actual Heat Transfer Rate (Q_{Actual})

$$\dot{Q} = \dot{m}_h c_{ph}(T_{h,in} - T_{h,out}) \quad (3\text{-}6)$$

Where	m_h	=	1077.6	kg/h
	$C_{p,h}$	=	4.260	kJ/kg °C
	$T_{h,in}$	=	103.5	°C
	$T_{h,out}$	=	49.6	°C

Therefore, the actual heat transfer rate is approximately

Q_{max}	=	(1,077.6) (4.182) (103.5-49.6)kJ/h	
	=	245,457.36	kJ/h
Or	=	68.1826	kW

2.3 Calculate Effectiveness (η)

$$\eta = \frac{Q}{Q_{max}} = \frac{Actual\ heat\ transfer\ rate}{Maximum\ possible\ heat\ transfer\ rate} \quad (3\text{-}8)$$

Where	Q_{Actual}	=	68.1826	kW
	Q_{max}	=	94.8737	kW

Therefore, the effectiveness is approximately

η	=	68.1826 / 94.8737
	=	0.7186

2.4 Calculate Number of Transfer Unit (NTU)

$$NTU = -\frac{1}{\sqrt{1+c^2}} ln\left(\frac{2/\eta - 1 - c - \sqrt{1+c^2}}{2/\eta - 1 - c + \sqrt{1+c^2}}\right) \quad (3\text{-}3)$$

Where	c	=	0.0356
	η	=	0.7186

Therefore, the number of transfer unit is approximately

$$NTU = -\frac{1}{\sqrt{1+(0.035)^2}} ln\left(\frac{2/0.7186 - 1 - 0.035 - \sqrt{1+(0.035)^2}}{2/0.7186 - 1 - 0.035 + \sqrt{1+(0.035)^2}}\right)$$

$= 1.3017$

2.5 Calculate the Overall Heat Transfer Coefficient (U)

$$U = \frac{NTU \times (\dot{m}Cp)_{min}}{A_s} \tag{3-4}$$

Where	NTU	=	1.3017	
	m_h	=	1,077.6	kg/h
	$C_{p,h}$	=	4.260	kJ/kg °C
	A_s	=	30.2	m^2

Therefore, the overall heat transfer coefficient is approximately

$$U = \frac{1.3017 \times 1{,}077.6 \times 4.182}{30.2} \quad m^2$$

$$= 0.054 \quad kW/m^2\,°C$$

3. The Outlet Temperature of the Stream Calculation

3.1 Calculate the Maximum Heat Transfer Rate (Q_{max})

$$\dot{Q} = \dot{m}_h c_{ph}\left(T_{h,in} - T_{c,in}\right) \tag{3-5}$$

Where	m_h	=	538.8	kg/h
	$C_{p,h}$	=	4.182	kJ/kg °C
	$T_{h,in}$	=	103.5	°C
	$T_{c,in}$	=	28.5	°C

Therefore, the actual heat transfer rate is approximately

Q_{max}	=	(538.8) (4.182) (103.5-28.5)	kJ/h
	=	170,775	kJ/h
Or	=	47.4375	kW

3.2 Calculate Number of Transfer Unit (NTU)

$$NTU = \frac{A_s \times U}{(\dot{m}Cp)_{min}} \tag{3-4}$$

Where	A_s	=	30.2	m^2
	U	=	0.054	kW/m^2 °C
	m_h	=	538.8	kg/h
	$C_{p,h}$	=	4.182	kJ/kg °C

Therefore, the number of transfer unit is approximately

$$NTU = \frac{1.3017 \times 1{,}077.6 \times 4.182}{30.2}$$

$$= 2.6022$$

3.3 Calculate Effectiveness (η)

$$\eta = 2\left\{1 + c + \sqrt{1+c^2}\left(\frac{1+exp\left(-NTU\sqrt{1+c^2}\right)}{1-exp\left(-NTU\sqrt{1+c^2}\right)}\right)\right\}^{-1} \quad (3\text{-}7)$$

Where c = 0.0356

NTU = 2.6022

Therefore, the effectiveness is approximately

$$\eta = 2\left\{1 + 0.0356 + \sqrt{1+(0.0356)^2}\left(\frac{1+exp\left(-NTU\sqrt{1+c^2}\right)}{1-exp\left(-NTU\sqrt{1+c^2}\right)}\right)\right\}^{-1}$$

$$= 0.9183$$

3.4 Calculate Actual Heat Transfer Rate (Q_{Actual})

$$\eta = \frac{Q}{Q_{max}} = \frac{Actual\ heat\ transfer\ rate}{Maximum\ possible\ heat\ transfer\ rate} \quad (3\text{-}8)$$

Where Q_{max} = 47.4375 kW

η = 0.9183

Therefore, the actual heat transfer rate is approximately

Q_{actual} = 47.4375 × 0.9183 kW

= 43.5627 kW

3.5 Calculate the Outlet Temperature of Stream ($T_{h,out}$)

$$\dot{Q} = \dot{m}_h c_{ph}\left(T_{h,in} - T_{h,out}\right) \quad (3\text{-}6)$$

Where m_h = 538.8 kg/h

$C_{p,h}$ = 4.182 kJ/kg °C

$T_{h,in}$ = 49.6 °C

Q_{actual} = 43.5627 kW

Therefore, the outlet temperature of the stream is approximately

$$T_{h,out} = 49.6 - \frac{43.5627}{538.8 \times 4.182} \quad °C$$

$$= 34.6 \quad °C$$

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  Internship Program 2017

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  Co-operative Program 2017