การสร้างแบบจำลองและหาภาวะที่เหมาะสมสำหรับหน่วยอัดความดันของก๊าซที่ได้จาก การแครกกิ้งในโรงงานเอทิลีน

นางสาวปิยฉัตร พุทธรักษา

วิทยานิพนธ์นี้เป็นส่วนหนึ่งของการศึกษาตามหลักสูตรปริญญาวิศวกรรมศาสตรมหาบัณฑิต

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ปีการศึกษา 2553

SIMULATION AND OPTIMIZATION FOR COMPRESSOR UNIT OF CRACKED GAS IN ETHYLENE PLANT

Miss Piyachat Puttaraksa

A Thesis Submitted in Partial Fulfillment of the Requirements
for the Degree of Master of Engineering Program in Chemical Engineering
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By	Miss Piyachat Puttararaksa
Field of Study	Chemical Engineering
Thesis Advisor	Professor Paisan Kittisupakorn, Ph.D.

Accepted by the Faculty of Engineering, Chulalongkorn University in Partial Fulfillment of the Requirements for the Master's Degree

....Dean of the Faculty of Engineering

(Associate Professor Boonsom Lerdhirunwong, Dr.Ing.)

THESIS COMMITTEE

....Chairman

(Professor Piyasan Praserthdam, Dr.Ing.)

....Thesis Advisor

(Professor Paisan Kittisupakorn, Ph.D.)

....Examiner

(Assistant Professor Anongnat Somwangthanaroj, Ph.D.)

....External Examiner

(Wachira Daosud,Dr.)

ปิยฉัตร พุทธรักษา : การสร้างแบบจำลองและหาภาวะที่เหมาะสมสำหรับหน่วยอัดความดันของก๊าซที่ได้จากการแครกกิ้งในโรงงานเอทิลีน. (SIMULATION AND OPTIMIZATION FOR COMPRESSOR UNIT OF CRACKED GAS IN ETHYLENE PLANT) อ. ที่ปรึกษาวิทยานิพนธ์หลัก: ศ.ดร. ไพศาล กิตติศุภกร, 82 หน้า.

งานวิจัยนี้มุ่งถึงการศึกษาปัจจัยที่เกี่ยวข้องกับการสะสมอุดตันโพลิเมอร์ในเครื่องอัดความดัน ที่ใช้อัดก๊าซที่ได้จากการแครกกิ้งในโรงงานเอทิลีน ด้วยการใช้รูปจำลองและแบบจำลองทางโปรแกรม HYSYS ในโรงงานผลิตโอเลฟินส์ที่ใช้ก๊าซผลิตมักใช้ wash oil ชะล้างหรือละลายโพลิเมอร์ที่เกิดขึ้นภายในเครื่องอัดความดัน และใช้ wash water ลดอุณหภูมิขาออกของเครื่องอัดความดัน เพื่อลดการสะสมอุดตันโพลิเมอร์ซึ่งปริมาณการใช้นั้นมักขึ้นอยู่กับค่าการออกแบบภาวะการทำงานของแต่ละโรงงาน ในทางปฏิบัติจริงนั้น ยากที่จะสามารถหาปริมาณที่เหมาะสมของ wash oil ได้ เนื่องจากปริมาณ wash oil ไม่ส่งผลต่ออุณหภูมิขาออกของเครื่องอัดความดันที่เป็นปัจจัยหลักในการพิจารณาอัตราการเกิดโพลิเมอร์ในเครื่องอัดความดันที่ใช้อัดก๊าซที่ได้จากการแครกกิ้ง ดังนั้นเมื่อพิจารณาอีกหนึ่งปัจจัยหลักคือ wash water ซึ่งสามารถใช้แล้วสามารถสังเกตการเปลี่ยนแปลงทางอุณหภูมิขาออกของเครื่องอัดความดันได้ชัดเจน wash water จึงถูกเลือกมาใช้สำหรับการทดลองนี้ด้วยโปรแกรม HYSYS ในการหาแบบจำลองทางกระบวนการด้วยโปรแกรม HYSYS ข้อมูลที่เป็นภาวะการทำงานของก๊าซที่ได้จากการแครกจะถูกรวบรวมและแปลผลหารูปจำลองทาง HYSYS และรูปจำลองที่ถูกต้องเหมาะสมที่สุดจะถูกนำมาใช้งาน ด้วยรูปจำลองที่ถูกต้องนั้นทำให้สามารถหาค่า wash water ที่เหมาะสมได้ ด้วยการปรับใช้ปริมาณ wash water ที่ค่า 1,200 ลิตรต่อชั่วโมงถึง 100 ลิตรต่อชั่วโมงเพื่อหาปริมาณการใช้ที่ทำให้อุณหภูมิขาออกของเครื่องอัดความดันมีค่าอย่างมากอยู่ที่ 90 องศาเซลเซียส จากการทดลองพบว่าเครื่องอัดความดันที่ 1, 2 และ 3 ได้ปริมาณอัตราการใช้ wash water ที่เหมาะสมดังนี้คือ 465 ลิตรต่อชั่วโมง, 380 ลิตรต่อชั่วโมงและ 505 ลิตรต่อชั่วโมง ตามลำดับ

ภาควิชา วิศวกรรมเคมี ลายมือชื่อนิสิต ปิยฉัตร พุทธรักษา
สาขาวิชา วิศวกรรมเคมี ลายมือชื่อ อ.ที่ปรึกษาวิทยานิพนธ์หลัก ไพศาล กิตติศุภกร
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KEYWORDS : ETHYLENE PLANT / COMPRESSOR / FOULING / WASH WATER

PIYACHAT PUTTARAKSA: SIMULATION AND OPTIMIZATION FOR COMPRESSOR UNIT OF CRACKED GAS IN ETHYLENE PLANT. ADVISOR: PROF. PAISAN KITTISUPAKORN, Ph.D., 82 pp.

This work is aimed at studying factors involved in polymer fouling for a cracked gas compressor in an ethylene plant by developing a HYSYS model and simulation. Wash oil and water have usually been decided to disperse or dissolve some polymers generated inside the compressor and to reduce the discharge temperature of the compressor respectively. They have also been used for an olefins plant's gas feed type in order to decrease high polymer fouling with various amount of feed rate subjecting to operating condition design. However, in practice, wash oil has less impact on compressor discharge temperature, the main factor to consider high potential polymer fouling in the cracked gas compressor, and therefore has usually been hard to determine the feed rate. On the other hand, wash water has been employed to handle the variation of the temperature. Then, the flow rate of wash water needed to handle the temperature has been calculated by the HYSYS program. To simulate the process by HYSYS, information regarding the cracked gas operating condition has been gathered to formulate the HYSYS model and then has been used to validate the model. With the accurate model, the optimal wash water flow rate is obtained with respect to its constraint of 1,200 l/hr to 100 l/hr to achieve the discharge temperature of the compressor at 90 °C. Optimization results have shown that the optimal wash water flow rates of the 1st, 2nd and 3rd stage compressors are at 465 l/hr, 380 l/hr and 505 l/hr, respectively.

Department : Chemical Engineering	Student's Signature Piyachat .P
Field of Study : Chemical Engineering	Advisor's Signature Paisan Kittisupakorn
Academic Year : 2010	

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CONTENTS

Page

Page

LIST OF TABLES

LIST OF FIGURES

Figure Page

CHAPTER I

INTRODUCTION

1.1 Important and reasons for this thesis

Ethylene Background

Ethylene is the largest volume industrially produced organic material. Current worldwide production is about 95 Mio. t/year and is projected to increase for the foreseeable future. A typical modern plant produces in excess of 800,000 t/year. Feedstock to ethylene plants ranges from light Ethane/Propane mix to heavy naphtha and vacuum gas oils. Most plants are designed with raw material flexibility in mind. Majority of ethylene produced is used in the production of polymers and ethylene derivatives such as ethylene oxide and glycol. A typical ethylene plant also makes a number of other important chemicals such as propylene, butadiene and pyrolysis gasoline. In the past years, Ethylene plants have evolved into highly integrated, highly flexible processing systems that can profitably adjust to changing raw material availability and market demands for Olefins products. Advanced process control technologies are used in Olefins plants and have greatly improved products quality, plant efficiency and resulted in quick payback of the investment. Typical process features of an ethylene process are short residence time in the furnace, high selectivity, feedstock flexibility, operational reliability and safety, easy start-up, and energy efficiency. Process analytics is a key issue for process control by online monitoring the various process streams in ethylene and propylene production. Process analytics maximizes yields and ensures product quality specifications. Various feedstocks (liquid and gaseous) are used for the production of ethylene. The principal feedstocks are naphthas, a mixture of hydrocarbons in the boiling range of 30 to 200 °C. Depending on the origin, naphtha composition and quality can vary over a wide range requiring quality control of the feed mixtures. Preferably in the US and the Middle East light feedstocks (natural gas, ethane, propane, butane) are used. Gas oils (crude oil fractions) are also gaining importance as feedstocks in some areas of the world.

Ethylene is the lightest olefin. It is a colorless and flammable with a slightly sweet smell at normal condition, i.e., ambient temperature and one atmosphere. Ethylene is also one of the most important olefinic hydrocarbons in the petrochemical industry. The importance comes from its highly reactive double bond in its chemical structure. It is produced mainly from petroleum- based feedstocks by thermal cracking in the presence of steam. With this double bond, ethylene can be involved in all kinds of reactions - addition, oxidation, polymerization, among many others - to convert to the final product or intermedial product in the petrochemical engineering industry. In addition, ethylene is also a major raw material to produce plastics, textiles, paper, solvents, dyes, food additives, pesticides and pharmaceuticals. So, the ethylene's use can be extended into the packaging, transportation, construction, surfactants, paints and coatings and other industries. Ethylene may be polymerized directly to produce polyethylene, the world's most widely used plastic. Ethylene can also be chlorinated to produce 1,2-dichloroethane, a precursor to the plastic polyvinyl chloride, or combined with benzene to produce ethylbenzene, which is used in the manufacture of polystyrene, another important plastic. Smaller amounts of ethylene are oxidized to produce chemicals including ethylene oxide, ethanol, and polyvinyl acetate. Ethylene quality depends on users requirements in downstream processes. No single chemical grade ethylene exists, but ethylene content normally exceeds 99.7%. Sulfur, oxygen, acetylene, hydrogen, carbon monoxide and carbon dioxide are the most troublesome impurities that must be controlled carefully. Although ethylene is considered to be the major product from an olefins plant, the by-products are also of great importance when considering the plant-wide economics. In the mid-1950s, the by-products of propylene and C4's were generally burned with the residue gas as fuel gas and the pyrolysis gasoline was blended into a large gasoline pool without hydrotreating. In the early 1960s, people gradually realized the importance of those by-products and made profit from them. The propylene can be used to produce polypropylene, isopropanol, acrylonitrile and cumene. The pyrolysis gasoline needs hydrotreating before the blending into a gasoline pool. Currently, many plants control the propylene/ethylene weight ratio in a certain range to satisfy the demand for propylene. The applications of ethylene are numerous and ethylene derivatives are traded around the world as for examples below.

Polyethylenes of various density and melt flow account for more than 50% of world ethylene demand. The primary use of polyethylene is in film applications for packaging, carrier bags and trash liners. Other applications include injection moulding, pipe extrusion, wire and cable sheathing and insulation, as well as extrusion coating of paper and cardboard.

Ethylene oxide is a key raw material in the production of surfactants and detergents. It is also used to manufacture ethylene glycols, which are in turn used in soft drinks and food packaging and textiles, and to make ethylene oxide glycol ether solvents.

Styrene monomer is used principally in polystyrene for packaging and insulation, as well as in styrene butadiene rubber for tyres and footwear.

Linear higher olefins are used as base materials for the manufacture of detergents, plasticisers, synthetic lubricants and additives, but also as comonomers in the production of polyethylenes

Ethylene is usually transported by pipeline in gaseous form from the producing plant to the purchasing plant, although a relatively small quantity of liquefied ethylene is moved by tank truck. In the United States, Texas and Louisiana are the major ethylene producing and consuming areas and numerous pipeline networks are constructed to transport gaseous ethylene.

Ethylene production

Ethylene is a basic building block of the chemical industry, and is the link between chemical companies and petroleum refiners. An ethylene plant is often and correctly called an olefin plant, because of the fact that the end products are olefins. The ethylene or olefin plant will yield mostly ethylene, but will also break the feedstock into a number of usable byproducts. Olefin plants can be subdivided into the following major sections - Cracking and Quench, Compression and Treating, Drying and Chilling, and Product Separations.

In an olefin plant, crude oil or natural gas is thermally cracked in a reactor furnace, and in coal gasification plants the coal is gasified in a reactor under the effect of steam and oxygen. The gas discharged from the gasification reactor or olefin plant furnace is an up to 20-component mixture of hydrogen, carbon oxides, hydrocarbons and other trace elements. The function of cracked gas compressor is to increase the pressure of the mixed gas for further treatment in the downstream stages of the process. The cracked gas to be compressed in compressor contains a great number of components, often with a corrosive, erosive or polymerizing effect. These components have a bearing on the thermodynamic and mechanical behavior of the machines. Since any cracked gas compressor failure automatically means an interruption of plant operation that may require an unscheduled shutdown negatively affecting plant economics, special precautions must be taken in the design and application of these machines to ensure reliable operation even under extreme working conditions.

Simulation method

To assess the technological feasibility and obtain material and energy balances for a preliminary economic analysis, complete process simulations were performed. Despite some expected differences between a process simulation and real-life operation, process simulators are commonly used to provide reliable information on process operation, owing to their vast component libraries, comprehensive thermodynamic packages and advanced computational methods. HYSYS (HYprotech SYStem) was used to conduct the simulation. HYSYS was selected as a process simulator for both its simulation capabilities and its ability to incorporate calculations using the spreadsheet tool.

In this research, simulation by HYSYS is used to find model and verify operating condition of cracked gas compressor in ethylene plant. The results from simulation program will be planned to use for optimization that can reduce cost from real operation.

1.2 The objectives of this thesis

1.2.1. To validate the model of cracked gas compressor in ethylene plant by HYSYS.

1.2.2. To find an optimal wash water flow rate which is the major variable impacting on high polymer fouling rate of cracked gas compressor unit by HYSYS.

1.3 The scopes of this thesis

1.3.1. Gather information and operating data of cracked gas compressor unit in existing ethylene plant.

1.3.2. Validate the compressor process model by using operating data and program HYSYS.

1.3.3. Vary the major factor or wash water flow rate by HYSYS with validated model.

1.3.4. Determine the operating condition with optimal amount of wash water flow rate.

CHAPTER II

THEORY AND LITERATURE REVIEWS

2.1 Ethylene process

Ethylene plant process design and proprietary technology is offered by the following U.S. engineering firms/design constructors: ABB Lummus Global, The M.W. Kellogg Company, Stone & Webster, and Brown & Root/Braun, each firm offers proprietary furnace technology.

Ethylene plants are originally designed to crack a particular feedstock - LPG (ethane, propane, butane) or liquid feed (naphthas, condensates, gas oils). In general, there is little fundamental difference in the process design of the two varieties of ethylene plants. The major difference is that since liquid crackers produce a large volume of heavy pyrolysis liquids, there are more pieces of equipment used to remove and purify these liquids. Olefin plants can be subdivided into the following major sections - Cracking and Quench, Compression and Treating, Drying and Chilling, and Product Separations.

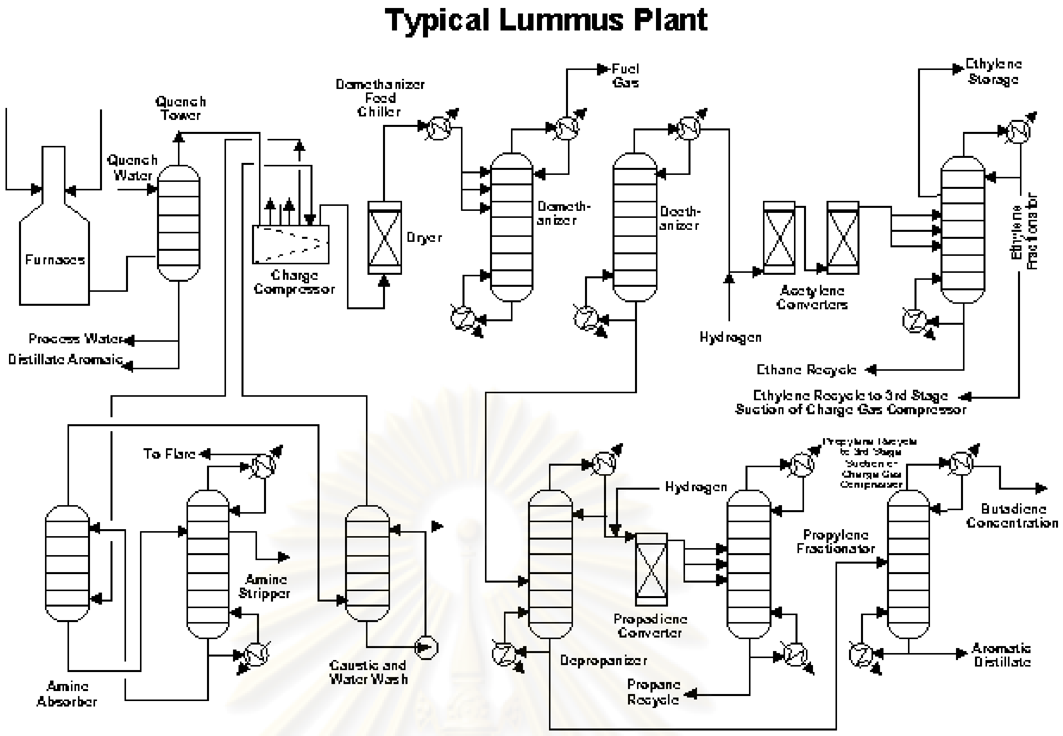


Figure 2.1 Ethylene plant diagram from Lummus Technology

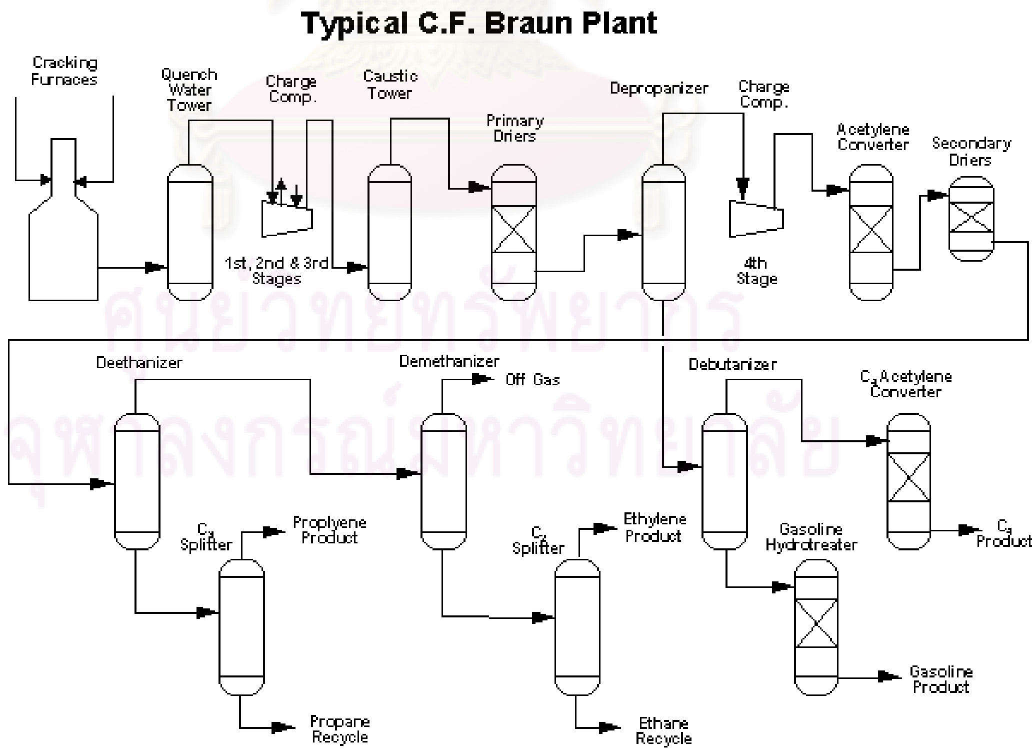


Figure 2.2 Ethylene plant diagram from Brown & Root/Braun Technology

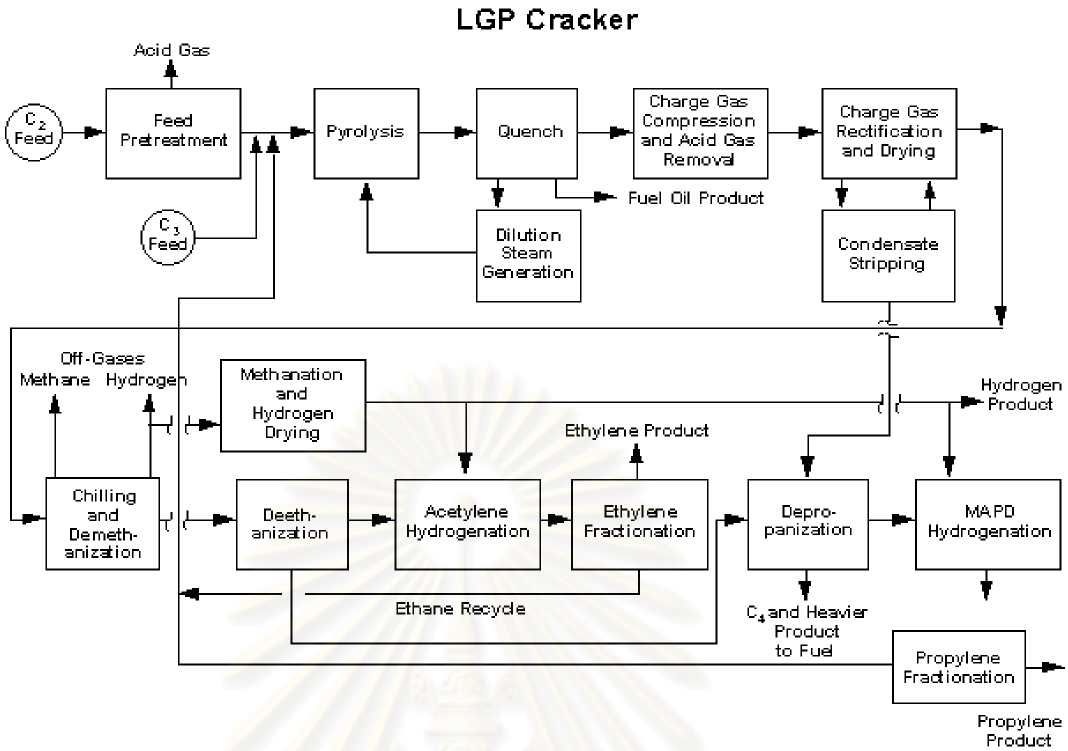


Figure 2.3 Block diagram of gas cracker or LPG cracker

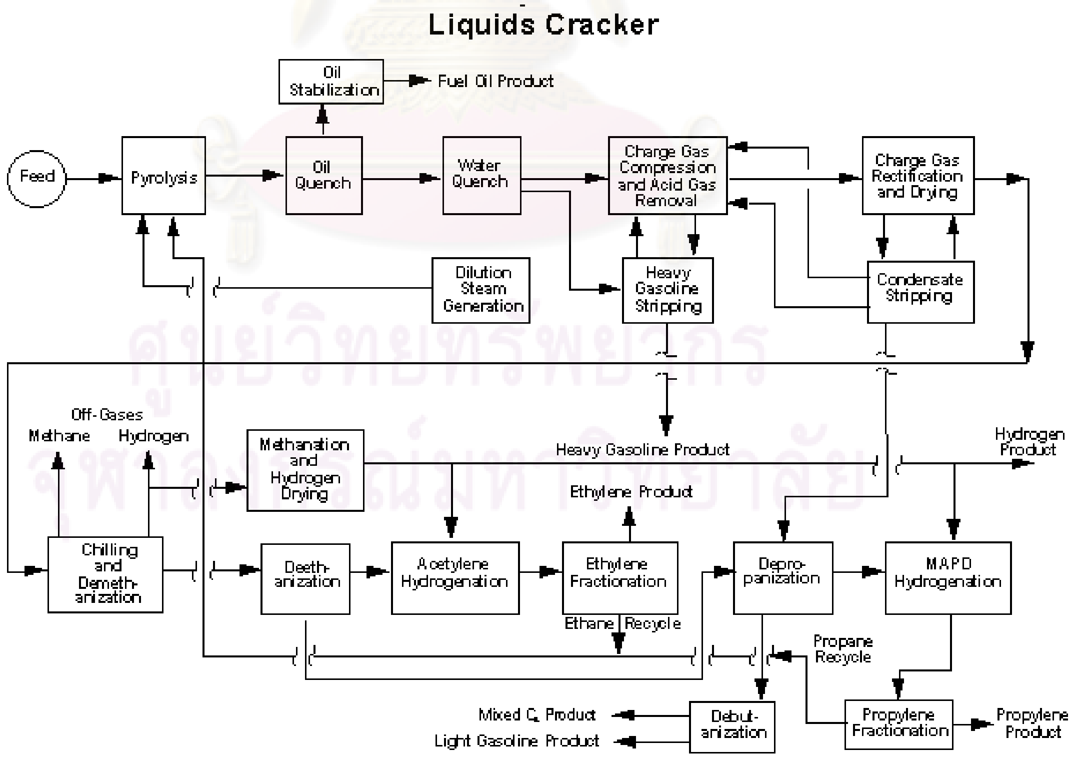


Figure 2.4 Block diagram of Liquid cracker

2.1.1 Cracking and Quench

The cracking or pyrolysis section of an ethylene plant contains between 4 and 20 furnaces with a normal average of 8 - 12 furnaces per plant. The furnace feedstocks are preheated externally and vaporized in the convection section. The steam-hydrocarbon mixture then enters the pyrolysis coils where it is further heated and cracked into products. For the pyrolysis reactions, hydrocarbon partial pressure and temperature, as functions of coil length and time, are of particular importance. The total residence time the gas spends in the pyrolysis coil of a modern furnace is generally in the range of 0.2 - 0.6 seconds. The gas outlet temperature is in the range of 1450 - 1550° F depending upon feedstock and furnace design. The product gas is sent to closely connected transfer line exchangers in which the temperature is rapidly reduced by several hundred degrees by generating high pressure saturated steam at approximately 1500 psig. Given the extreme process gas temperature and the high steam pressure, the transfer line exchangers have to be of very special design. Two or more transfer line exchangers, together with a common steam drum, are usually built right into the furnace structure.

2.1.2 Water Quench and Dilution Steam Generator

In the base of the tower, hot quench water is separated from heavy pyrogasoline and tars which is condensed along with the dilution steam. The bulk of the hydrocarbons is pumped to the primary fractionator to serve as reflux, while the net product is sent to the heavy pyrogasoline stripper for the removal of C4 and lighter hydrocarbons. The net dilution steam condensed is withdrawn from the quench tower and, after a second more effective water-hydrocarbon phase separation, it is sent to the process water stripper for the removal of dissolved hydrocarbons and gases. The treated process water is then pumped to the dilution steam generators which are heated with quench oil and medium-pressure steam. The net process water discharge from the system is minimal since only the small quantity of the live stripping steam fed to the fuel and gas oil stripper represents a net inflow. In an LPG based cracking plant, a water quench tower alone is commonly used to cool the cracked gas immediately downstream of the TLEs. In this case, very little heavy oil and pyrogasoline is produced by LPG cracking. The hot quench water is used for low level heat recovery while the net condensed dilution steam is reused as dilution steam boiler feedwater.

2.1.3 Compression and Treating

The pyrolysis gas leaves the water quench tower at approximately 80 - 100°F and at a pressure slightly above atmospheric. Most processes call for compression to a pressure of 525 - 550 psig which appears optimal for the subsequent cryogenic treatment. Modern olefin plants employ radial centrifugal compressors to boost operating pressure. Plants based on gaseous (LPG) feedstocks commonly employ four stages of compression, while naphtha or gas oil plants employ five stages of compression. Five stages will reduce compressor discharge temperature which, in turn, reduces the fouling rate of the machine. The typical pyrolysis gas compressor consists of two or three compressor casings driven by a single extraction/condensing steam turbine. High pressure superheated steam from the cracking furnace waste heat boilers is used to drive the turbine. Some older ethylene plants operate electric driven process gas compressors. Water and hydrocarbons condensed between stages are separated from the pyrolysis gas in interstage separators (also called suction drums). The water is returned to the quench water system while the hydrocarbons formed in the first three stages are sent to the heavy pyrogasoline stripper, while the condensate from the last one or two stages is sent to the condensate deethanizer (condensate stripper). "Treating" refers to the removal of contaminants (acid gas and water) prior to separation and purification of the olefins. Hydrogen sulfide and carbon dioxide are commonly removed between the third and fourth stages of compression. This location is ideal because the actual gas volume has been reduced significantly in the first three stages of compression while the acidic components are still present in the gas stream and have not contaminated any products that are separated ahead. In those plants, processing feedstocks with a sulfur content of a few hundred ppm, scrubbing with a dilute caustic soda solution has proven most economical. A three-stage scrubbing tower is provided to obtain maximum utilization of the caustic soda. The solution circulating over the middle section of the scrubbing tower typically contains between 8 and 12% free sodium hydroxide while the solution in the lower section will contain between 2% and 5% free caustic. The top section uses fresh water to minimize caustic entrainment in the effluent gas. Relatively weak caustic solutions are preferred to avoid the precipitation of sodium salts and to minimize the formation of organic sodium complexes. The pyrolysis gas leaving the caustic scrubber contains less then 1 ppm of acid gases and thus assures that the final products of the plant will meet specifications in this respect. The spent caustic represents, environmentally, the most troublesome liquid effluent of ethylene plants and requires substantial treatment before it can be safely discharged. Plants designed to process feedstocks with significant sulfur contents (i.e., in excess of 500 ppm approximately) often contain a regenerative acid gas removal system upstream of the caustic scrubber. Regenerative systems alone cannot achieve the essentially complete removal of acid gases from the pyrolysis gas as required. These systems, which may employ monoethanolamine or diethanolamine as solvents, would be of the standard absorber-regenerator design if it were not for the present of diolefin compounds in the pyrolysis gas which tend to dissolve in, or be weakly bound to the solution and through polymerization, may cause severe operational problems in the regenerator. These systems must contain facilities for the removal of hydrocarbons from the rich solution. In fact, it was the development of these solution purification techniques that, after unsatisfactory early attempts, made possible the application of regenerative acid gas removal systems for the bulk removal of hydrogen sulfide and carbon dioxide from pyrolysis gas.

2.1.4 Charge Gas Drying

Complete removal of water from the pyrolysis gas is required in preparation for the cryogenic treatment. Most processes employ a single adsorptive drying system located immediately after the final stage of pyrolysis gas compression. Molecular sieves are the preferred desiccants because of their selectivity. Of two or more dryers normally provided, one or more in normal operation while the other is regenerated, recooled, or held in a standby position. High pressure methane heated with steam or in a direct-fired heater to a temperature of approximately 450°F is the preferred regeneration medium. The dryers are designed for onstream times of 24 to 48 hours between successive regenerations. To limit the quantity of desiccant required and thus the size of the vessels, the pyrolysis gas, upstream of the dryers, is cooled with propylene refrigeration to above the hydrate inception temperature, the value of which depends on the system pressure.

2.1.5 Charge Gas Chilling

The cryogenic treatment involves the successive chilling of the pyrolysis gas in heat exchange with boiling refrigerant and with cold process streams that have to be vaporized and/or reheated. As the pyrolysis gas temperature is reduced, more and more condensate is formed. There exist numerous process variations on the sequence of condensate separation, fractionation, and heat exchange. The vast majority (95%) of the hydrogen in the cracked gas is separated and removed in the cryogenic section. Brazed aluminum plate-fin exchangers are generally used for the multipass cryogenic heat transfer services and are sometimes even used for the refrigerated chillers with thermosyphon circuits for refrigerant circulation. Much of the cryogenic equipment is often installed in a rectangular carbon steel container, commonly called cold box, with its void spaces filled with perlite or rockwool for insulation. Considering the complexity of the system, this mechanical design is characterized by surprising compactness and minimum space requirements.

2.1.6 Product Separations

The product separations area of an olefins plant consists of a number of distillation columns and reactors. They are used to purify and separate the cracked gas into salable polymer grade ethylene and propylene. In most ethylene units, the following equipment is included in the product purification section: C1 -C4 columns and reboilers, acetylene and propadiene converters, a C2 splitter, and a C3 splitter. In the demethanizer, methane and remaining hydrogen product is taken overhead. The recovered methane becomes fuel for the cracking furnace operation. Hydrogen is recycled back to the cold box where it is further purified and used in the acetylene and propadiene converters. There is usually no process fouling problems in the demethanizer due to its very low operations temperatures. The deethanizer splits the ethylene and ethane overhead. The C2 stream is sent through an acetylene converter where residual acetylene is converted to ethylene and ethane. Acetylene converter operation is important since the ethylene product must meet tight specifications on acetylene content. In the C2 splitter, polymer grade ethylene is taken overhead and the bottoms product ethane is usually recycled back to a cracking furnace. The deethanizer column is usually the first area where polymer fouling may be noted. Typically, column fouling is minor, but reboiler fouling can be quite severe. The depropanizer column splits C3s overhead. The overhead product is then sent through the methylacetylene/propadiene converter. After treatment, the C3 splitter separates propylene overhead from a propane bottoms product. Propane is either used for fuel or is cycled back to a cracking furnace. The depropanizer is usually the first column where severe tray and reboiler fouling becomes evident. Some of the more modern plants have installed a separate depropranizer column bottoms section in anticipation of severe fouling due to polybutadiene. Peroxides, metals, and heat initiate polymer reactions. In the debutanizer, the mixed C4s (butadiene, butene and butanes) are taken overhead. The overhead product becomes feed to a butadiene extraction/recovery plant. The bottoms product is raw pyrolysis gasoline, sometimes also called DAC (Debutanized

Aromatic Concentrate). Some problems have been noted with polymerization in the debutanizer overhead condensers, column and reboilers.

2.2 Compressor Types

Compressor is device for increasing the pressure of a gas by mechanically decreasing its volume. It is key component in many manufacturing and process industries. The major compressor classes are positive displacement and dynamic.

An example of the positive displacement class is the bicycle pump or fireplace bellows, both of which change the volume of a chamber to compress air. If a piston inside a cylinder forms the chamber, the compressor is known as a reciprocating type. These are further subdivided into single-acting and double-acting. In a single-acting type, only one piston face compresses the air; double-acting types use both faces alternately. Reciprocating compressor sizes range from fractional horsepower to more than 600 hp.

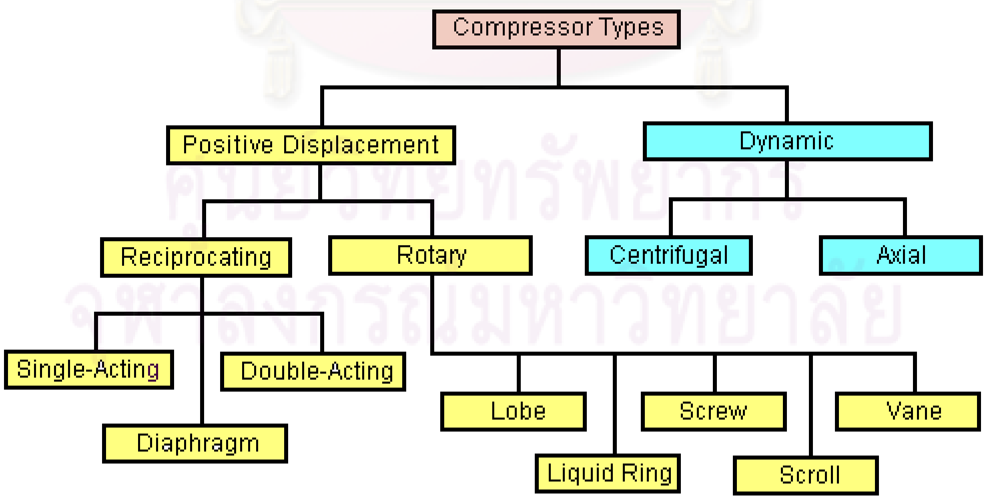


Figure 2.5 The main types of gas compressors

Another type is the rotary positive displacement compressor, in either helical screw or sliding vane varieties. The helical screw compresses air between a meshing rotating rotor and screw assembly. Helical screw compressors are available in sizes from about 3 hp to several thousand horsepower. The sliding vane compressor uses a set of sliding vanes placed in slots on a rotor eccentrically mounted in a cylindrical casing. As the rotor spins, centrifugal force presses the vanes against the casing wall to compress air between the vanes and the casing.

The second major compressor class is the dynamic type, which compress by converting air velocity into air pressure using blades mounted on a rotating shaft. In centrifugal compressors, air enters near the base of the impeller blades, accelerates along the blade and exits near the ends of the blades at the circumference of the compressor case. Centrifugal compressors range in size from about 100 hp to several thousand horsepower. In an axial flow compressor, the air enters and exits along the axis of the shaft, usually after passing through several stages of rotor blades. Each set of rotating blades is separated from the next by nonrotating stator blades. Air compressors in aircraft jet engines are a common example of the axial type. Axial-flow compressors are available in sizes from a few hundred horsepower to several thousand horsepower.

Both positive-displacement and dynamic compressors can be single-stage or multistage. Multiple stage compressors need two or more stages to reach the final output pressure; the output of one stage being the input to the next. Cooling the air between stages improves compressor efficiency.

Each compressor type reciprocating, screw, rotary vane, centrifugal and axial has typical operating characteristics. There is, however, overlap and, for a given application, one might have a choice of types. Some important characteristics are flow, pressure, capacity control and lubrication.

2.2.1 Centrifugal compressor

Centrifugal compressor is an integral part of the petrochemical industry, finding extensive use because of their smooth operation, large tolerance of process fluctuations, and their higher reliability compared to other types of compressors. Centrifugal compressors range in size from pressure ratios of 1:3 per stage to as high as 12:1 on experimental models. The proper selection of a compressor is a complex and important decision. The successful operation of many plants depends on smooth and efficient compressor operations. To ensure the best selection and proper maintenance of a centrifugal compressor, the engineer must have a knowledge of many engineering disciplines.

Centrifugal compressors are fluid flow dynamic machines for the compression of gases according to the principals of dynamics. The bladed impeller with its continual internal flow serves as an element of energy transfer to the gas. Pressure temperature and velocity of the gas leaving the impeller are higher than at the inlet. Diaphragms or diffusers arranged after the impeller helps in diverting the gas velocity, thus further increase in pressure and temperature is achieved by the conversion of the kinetic energy into pressure energy. During energy transfer in the impeller, the gas flows from the inside in an outward direction. It is therefore subjected to the change of the centrifugal field, through which the attainable pressure ratios are substantially higher than those of axial compressors. The radial direction of flow in the impellers again requires radially arranged diffusers, which increases the outer diameters of the casing to about double the impeller diameters. Generally centrifugal compressor are used for high capacity and low pressure and though initial cost is high but, lower maintenance and running cost places these compressors to compare with high efficiency reciprocating compressor.

Centrifugal compressors are generally manufactured in two configurations:

1) Horizontally split Construction
2) Barrel type construction

Compressors of both the configurations are in use and the discharge pressure of the gas directs the designs. Centrifugal compressors are also manufactured in integral gear type construction. Centrifugal compressors are composed of outer casing which contains stator part called a diaphragm bundle, a rotor formed by a shaft with one or more impellers, a balance drum, thrust collar etc. The rotor is driven by means of coupling hub and is held in position radially by journal bearings and axially by a thrust bearing. Sealing system in centrifugal compressors is very interesting and a bit complicated. Sealing of by-passing gas from one stage to another is required between all the inter stages and also at both the shaft ends as well. In general rotor and stators are fitted with labyrinth seal rings sealing. In toxic gas services 100% leak-proof seals are required which may be of different type depending on the service of the compressors. Oil deflectors and oil labyrinths are used as oil seal on both ends of the rotor.

Gas is drawn into the compressor through suction nozzle of the compressor and enters in to an annular chamber called inlet volute, flowing towards the center of the impeller from all directions in a uniform radial pattern. The rotating impeller imparts energy through its wanes and pushes the fluid outwards raising its velocity (kinetic energy) and pressure as it passes through the impeller shroud. The outlet fluid leaves the impeller tangentially and then enters into another circular chamber called diffuser, where its velocity (kinetic energy) is converted into pressure energy. After this increase in pressure of the fluid in one stage again it enters into second stage impeller eye and cycle goes on till the final discharge of the pressurized gas from the compressor discharge nozzle.

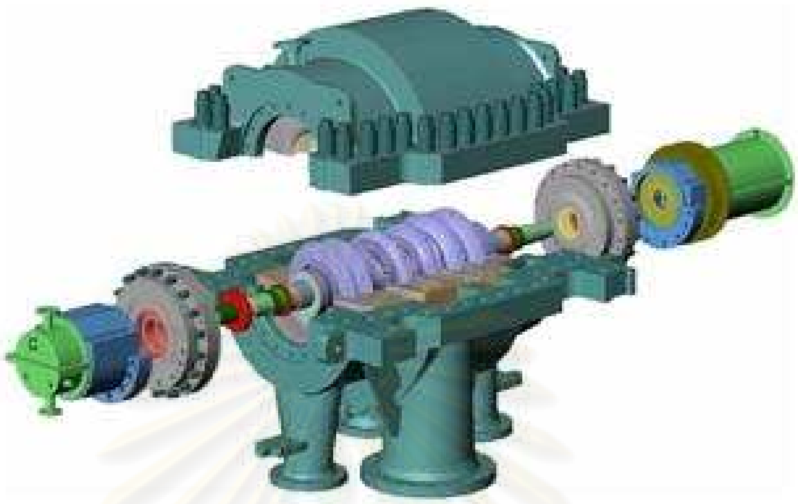


Figure 2.6 Horizontally split Construction

Horizontally split casing consists of two halves joined along the center line by casing bolts. These type of compressors are generally made up to 40 kg/cm^2 discharge pressure. All the suction nozzle, discharge nozzle, lubricating lines and other connections are generally located in the lower half casing so that during maintenance / inspection, only the upper half casing can be removed easily and gain access to all internal components. The material of construction of casing depends upon the operating parameters like pressure, temperature, gas handled etc.

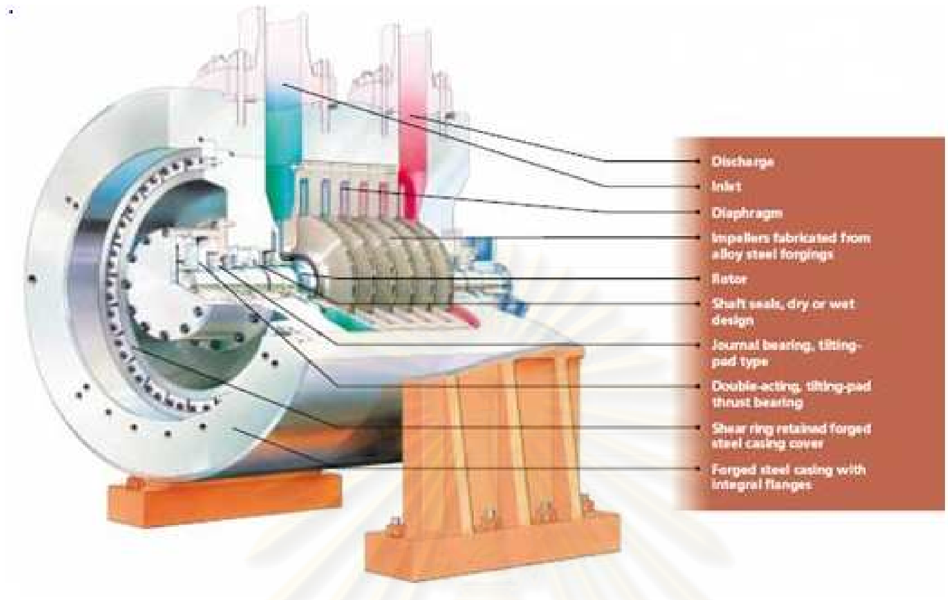


Figure 2.7 Barrel type construction

2.3 Fouling in charge gas compressor or cracked gas compressor

2.3.1 Causes of Fouling

The furnace feedstock being cracked relate with fouling potential in the compressor area. In NGL crackers, the fouling rate can be classified as moderate to high, depending on discharge temperatures. In general, compressors receiving feed from naphtha and gas oil crackers exhibit a much lower fouling potential. This is believed to be due to the large quantities of aromatics contained in the raw cracked gas itself. These aromatic compounds provide a solubilizing and washing action of polymers generated by the heat of compression. In contrast, the cracked gas originating from an NGL feedstock contains only trace quantities of aromatics; thus, supplemental wash oil injection becomes critical to minimize polymer fouling. Plants cracking large quantities of ethane exhibit the highest fouling potential.

2.3.2 Fouling mechanisms

Fouling in process gas compressors is common in most ethylene plants regardless of the feedstock. Deposits are formed by polymerization of diolefins, mainly butadiene and isoprene, present in the cracked gas. The polymerization reactions follow a radical chain mechanism initiated, in most cases, by thermal homolysis of impurities present in the monomer. These impurities include peroxides or hydroperoxides formed by oxygen reaction with the feedstock (for example during storage). A general free radical polymerization mechanism is shown in Figure 2.8 it consists of a sequence of three steps: initiation, propagation, and termination.

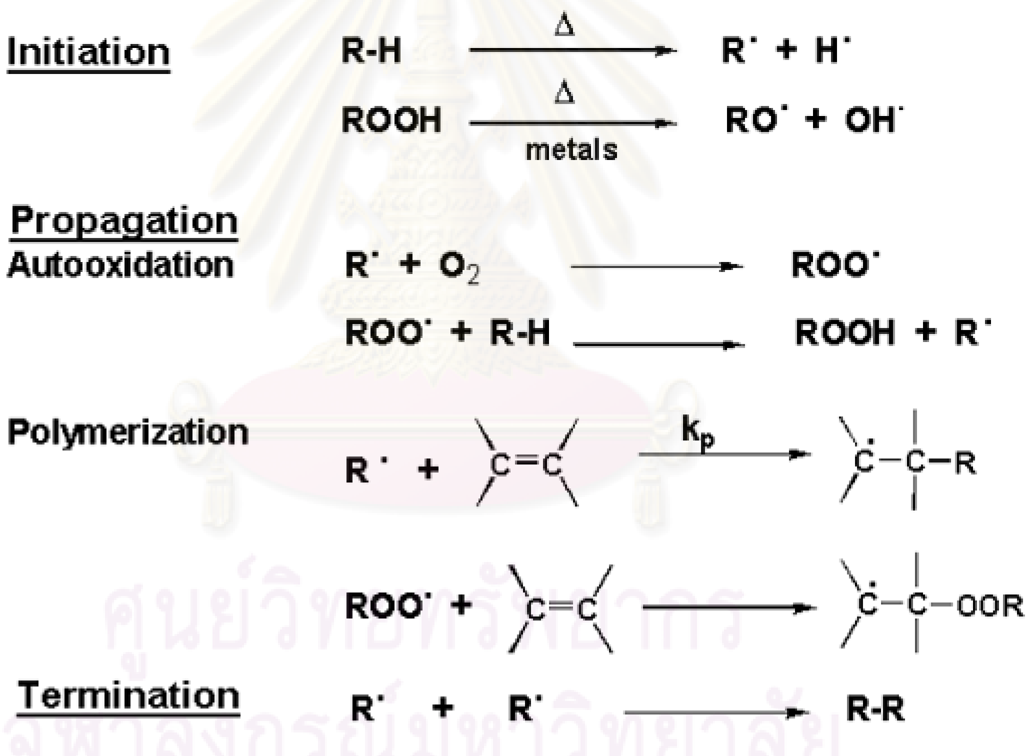


Figure 2.8 Free radical Polymerization mechanisms

In the initiation step, a free radical is generated by heat and/or metals from impurities present in the olefinic streams. These radicals, called initiators, can further react in two ways depending on whether oxygen is present in the medium. The initiation step does not involve the thermal production of radicals directly from ethylene because the dissociation energy in this case is too high (above 170 kJ/mole). In the presence of oxygen, the initiator will rapidly react with it to form peroxy radicals which at elevated temperatures will add to the vinyl monomer, a diolefin in our case, and generate a larger free radical species. The propagation with oxygen is frequently called autooxidation. In the absence of oxygen, the initiators add to a monomer molecule to form a new free radical. The process is repeated hundreds of times and monomer molecules are successively added, with a rate constant kp, to the propagating chain which continues to grow in molecular weight to form a polymer. The discharge temperature of compressors has a direct relationship to fouling potential. When the temperature is high, the potential fouling rate is also high as shown in Figure 2.9.

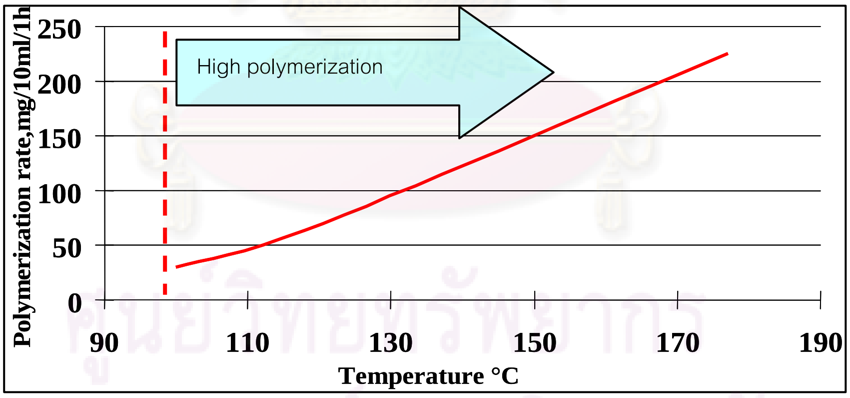


Figure 2.9 Temperature impact on polymerization rate

2.3.3 Wash oil injection

Fouling potential in charge gas compressors is highest when NGL feedstocks are cracked. Another factor that has a direct relationship to fouling potential is the compressor discharge temperature. Modern ethylene units are designed with 5 stages of compression rather than 4 stages used under older plant designs. The additional stage reduces the compression ratio per stage, which in turn reduces the discharge temperature. This minimizes compressor fouling. Wash oils are sprayed into the suction line of compressors or into individual impellers to wet compressor internals and wash away polymer forming in the machine. Typical wash oil injection rates range from 0.3 - 1.0 GPM/stage. The wash oil used may be an internally recovered aromatic oil(dripolene) or purchased oil (FCC light cycle oil). The characteristics of a good quality wash oil are as follows:

2.3.3.1 High Aromatic Content - Total aromatics should be between 15% - 95%. Aromatics are a good polymer solubilizing agent. High molecular weight aromatics are preferred.

2.3.3.2 High Boiling Point - The preferred boiling point range is 350°F - 600°F. The wash oil must remain liquid once inside the machine to assure proper wetting of compressor internals. Low boiling range oils will flash or vaporize in the first and second compressor stages reducing their solubilizing action.

2.3.3.3 Low Diolefin Levels - Some wash oils recovered from other sections of the ethylene plant may have high diolefin and peroxide concentrations which accelerate compressor fouling. In cases like this, an antioxidant should be added to this wash oil to minimize peroxide and gum formation until the oil is injected into the compressors.

2.3.3.4 Low Existent Gum - High gums will promote compressor fouling. Existent gums should be less than 25 mg/100 ml.

2.3.4 Wash water injection

Some ethylene producers inject water into the process gas compressor along with aromatic wash oil injection. Water functions to reduce compressor fouling tendency by reducing the discharge temperature. If water is used, it must be deaerated boiler feedwater with nil levels of dissolved solids and oxygen. Another good water source is steam condensate. Because excessive water injection can cause erosion of compressor internals, a water injection rate guideline is 1% water based on total gas throughput in pounds per hour. For example, a compressor processing 250,000 lbs/hr cracked gas should inject 5 GPM water. Water is sometimes intermittently injected to remove caustic salts due to caustic tower foaming/carryover. Water injection will add significantly to the amount of liquid in the downstream knockout (or suction) drum. Extra water will require proper level controllers and facilities to purge water continuously from the knockout drum. It has been well known that the wash water is employed to reduce compressor fouling tendency by reducing the discharge temperature; the discharge temperature of compressors has a direct relationship to fouling potential. When the temperature is high, the potential fouling rate is also high.

2.4 Simulation and optimization

Nowadays, many production plants use comprehensive thermodynamic packages and advanced computational to conduct simulation to obtain material and energy balance for preliminary economic analysis. There are the methods of simulation for optimization as follow:

Alex et al. [1] showed that four continuous biodiesel processes were designed and simulated in HYSYS. The first two employed traditional homogeneous alkali and acid catalysts. The third and fourth processes used a heterogeneous acid catalyst and a supercritical method to convert a waste vegetable oil feedstock into biodiesel. While all four processes were capable of producing biodiesel at high purity, the heterogeneous and supercritical processes were the least complex and had the smallest number of unit operations. Material and energy flows, as well as sized unit operation blocks, were used to conduct an economic assessment of each process. Total capital investment, total manufacturing cost and after tax rate-of-return were calculated for each process. The heterogeneous acid catalyst process had the lowest total capital investment and manufacturing costs, and had the only positive after tax rate-of-return.

Cardoso et al. [3] presented a simulated annealing-based algorithm (MSIMPSA) suitable for the optimization of mixed integer non-linear programming (MINLP) problems was applied to the synthesis of a non-equilibrium reactive distillation column. A simulation model based on an extension of conventional distillation is proposed for the simulation step of the optimization problem. In the case of ideal vapor-liquid equilibrium, the simulation results are similar to those using the GAMS environment and to those obtained with the AspenPlus modular simulator. The results show that the optimized objective function values are very similar, and mostly independent of the number of trays and of the reaction distribution.

Gao et all. [4] presented Braun K10 equation as the thermodynamic method, ethylene quench system was simulated using software PRO/II. Instead of pseudo-components, the pyrolysis oil was characterized by real components. Real components were selected from paraffin, naphthalene and aromatic groups (PNA) for each cutting interval of true boiling point (TBP) curve. Compositions of PNA components in each interval were supposed to be the same with that in whole pyrolysis oil. Simulation results were consistent with measured data and better than using pseudo-components, which proved that the new approach could be an effective way to calculate quench system.

Paul et al. [5] Acetone–methanol, methyl acetate–methanol and methanol–chloroform binary extractive single column distillation systems were simulated with the HYSYS software platform, to investigate the effects of solvent feed entry stages, solvent split stream feed and solvent condition on the separation. Water was used in all of the simulations as the solvent. The simulations supported data and findings from experimental column studies of the same systems. A rigorous simulation of the acetone–methanol system including a secondary stripping column and recycle loop was established to simulate an industrially relevant situation. This simulation enabled an economic evaluation of the process to be made. It was found for feed mixtures containing 25, 50 and 75 mol% methanol, the optimum reflux ratios were found to be 3.5, 3.5 and 4.2, respectively. As a consequence one column design could separate binary feed of varying composition between 25 and 75% methanol. The optimum number of ideal stages for the primary column for an equimolar binary feed was determined to be 73. When maintaining a constant solvent flow, the distance between the split feed entry stages had no effect on the economic potential (EP) of the system.

Aspelund et al. [6] developed a gradient free optimization-simulation method for processes modelled with the simulator Aspen HYSYS. The tool is based on a Tabu Search (TS) and the Nelder-Mead Downhill Simplex (NMDS) method. The local optima that result from the TS are fine-tuned with NMDS to reduce the required number of simulations. The tool has been applied to find the total refrigerant flow rate, composition and the refrigerant suction and condenser pressures that minimize the energy requirements of a Prico process. The main strength of this method is that it has a high probability of obtaining a better solution with significantly fewer simulation runs than other metaheuristic methods. Also, by changing the TS step size it is possible to influence the initial search pattern, thereby taking advantage of already gained process knowledge to decrease the optimization time. The method is general and can be applied to other processes modelled in Aspen HYSYS.

Manuel et al. [7] simulated a continuous process for biodiesel production by using Aspen HYSYS V7.0 software. As fresh feed, feedstocks with a mild acid content have been used. The process flowsheet follows a traditional alkaline transesterification scheme constituted by esterification, transesterification and purification stages. Kinetic models taking into account the concentration of the different species have been employed in order to simulate the behavior of the CSTR reactors and the product distribution within the process. The comparison between experimental data found in literature and the predicted normalized properties, has been discussed. Additionally, a comparison between different thermodynamic packages has been performed. NRTL activity model has been selected as the most reliable of them. The combination of these models allows the prediction of 13 out of 25 parameters included in standard EN-14214:2003, and confers simulators a great value as predictive as well as optimization tool.

Soos et al. [8] presented the design of a separation column for recovery of 1,2 dichloropropane (DCP) from the off-gas released during propylenechlorhydrine (PCH) synthesis. The aim of this separation is to recover DCP with a purity of 99.99 mass%. The simulation of the separation column is performed by means of the professional simulation program HYSYS. The design of the rectification column is based on complete experimental measurements. The calculation is compared with the design of a separation column, which uses for the computation of the equilibrium composition the corresponding database of the simulation program, and with the calculation, in which the complete set of equilibrium data is predicted from the UNIFAC method. By comparison of the results of the rectification column design quite different values of parameters which characterising the rectification column (i.e. reflux ratio, flows and composition of phases on the stages. . .) were obtained. These could lead to false results in designing the equipment. Within the frame of this work experimental vapour-liquid equilibrium (VLE) data at a pressure of 98.66 kPa and binary solubilities are presented for the system composed of following components: water, propylene oxide (PO), propanal (PA), and DCP. The dependence of activity coefficients on the composition is given by the UNIQUAC equation. The experimental results are compared with those predicted by means of the HYSYS database and with those predicted from the UNIFAC method.

Petter et al. [9] A real time dynamic simulator of the Hydro Rafnes VCM plant has been developed. The simulator model covers the cracking and quench section of the plant, as well as the HC1 column and refrigeration unit. A high fidelity emulation of the distributed control system, Honeywell TDC 3000, is an essential part of the simulator. The simulator is mainly used for operator training, but has also been very useful during tests of the control strategy of a new refrigeration compressor. New development for this specific simulator project includes development of a cracker furnace model and thermodynamic routines for "On-line K-value" computation and robust flash routines for mixtures with one dominant component and small amounts of impurities.

CHAPTER III

Methodology

3.1 Research Methodology

This study is based on the information from real ethylene plant. Thus the compressor simulation is set by using design operating data and actual data to formulate the HYSYS model and then is used to validate the model. With the accurate model, the optimal wash water flow rate is determined. The methodology diagram explaining the process of this research is shown in Figure 3.1.

This research is divided into three major parts:

(i) Ethylene plant description and operating data.

(ii) Validation model.

(iii) Wash water optimization for this thesis.

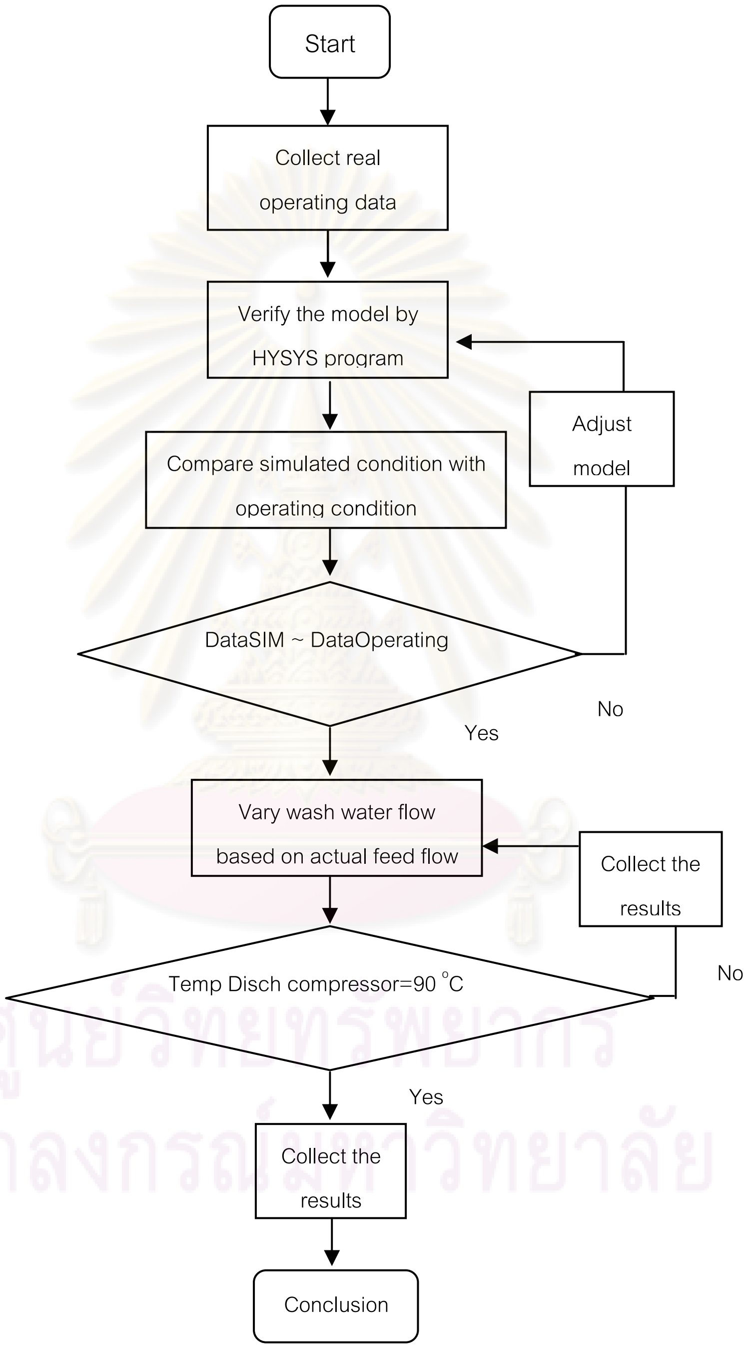


Figure 3.1 Flow diagram of research methodology

3.1.1 Ethylene plant description and operating data

The studied plant is the large ethylene plant in Thailand which is designed to crack LPG feedstock. This plant has production process like general ethylene plant that has about 7 major sections and use KBR (Kellog/Brown & Root) technology as design engineering.

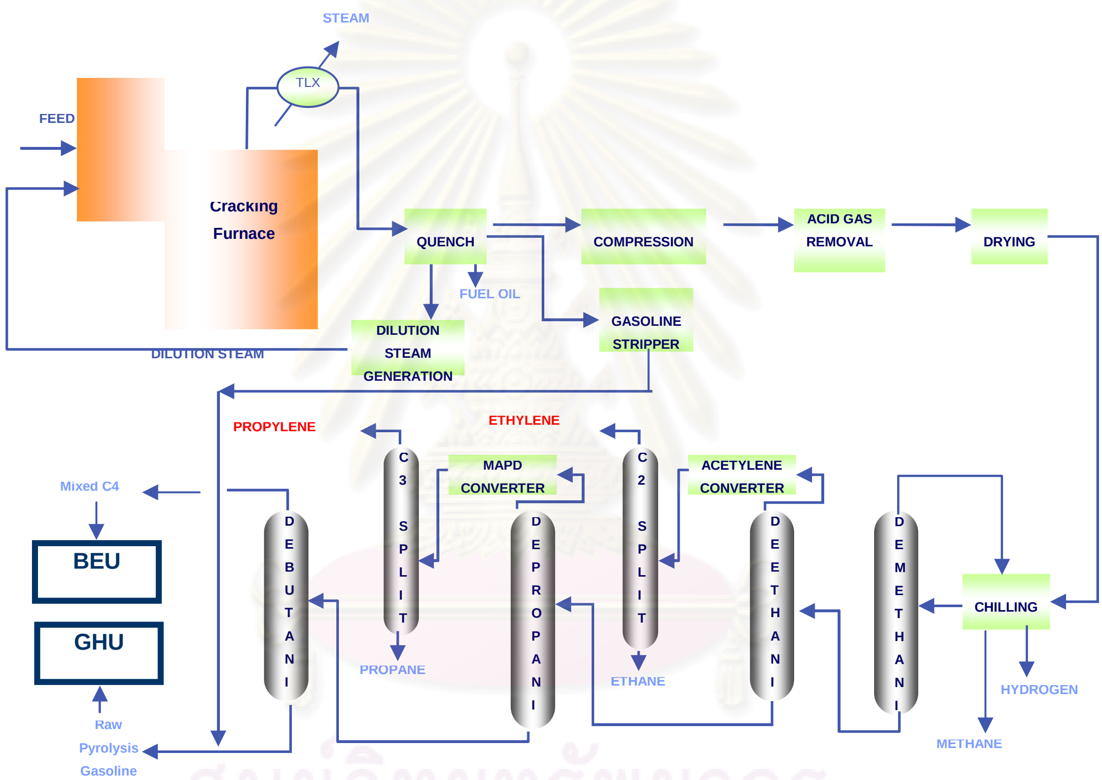


Figure 3.2 Ethylene plant diagram

3.1.1.1 Ethylene Section

This studied ethylene plant can be explained into the following 7 major sections.

1. Cracking section

This section, the feedstocks are preheated, mixed with dilution steam, and thermally cracked in the pyrolysis coils at temperatures between 1500 - 1650°F.

2. Quenching section

The cracked gases leaving cracking section are cooled by direct water quenching (water quench tower or spray gas cooler)

3. Compression section

Compressor has a role on compression or boosting pressure to liquefy light components (hydrogen & methane) at available lowest temperature. This section is the most important for this research that has more detail in next chapter.

4. Acid gas removal section

This section refers to the removal of contaminants (acid gas and water) prior to separation and purification of the olefins.

5. Drying section

Drying section can completely remove water from cracked gas that is required in preparation for the cryogenic treatment.

6. Chilling section

The cracked gas or light component from drying section will be condensed before next separation.

7. Purification section

There is used to purify and separate the cracked gas into salable polymer grade ethylene and propylene.

3.1.1.2 Compressor unit

The existing ethylene plant we research has 4 stages of compressor, however, the 4th stage of compressor is neglected because it operate with low temperature and the inspection result after plant turnaround shows us in low polymer fouling. So plant considers to remove wash water for this stage, thus we find the simulation model for the 1st, the 2nd and the 3rd stages of compressors. A diagram of cracked gas compressor unit for this plant is shown in Figure 3.3.

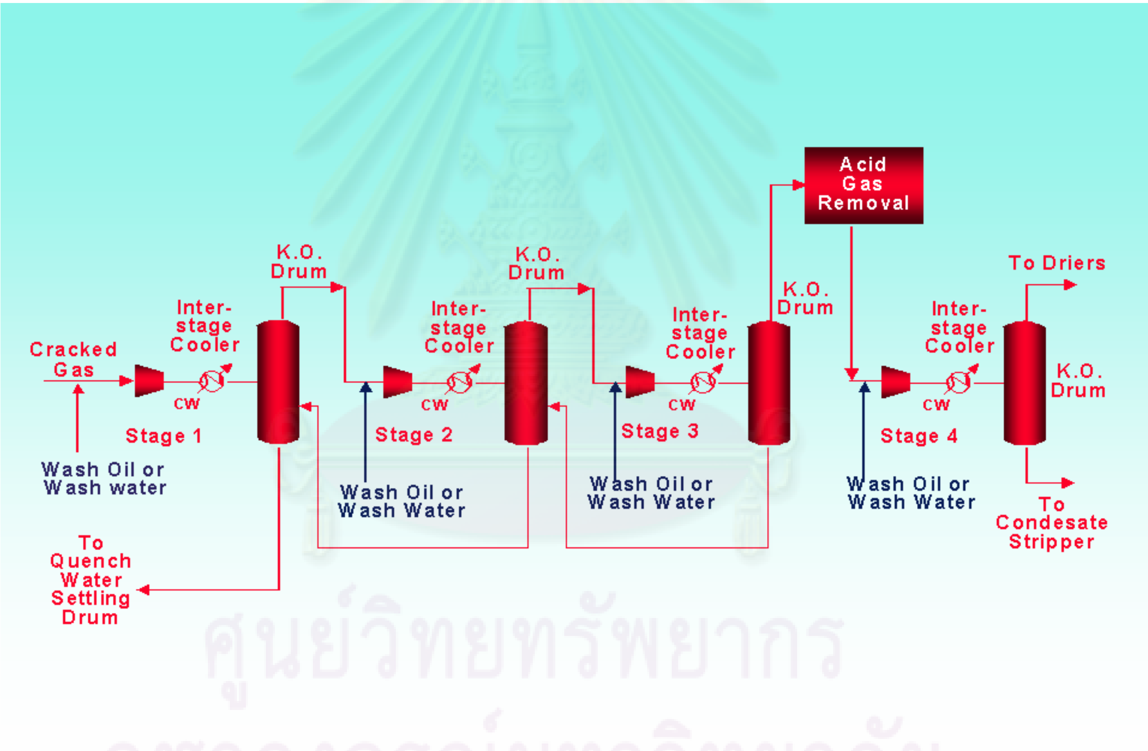


Figure 3.3 Compressor unit diagram for existing plant

For one stage of compressor, cracked gas from cracking comes and is compressed to liquefy light component. Then cracked gas with higher pressure will be condensed by heat exchanger which has a role as an inter-stage cooler before pass to knock out drum. Knock out drum allow total gas-liquid to set inside, gas phase will go through the top of drum to the next stage of compressor while liquid phase go through at the bottom of drum and back to quench system.

Real image of compressor look like Figure 3.4. The compressor must be strong and can support total cracked gas volume for production per day. The major complements of compressor are casing (inner and outer) and impeller that channel and impeller of compressor can be fouled with polymer. Fouling in cracked gas compressor can negatively affect plant economics from unplanned schedule plant shutdown.

Figure 3.4 Cracked gas compressor

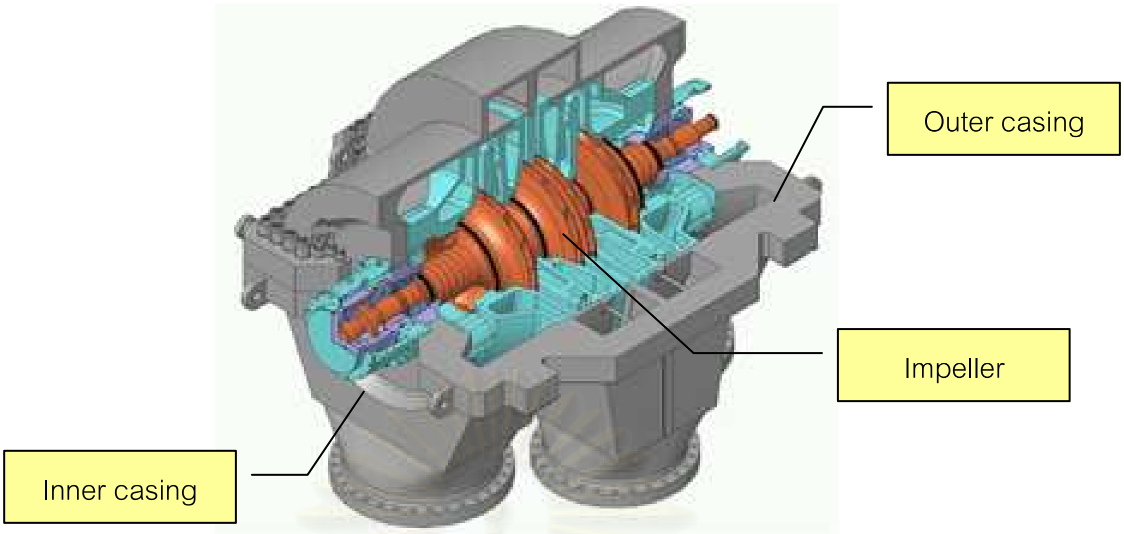


Figure 3.5 Major complements of cracked gas compressor

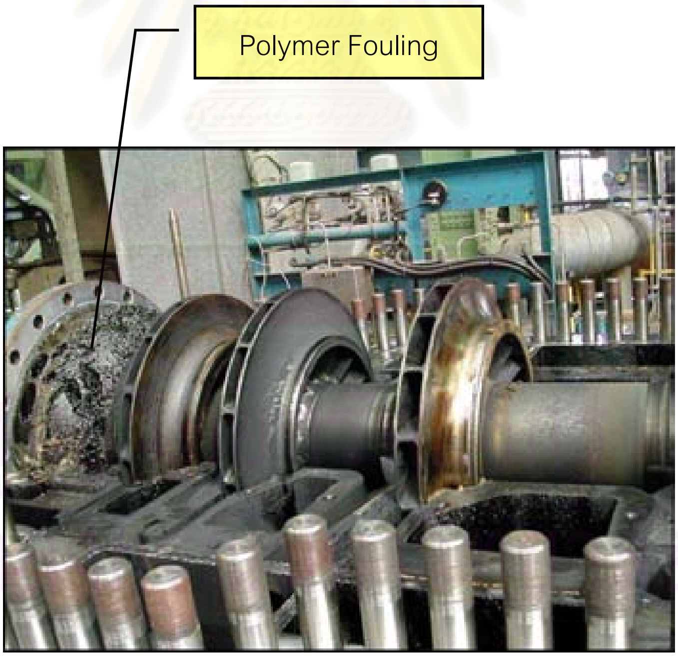


Figure 3.6 Polymer fouling in compressor

3.1.1.3 Plant information

This study is based on the information gathering from an existing ethylene plant. Cracked gas from cracking has many components that the most of them are hydrocarbon. Design operating data is highly important because it must use as input-output in simulation program to validate model before optimization. Besides the design operating data, the actual operating data is also important and use to recheck the reasonable of model. The design operating data and actual operating data are received as follow.

The design operating data

- Material balance of

COMPOSITION
Hydrogen
Carbon monoxide
Carbon dioxide
Hydrogen sulfide
Methane
Acetylene
Ethylene
Ethane
Propylene
Propadiene
propylene
propane
Vacetylene
1-Butyne

COMPOSITION
1, 2-Butadiene
1, 3 –Butadiene
1Butene
Cis 2 Butene
Trans 2 Butene
IsoButadiene
IsoButane
Butane
1,3 Cyclopentane
Cyclopentane
Trans 1, 3 Pentadiene
Isoprene
1Pentene
3M1Butene
N-Pentane
Iso-Pentane
Benzene
Hydrocarbon C6 non aromatic
Toluene
Hydrocarbon C7 non aromatic
Styrene
Ebenzene
Hydrocarbon C8 non aromatic
Hydrocarbon C9 and heavier
Water

- Design operating condition
    1. Inlet-outlet pressure each of equipments
    2. Inlet-outlet Temperature each of equipments

The equipments we mention are 3 compressors, 3 inter-stage coolers and 3 knock out drums

The actual operating data

For confirmation validated model, the actual operating condition of year 2009 is gathered. %polytropic efficiency, discharge temperature and wash water flow rate used in year 2009 are important. Filtered data to find minimum value, maximum value and trend of each parameter lead us understand how different between actual data and simulation results.

3.1.2 Validation model

Simulation models are powerful tools for process design and retrofits. The power of process modeling to a wider range of users, allowing companies to expand the use of process models to solve plant operating problems due to the complexities such as equipment troubleshooting, performance improvement. This study use HYSYS as simulation tool to validate simulation model. HYSYS offers a high degree of flexibility because there are multiple ways to accomplish specific tasks. This flexibility combined with a consistent and logical approach to how these capabilities are delivered makes HYSYS an extremely versatile process simulation tool.

To solve with program HYSYS, we have to know primary interface element. There are five primary interface elements for interacting with HYSYS:

Table 3.1 Primary interface elements for interacting with HYSYS

Interface Element	Description
PFD	A property view containing a graphical environment for building your flowsheet and examining process connectivity. Process information can be displayed for each individual stream or operation as needed.
Workbook	A property view containing a collection of tabs that displays information in a tabular format. Each Workbook tab displays information about a specific object type. You can install multiple tabs for a given object type, displaying information in varying levels of detail.
Property View	A single property view that contains multiple tabs. HYSYS extensively uses these single property views, which include all information about a specific object (in other words, an individual stream or operation).
Summary View	Displays the currently installed streams and operations.
Simulation Navigation	A property view that provides a single location for viewing all stream and unit operation property views in the simulation case, regardless of the flowsheet they exist in.

For open a new case in HYSYS program, the Components tab is where you define the sets of chemical components used in the simulation. These component sets are stored in Component Lists and can include library pure components and/or hypothetical components. The data from design operating data is performed as components in list.

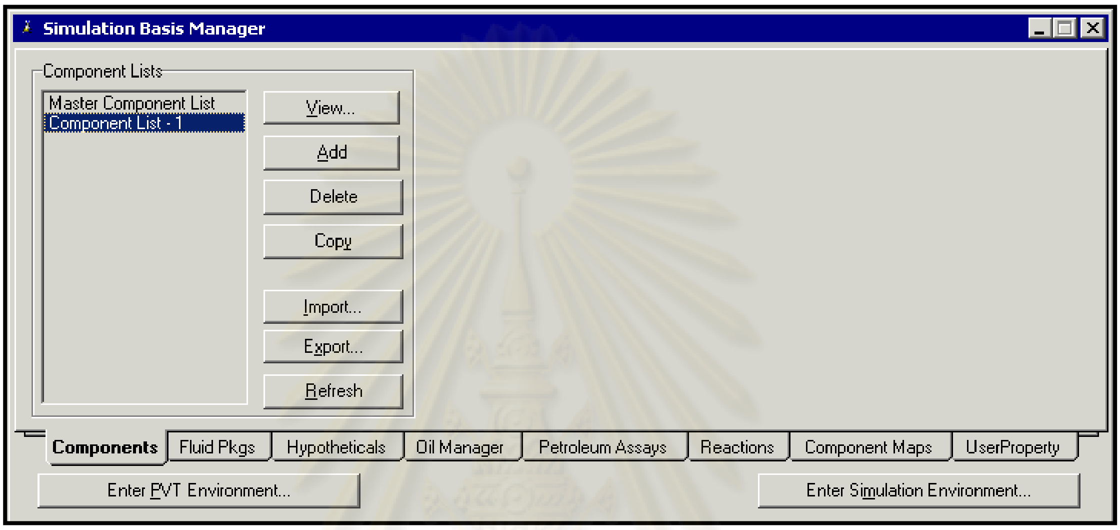


Figure 3.7 Components Tap

The Components tab contains a Master Component List that cannot be deleted. This master list contains every component available from "all" component lists. If components are added to any other component list, they are automatically added to the Master Component List. Also, if a component is deleted from the master list, it is deleted from all other component lists using that component.

The fluid package contains all the necessary information for pure component flash and physical property calculations. This allows defining all the required information inside a single entity. There are four key advantages to using fluid packages:

• All associated information is defined in a single location for easy creation and modification.

• Fluid packages can be exported and imported as completely defined packages for use in any simulation.

• Fluid packages can be cloned, reducing the time involved in creating and/or modifying complex fluid packages.

• Multiple fluid packages can be used in the same simulation.

In fluid packages tab, it allows to select thermodynamic equation in Lists of property package selection. The thermodynamic equation chosen will explain behavior or characteristic of input-output steam.

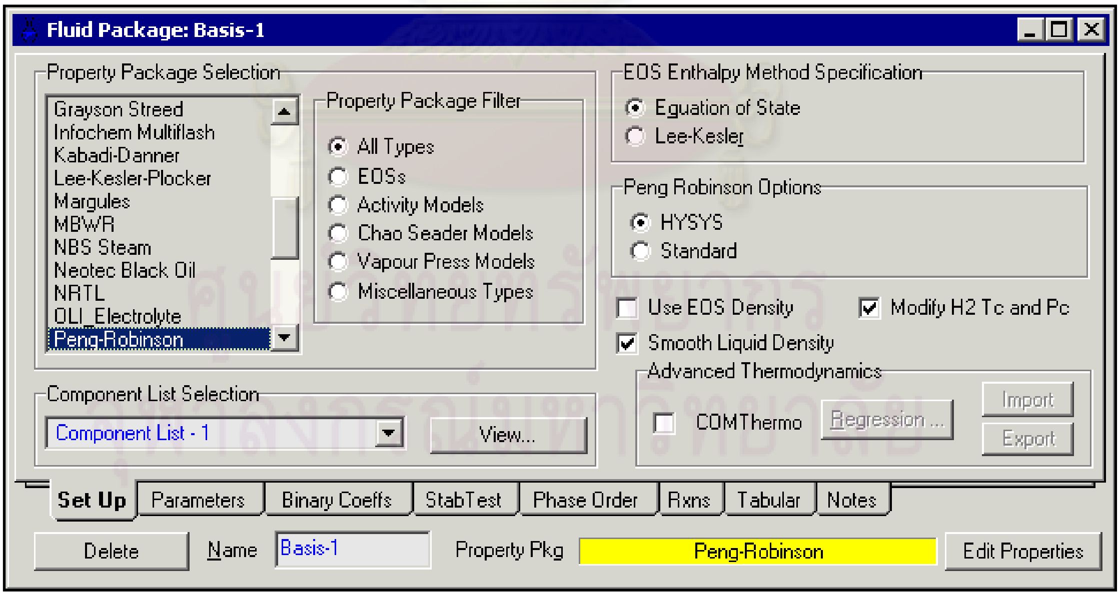


Figure 3.8 Fluid package Tap

Thermodynamic Equation

Thermodynamic equations describe the state of matter under a given set of physical conditions. It is a constitutive equation which provides a mathematical relationship between two or more state functions associated with the matter, such as its temperature, pressure, volume, or internal energy. Equations of state are useful in describing the properties of fluids, mixtures of fluids and solids. Peng-Robinson thermodynamic equation leads to solve this study.

Peng-Robinson equation of state

$$p = \frac{RT}{V_m - b} - \frac{a\,\alpha}{V_m^2 + 2bV_m - b^2}$$

$$a = \frac{0.45724\,R^2\,T_c^2}{p_c}$$

$$b = \frac{0.07780\,RT_c}{p_c}$$

$$\alpha = \left(1 + \left(0.37464 + 1.54226\,\omega - 0.26992\omega^2\right)\left(1 - T_r^{0.5}\right)\right)^2$$

$$T_r = \frac{T}{T_c}$$

In polynomial form:

$$A = \frac{a\alpha p}{R^2\,T^2}$$

$$B = \frac{bp}{RT}$$

$$Z^3-(1-B)\ Z^2+(A-2B-3B^2)\ Z-(AB-B^2-B^3)=0$$

where, ω is the acentric factor of the species, *R* is the universal gas constant and Z=PV/(RT) is compressibility factor.

The Peng-Robinson equation was developed in 1976 in order to satisfy the following goals.

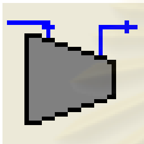
1. The parameters should be expressible in terms of the critical properties and the acentric factor.
2. The model should provide reasonable accuracy near the critical point, particularly for calculations of the compressibility factor and liquid density.
3. The mixing rules should not employ more than a single binary interaction parameter, which should be independent of temperature pressure and composition.
4. The equation should be applicable to all calculations of all fluid properties in natural gas processes.

For the most part the Peng-Robinson equation exhibits performance similar to the Soave equation, although it is generally superior in predicting the liquid densities of many materials, especially nonpolar ones. The departure functions of the Peng-Robinson equation are given on a separate article.

After we add components list, fluid package and reaction, we can enter to simulation environment. When the window of simulation environment is reached, objects palettes are used to build simulation within the Simulation environment. The objects in HYSYS are streams, unit operations, and logical operations.

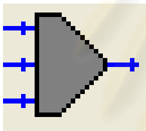
Objects are chosen for HYSYS simulation in this research as follow

1. Compressor



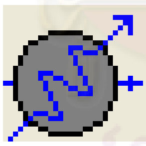
This object palette use as compressor for compress cracked gas.

2. Mixing tank



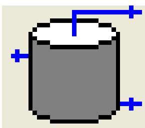
This object palette use as mixing tank that is the path of water injection.

3. Heat exchanger



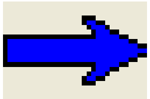
This object palette use as inter-stage aftercooler for reduce temperature of cracked gas after compression.

4. Gas-Liquid drum



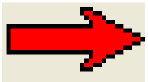
This object palette use as knock out drum that 2 phase of gas-liquid from inter-stage aftercooler come before pass to the next

5. Steam line



This object palette use as steam input-output.

6. Energy line



This object palette use as energy input-output

3.1.3 Wash water optimization for this thesis

After achieve simulation HYSYS model, the optimization of studied factor will be started. The maximum rate of wash water from actual operating data is gradually reduced in order to obtain optimal flow rate to achieve compressor discharge temperature at 90 °C in simulation model. This temperature 90 °C has less impact on high fouling in compressors. The optimal wash water flow rate was determined with its constraint of 1,200 l/hr to 100 l/hr.

CHAPTER IV

RESULTS AND DISCUSSION

4.1 Cracked gas compressor model

With the use of design operating conditions as input data on HYSYS program, the HYSYS model is developed and tested. The 4th stage of compressors is neglected because it has no problem from polymer fouling. Therefore, the HYSYS model is developed to handle the first 3 stages of compressors.

There are three steps of validation simulation model which the first, the second and the third steps are 1.simulate the model by no adding wash water, 2.improve the model by adding wash water and 3.improve the model by adding extra object.

4.1.1 Step one: Simulate the model by no adding wash water

This step is started with putting the design operating information to HYSYS program. The series of this model combine with 3 compressors, 2 inter-stage coolers and 2 gas – liquid drums. The simulation results show difference of discharge temperature for compressor stage 1 and %temperature difference is 19.3% while other parameters look suitable.

Table 4.1 Operating data and simulation results of compressor stage 1

Parameters		Compressor stage 1	
		Suction	Discharge
Design Operating condition	Mass flowrate, kg/hr	119,016.00	119,016.00
	Pressure, kg/cm^2	1.73	3.71
	Temperature, ^oC	39.45	87.50
Condition from simulation (no wash water)	Mass flowrate, kg/hr	119,016.00	119,016.00
	Pressure, kg/cm^2	1.73	3.71
	Temperature, ^oC	39.45	104.40

$$\%difference = \frac{|operating\ data - simulation\ data|\ x\ 100}{operating\ data}$$ .... (Eq.4.1)

For compressor stage 2, the simulation results show difference of discharge temperature and total mass flow rate. They are 18.9%of temperature difference and 3.27%of total mass flow rate difference.

Table 4.2 Operating data and simulation results of compressor stage 2

Parameters		Compressor stage 2	
		Suction	Discharge
Design Operating condition	Mass flowrate, kg/hr	123,042.00	123,042.00
	Pressure, kg/cm^2	3.41	8.42
	Temperature, ^oC	33.51	98.13
Condition from simulation (no wash water)	Mass flowrate, kg/hr	119,016.00	119,016.00
	Pressure, kg/cm^2	3.41	8.42
	Temperature, ^oC	33.51	109.8

For compressor stage 3, the simulation results show difference of discharge temperature and total mass flow rate. They are 14.9% of temperature difference and 2.5%of total mass flow rate difference.

Table 4.3 Operating data and simulation results of compressor stage 3

Parameters		Compressor stage 3	
		Suction	Discharge
Design Operating condition	Mass flowrate, kg/hr	121,176.00	121,176.00
	Pressure, kg/cm^2	8.08	17.99
	Temperature, ^{o}C	35.76	92.64
Condition from simulation (no wash water)	Mass flowrate, kg/hr	114,157.00	114,157.00
	Pressure, kg/cm^2	8.08	17.99
	Temperature, ^{o}C	35.76	104.30

In step one, the simulation results show that %difference of discharge temperature is higher than 10% of all stages. This error is performed because it has no wash water in model. So, next step gives us more effective simulation model by adding wash water injection.

4.1.2 Step two: improve the model by adding wash water

This step is improved from step one by adding actual maximum wash water flow rate from the existing plant to the model. The maximum wash water flow rate of compressor stage 1, stage 2 and stage 3 are 1,200, 1,200 and 780 kg/hr, respectively. The FPD case shows series of this model combined with 3 compressors, 2 inter-stage coolers, 2 gas – liquid drums and 3 mixing tanks. The modified model can provide more accurate prediction of the simulated condition. The comparison between operating data and simulation model after add wash water can be shown in Table 4.4-4.6.

Table 4.4 Operating data and simulation results of compressor stage 1: Add wash water

Parameters		Compressor stage 1	
		Suction	Discharge
Design Operating condition	Mass flowrate, kg/hr	119,016.00	119,016.00
	Pressure, kg/cm^2	1.73	3.71
	Temperature, ^oC	39.45	87.50
Condition from simulation (add wash water)	Mass flowrate, kg/hr	119,016.00	119,016.00
	Pressure, kg/cm^2	1.73	3.71
	Temperature, ^oC	39.45	93.54

The simulation results after completely run program HYSYS still show difference of discharge temperature for compressor stage 1 and %temperature difference is 6.9% while other parameters look suitable.

Table 4.5 Operating data and simulation results of compressor stage 2: Add wash water

Parameters		Compressor stage 2	
		Suction	Discharge
Design Operating condition	Mass flowrate, kg/hr	123,042.00	123,042.00
	Pressure, kg/cm^2	3.41	8.42
	Temperature, ^oC	33.51	98.13
Condition from simulation (add wash water)	Mass flowrate, kg/hr	120,216.00	120,216.00
	Pressure, kg/cm^2	3.41	8.42
	Temperature, ^oC	33.51	96.41

For compressor stage 2, the simulation results show small difference of discharge temperature and total mass flow rate. They are 1.8%of temperature difference and 2.3%of total mass flow rate difference.

Table 4.6 Operating data and simulation result of compressor stage 3: Add wash water

Parameters		Compressor stage 3	
		Suction	Discharge
Design Operating condition	Mass flowrate, kg/hr	121,176.00	121,176.00
	Pressure, kg/cm^2	8.08	17.99
	Temperature, ^{o}C	35.76	92.64
Condition from simulation (add wash water)	Mass flowrate, kg/hr	118,953.00	118,953.00
	Pressure, kg/cm^2	8.08	17.99
	Temperature, ^{o}C	35.76	101.50

The results of simulation of compressor stage 3 show small difference of discharge temperature and total mass flow rate. They are 9.6% of temperature difference and 1.8%of total mass flow rate difference.

In step two, the simulation results show that % difference of discharge temperature is better than step one and %difference of discharge temperature of all stages are lower than 10%. However, simulation model should be improved to provide next process of optimization. So, next step is carried out to provide effective simulation model by adding extra object in order to replacement of heat loss.

4.1.3 Step three: improve the model by adding extra object.

For the next step of wash water optimization, the model has to finally improve by adding more one heat exchanger per stage of compressor to adjust discharge temperature as similar to actual discharge temperature in average. The maximum wash water flow rate and average discharge temperature of actual operating data are considered in this step as baseline of simulation.

Table 4.7 Maximum wash water flow rate and discharge temperature of compressor from actual operating data

Description	Compressor stage 1	Compressor stage 2	Compressor stage 3
Maximum Wash water, kg/hr	1200	1200	780
Maximum Discharge Temp, ^{o}C	91.26	90.36	87.93
Minimum Discharge Temp, ^{o}C	75.50	75.66	87.30
Average Discharge Temp, ^{o}C	83.38	83.01	87.62

Series of final simulation model combine with 3 compressors, 3 mixing tanks, 5 heat exchangers and 2 gas – liquid drums. The extra heat exchanger per stage has a role as a factor replacing heat loss of model. This method let us to get simulation condition like baseline condition. The baseline condition obtained from the step three is not equivalent to the designed operating condition because an error from the simulation is taken place with respect to more than 1 scenario given. However, correlation analysis between designed operating information and actual information is in the acceptable range: that is in the range of 4-10 %error.

Therefore, this model can be applicable to provide simulation of other cracked gas compressor units according to several olefins feed types with the concern of feed composition change.

4.2 Wash water optimization

The key of this research is to carry out wash water optimization. The optimal wash water flow rate is obtained from analysis relationship between simulation results and actual data. Actual data from real operation is considered to determine variation range of wash water flow rate relating with discharge temperature. The process to find out optimal wash water flow rate is done when wash water is gradually reduced in order to obtain optimal flow rate to achieve compressor discharge temperature at 90 ^{o}C which has less impact on high fouling rate in compressors. From actual data, it shows maximum data, minimum data and average data to arrange analysis information.

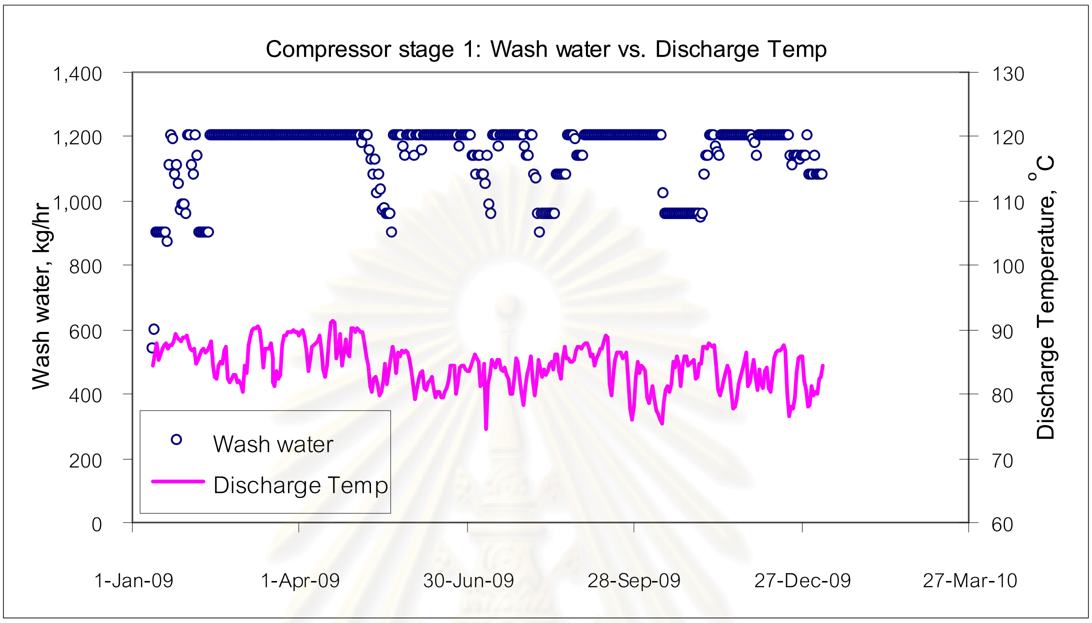


Figure 4.1 Actual data: Wash water flow rate vs. Discharge Temperature of compressor stage 1

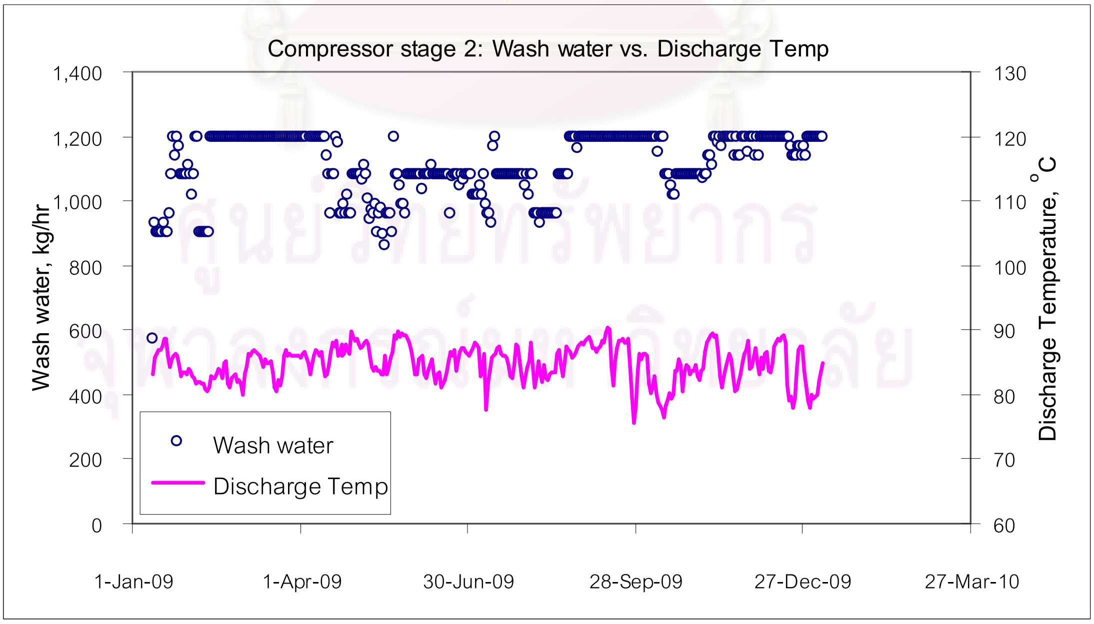


Figure 4.2 Actual data: Wash water flow rate vs. Discharge Temperature of compressor stage 2

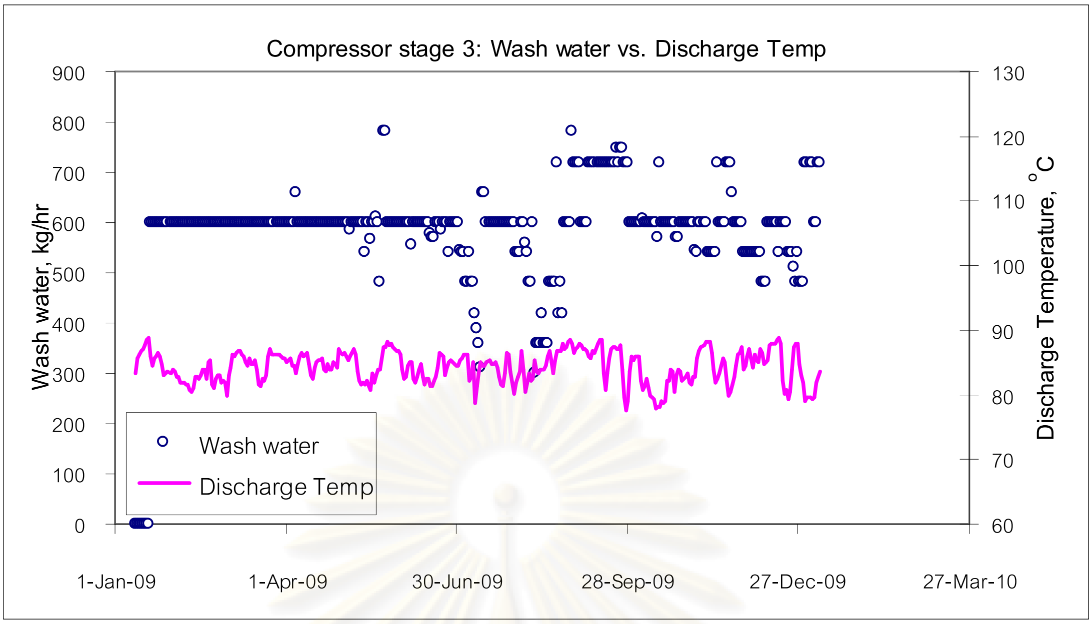


Figure 4.3 Actual data: Wash water flow rate vs. Discharge Temperature of compressor stage 3

Figure 4.1-4.3 show span of actual wash water flow rate which can control compressor discharge temperature to be in acceptable condition. Compressor of each stage has own span using as information comparing with simulation optimization analysis. At the lowest of discharge temperature of baseline temperature 83 °C to 90 °C, it is conducted to perform as independent variable. Wash water flow rate from simulation results by vary discharge temperature are shown in Table 4.8 and Figure 4.4.

Table 4.8 Wash water flow rate results from simulation model

Discharge Temperature, °C	Wash water flow rate , kg/hr		
	Compressor stage 1	Compressor stage 2	Compressor stage 3
83	1,241	1,200	1,315
84	1,128	1,080	1,200
85	1,015	965	1,080
86	907	845	964
87	800	730	849
88	690	615	737
89	575	495	620
90	465	380	505

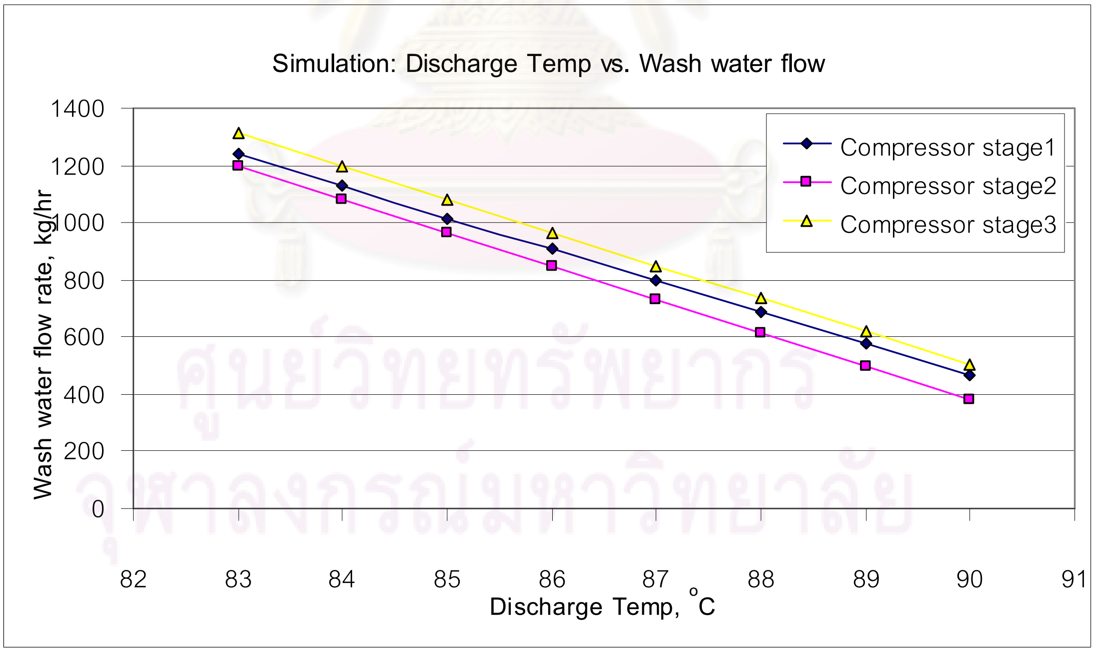


Figure 4.4 Relationship between wash water flow rate and discharge temperature from simulation model for 3 stage of compressors by vary discharge temperature

Besides discharge temperature, the optimal wash water flow rate is independent variable and determined with its constraint of 1,200 kg/hr to 100 kg/hr. The maximum wash water flow rate 1,200 kg/hr can be gradually reduced to 100 kg/hr to find out additional information to analyze optimal wash water flow rate at optimal discharge temperature. Discharge temperature from simulation results by vary wash water flow rate are shown in Table 4.9 and Figure 4.5.

Table 4.9 Discharge Temperature results from simulation model

Wash water flow rate , kg/hr	Discharge Temperature, ^{o}C		
	Compressor stage 1	Compressor stage 2	Compressor stage 3
1200	83.4	83.01	83.98
1100	84.3	83.86	84.84
1000	85.2	84.71	85.71
900	86.1	85.56	86.57
800	87.0	86.42	87.44
700	87.9	87.27	88.3
600	88.8	88.12	89.17
500	89.7	88.98	90.03
400	90.6	89.83	90.9
300	91.5	90.69	91.77
200	92.4	91.54	92.64
100	93.3	92.4	93.51

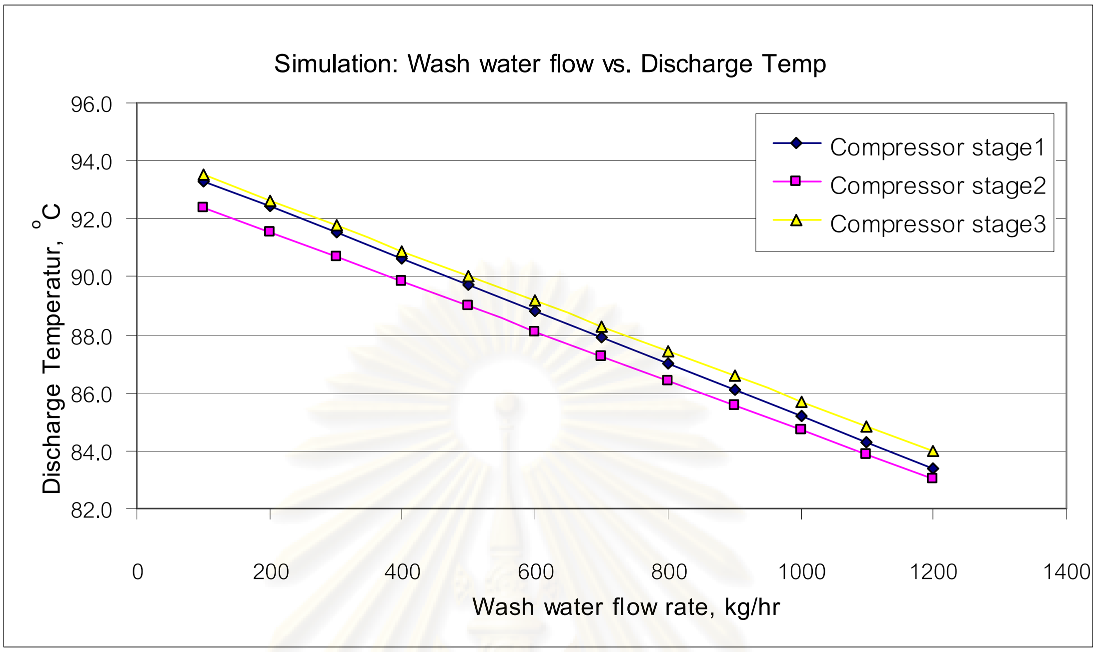


Figure 4.5 Relationship between wash water flow rate and discharge temperature from simulation model for 3 stage of compressors by vary wash water flow rate

Generally, polymerization rate in compressor is related with discharge temperature. Thus, this study considers 3 factors: polymerization rate, discharge temperature and wash water flow rate to analyze correlation among them. This relation can be shown in Figure 4.6-4.8.

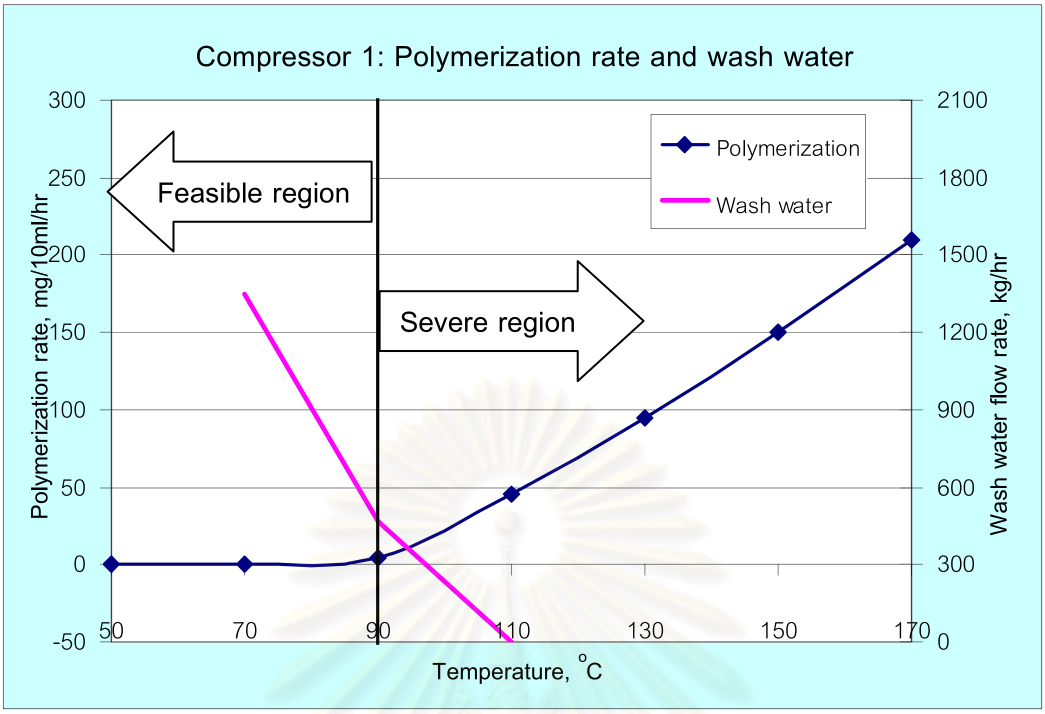


Figure 4.6 Relationship between polymerization rate and wash water flow rate of compressors stage 1

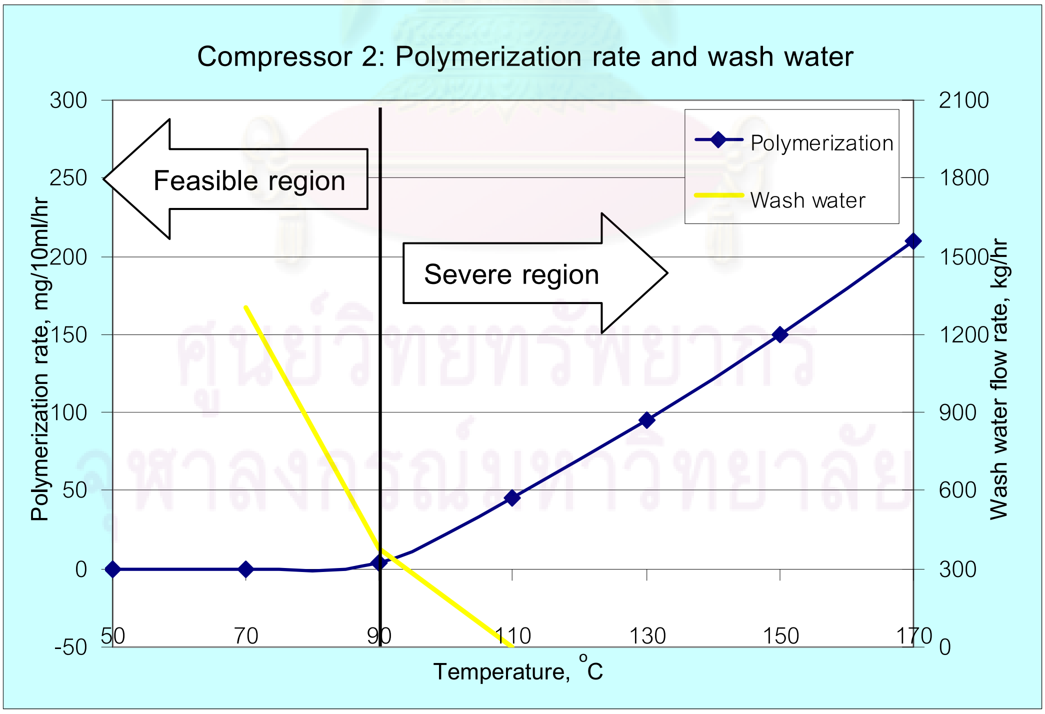


Figure 4.7 Relationship between polymerization rate and wash water flow rate of compressors stage 2

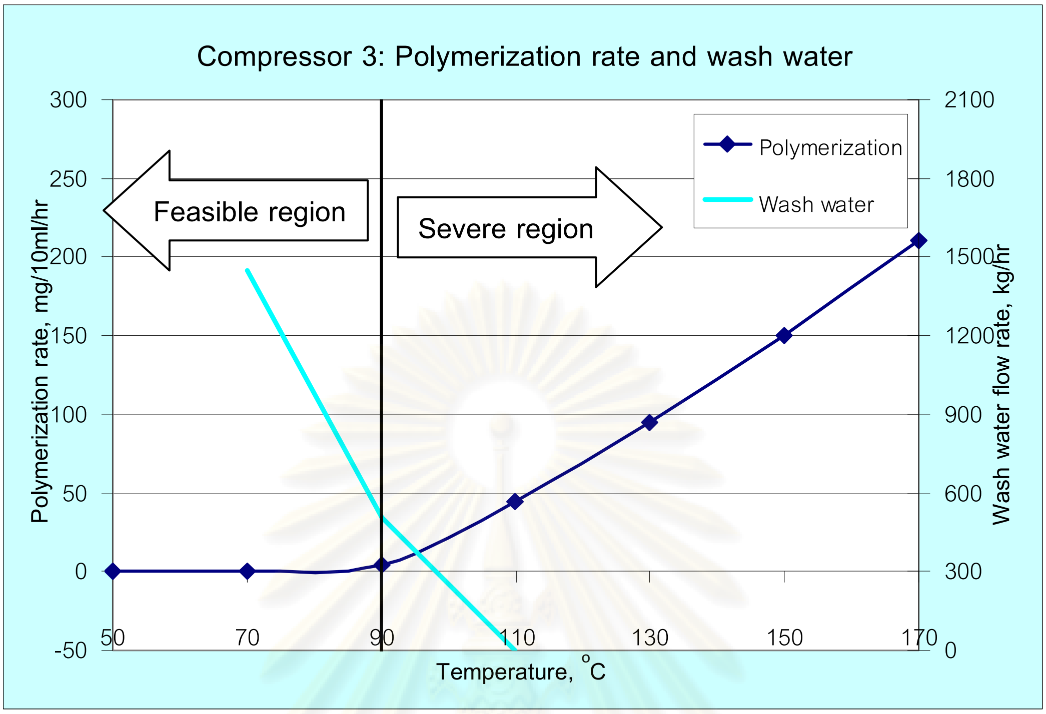


Figure 4.8 Relationship between polymerization rate and wash water flow rate of compressors stage 3

When analyze from relationship between polymerization rate and wash water flow rate as figures above, we get safety area or feasible region to verify optimal wash water flow rate. The feasible region show range of wash water flow rate that can control compressor discharge temperature at low polymerization rate. Severe region is high polymerization area that shows improper wash water flow rate controlling compressor discharge temperature.

Table 4.10 Actual data and simulation data of wash water flow rate

Description		Compressor stage 1	Compressor stage 2	Compressor stage 3
Actual data	Maximum Wash water, kg/hr	1200.0	1200.0	780.0
	Minimum Wash water, kg/hr	540.0	570.0	0.0
Simulation	Minimum Wash water, kg/hr (at discharge temp 90°C)	465.0	380.0	505.0

As compare with actual data and simulation data, compressor stage 1 and stage 2 use wash water flow rate to control discharge temperature in acceptable rate and in feasible region that safe for high polymerization rate. However, wash water flow rate using at compressor stage 3 has many actual data are in severe region and minimum rate of wash water is 0.0 kg/hr while minimum rate from simulation is 505 kg/hr. So, wash water for compressor stage 3 should be maintained at least 500 kg/hr to minimize high fouling from polymerization.

4.3 Saving obtained from optimization

Fouling in cracked gas compressor can negatively affect plant economics. In addition to energy lost and production loss, a fouled compressor may require an unscheduled shutdown. Discharge temperature control is the one of main factors for high polymerization rate in charge gas compressor. Improper wash water which can not control discharge temperature for a long time is considered as high impact on fouling in compressor that unscheduled shutdown can be occurred. The total loss from fouling and unscheduled shut down of charge gas compressor fouling is calculated from productivity lost, operating cost and compressor unit cleaning cost.

Total cost from fouling and unplanned shutdown = Productivity loss + Operating cost + Cleaning cost (Eq.4.2)

Productivity loss can be calculated from total cost of ethylene product loss during shut down period that plant disappear income from selling product.

Productivity loss = Productivity or ethylene price x Number of days during shutdown (Eq.4.3)

Table 4.11 Calculation total cost from productivity loss

Productivity loss		
Shutdown period	5	days
Ethylene production	912	Ton/day
Ethylene price	804	USD/Ton
Total cost	3,666,240	USD

Normally, studied ethylene plant has ethylene production rate 912 Ton/day and ethylene can sell as price 804 USD/Ton. For summary, total cost of productivity loss can be seen from Table 4.11 and it is 3,666,240 USD

Operating cost estimate from productivity reduction and steam driving turbine increasing during plant operate with fouling condition. Generally, plant shutdown is not occurred immediately when plant has sign of fouling. Plant will extend operation to prepare and provide procedure before shutdown and after start up to minimize loss during shutdown. The extension may 90 days after plant has sign of fouling.

Operating cost = Productivity reduction + Steam increasing during fouling (Eq.4.4)

Table 4.12 Calculation total cost from plant operates with fouling condition

Productivity reduction(1)		
Fouling period	90	Days
10% ethylene reduction	91.2	Ton/day
Ethylene price	804	USD/Ton
Cost	6,599,232	USD
Steam increasing(2)		
Fouling period	90	Days
Steam increasing	106	Ton/day
Steam price	750	USD/Ton
Cost	7,174,500	USD
Operating cost		
Total cost = (1)+(2)	13,773,732	USD

Operating cost is calculated from productivity reduction with percent of product cost reduction 10% and cost of steam increasing to try to drive discharge pressure to maintain desired discharge pressure for 90 days. Cost of steam is 750 USD/Ton and total operating cost is 13,773,732 USD as Table 4.12.

Cleaning cost is the expense paid for employing contractors and renting equipments to clean compressors and it is indicated as Equation 4.4.

Cleaning cost = Cost from cleaning compressor unit (Eq.4.5)

Table 4.13 Calculation total cost with considering plant cleaning

Cleaning cost		
Shutdown	1	Times
Wash oil cost for cleaning	19,000	USD
Offline cleaning cost	15,000	USD
Total cost	34,000	USD

Total cost from cleaning summary from cost of wash oil circulation in compressor unit 19,000 USD and cost of offline cleaning 15,000 USD such as man hour and mechanical cleaning cost. In table 4.13, the total cleaning cost is 34,000 USD.

Total saving from proper wash water flow rate control is saving from no payment due to fouling and unplanned shutdown.

Saving = Total cost from fouling and unplanned shutdown (Eq.4.6)

Table 4.14 Calculation of total saving with proper wash water flow rate

Saving = Total cost from fouling and unplanned shutdown		
Productivity loss	3,666,240	USD
Operating cost	13,773,732	USD
Cleaning cost	34,000	USD
Total saving	17,473,972	USD

Benefit of compressor discharge temperature control by optimal wash water flow rate is cost saving 17,473,972 USD per one times. For planning shutdown for 4 year, it can be subdivided to three cases considering %saving.

Table 4.15 Calculation total saving from use proper wash water flow rate

Cleaning cost	Cost of fouling, USD	Productivity cost, USD	%Saving
Meet plan of shutdown 4 year	0	1,070,542,080	3.26
Unplanned shutdown 1 time	17,473,972	1,053,068,108	1.63
Unplanned shutdown 2 times	34,947,944	1,035,594,136	0

Case one is the case of no fouling and it can maintain plant reliability to 4 years or meet shutdown planning. Case two and case three are situation of unplanned shutdown from fouling for one and two times before reaching shutdown planning, respectively, Comparison between case one and case three, it has the saving of 3.26% or 34,947,944 USD while it has the saving of 1.63% or 17,473,972 USD when compared between case one and case two.

CHAPTER V

CONCLUSION AND RECOMMENTDATION

5.1 Conclusion

Factors involved in polymer fouling for a cracked gas compressor in an ethylene plant were studied in this work based on the formulation of a HYSYS model and simulation. With the real ethylene plant's data, the HYSYS model was formulated and tested the characteristic of a cracked gas compressor. Validated model was improved in three steps. In first, step one was step of simulation the model by no adding wash water. It showed high % different of discharge temperature between design operating data and actual operating data, so step two was considered. Then, step two was step of improvement the model by adding wash water. This step could give better adjustment of % different of discharge temperature between design operating data and actual operating data of compressor all stages and % different was lower than 10%. Finally, step three was the step of improvement the model by adding extra object for replacing heat loss and the model from this step was used to analyze correlation between discharge temperature and wash water flow rate in the next.

Then, with this model, relationship between discharge temperature and wash water flow rate was determined by compressor discharge temperature constraint of 83 ^{o}C to 90 ^{o}C and wash water constraint of 1,200 l/hr to 100 l/hr. It was found that the minimum wash water flow rate in each stages determined to achieve the discharge temperature of the compressors at 90^{o}C were at 465 l/hr, 380 l/hr and 505 l/hr corresponding to the 1st, 2nd and 3rd stage compressors respectively. The optimal wash water flow employed to handle the fouling of the compressors was found by analysis of relationship between polymerization rate and wash water flow rate.

From analysis the relationship between polymerization rate and wash water flow rate, we got safety area or feasible region to verify optimal wash water flow rate. The feasible region showed range of wash water flow rate that can control compressor discharge temperature at low polymerization rate. Severe region was high polymerization area that shows improper wash water flow rate controlling compressor discharge temperature.

As compared to actual data with feasible area, the compressor stage 1 and stage 2 could give wash water used at the flow rate to control discharge temperature in acceptable rate and in feasible region that safe for high polymerization rate. However, wash water flow rate using at compressor stage 3 had many actual data in severe region and minimum rate of wash water had seen at 0.0 kg/hr while minimum rate from simulation was 505 kg/hr. So, wash water for compressor stage 3 should be maintained at least 505 kg/hr to minimize high fouling from polymerization

Next, benefit of proper wash water flow rate control was avoidance of unplanned plant shutdown from polymer fouling. The total loss could calculate from productivity loss, operating cost and compressor unit cleaning cost. Optimal wash water flow rate could save 17,473,972 USD per one times of fouling and shutdown period.

For baseline of planning shutdown every 4 years, emergency shutdown was predicted to be taken place 2 times as maximum., The cost savings of the plant were 1.63% and 3.26% regarding the reduction of one time and two times of fouling and shutdown period respectively.

5.2 Recommendation

Nowadays, many plants have economics saving as a goal. Besides wash water, optimization wash oil rate for cracked gas compressors is the one of vital objectives for project cost saving. In future direction, wash oil flow rate which is important factor impact on fouling in compressors should be considered. The interest of wash oil is its properties; aromatic content, boiling point, diolefins level and existent gum. This involves in polymerization rate control in compressors. In order to minimize economic loss of the process, the wash oil objective function and constraint should be studied.

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APPENDICES

APPENDIX A

Operating Data for Simulation

Compressor stage 1					
Steam description	Steam Suction	Steam Discharge	Outlet heat exchanger	Liquid from drum	Vapor from drum
phase	Mixed	Vapor	Mixed	Mixed	Vapor
Temperature, oC	39.45	87.5	38.6	33.09	33.21
Pressure, Kg/cm2	1.73	3.71	3.43	1.9	3.41
Total Mass Rate, Kg/hr	119016	119016	119016	10845	121841.8
Hydrogen	3538.5	3538.5	3538.5	0.1	3538.5
Carbon monoxide	29.5	29.5	29.5	0	29.5
Carbon dioxide	66.8	66.8	66.8	0	66.9
Hydrogen sulfide	10.4	10.4	10.4	0	10.5
Methane	15786.7	15786.7	15786.7	1.9	15795
Acetylene	769.2	769.2	769.2	0.6	771.8
Ethylene	53305.5	53305.5	53305.5	28.2	53420.6
Ethane	21066.9	21066.9	21066.9	16.2	21131.8
Propylene	341.4	341.4	341.4	1.7	349.4
Propadiene	341.7	341.7	341.7	1.4	348.4
propylene	6295.8	6295.8	6295.8	15.7	6363.8
propane	1805.7	1805.7	1805.7	4.9	1825.1
Vacetylene	136.6	136.6	136.6	1.8	145.6
1-Butyne	58.5	58.5	58.5	0.8	62.7
1, 2-Butadiene	24.8	24.8	24.8	0.4	26.8
1, 3 –Butadiene	2455.6	2455.6	2455.6	23.2	2562
1Butene	430.4	430.4	430.4	3.7	447.1
Cis 2 Butene	80.9	80.9	80.9	1	85.6
Trans 2 Butene	110.4	110.4	110.4	1.3	116.1
IsoButadiene	642.8	642.8	642.8	5.4	667

Compressor stage 1					
Steam description	Steam Suction	Steam Discharge	Outlet heat exchanger	Liquid from drum	Vapor from drum
phase	Mixed	Vapor	Mixed	Mixed	Vapor
1,3 Cyclopentane	465.7	465.7	465.7	22.8	603.7
Cyclopentane	63.3	63.3	63.3	3.9	86.1
Trans 1, 3 Pentadiene	226.3	226.3	226.3	11.5	289.1
Isoprene	96.4	96.4	96.4	3.3	113.2
1Pentene	70.9	70.9	70.9	2.2	81.5
3M1Butene	3.8	3.8	3.8	0.1	4.2
N-Pentane	4.4	4.4	4.4	0.2	5.1
Iso-Pentane	5.4	5.4	5.4	0.1	6.1
Benzene	2454	2454	2454	1020.2	7069.5
C6 non aromatic	140.9	140.9	140.9	22.6	239.9
Toluene	608	608	608	644.7	1306.3
C7 non aromatic	84.2	84.2	84.2	66.1	219.1
Styrene	181	181	181	180.9	118.4
Ebenzene	476.5	476.5	476.5	1191.4	842.3
C8 non aromatic	144.4	144.4	144.4	327.5	363.6
C9 and heavier	1090.7	1090.7	1090.7	2117.7	744.3
Water	5162.9	5162.9	5162.9	5117.7	1529.8

Compressor stage 2					
Steam description	Steam Suction	Steam Discharge	Outlet heat exchanger	Liquid from drum	Vapor from drum
phase	Mixed	Vapor	Mixed	Mixed	Vapor
Temperature, oC	33.51	98.13	38	33.76	34.85
Pressure, Kg/cm2	3.41	8.42	8.14	3.76	8.08
Total Mass Rate, Kg/hr	123041.8	123041.8	123041.8	13670.9	119976.3
Hydrogen	3538.5	3538.5	3538.5	0.3	3539.1
Carbon monoxide	29.5	29.5	29.5	0	29.5
Carbon dioxide	66.9	66.9	66.9	0.1	66.9
Hydrogen sulfide	10.5	10.5	10.5	0.1	10.5
Methane	15795	15795	15795	10.2	15808.4
Acetylene	771.8	771.8	771.8	3.3	776.9
Ethylene	53420.6	53420.6	53420.6	143.3	53619.5
Ethane	21131.8	21131.8	21131.8	81	21248.4
Propylene	349.4	349.4	349.4	9.7	366.2
Propadiene	348.4	348.4	348.4	8.1	362.1
propylene	6363.8	6363.8	6363.8	83.6	6498.3
propane	1825.1	1825.1	1825.1	24.3	1863
Vacetylene	145.6	145.6	145.6	10.9	162.2
1-Butyne	62.7	62.7	62.7	5	70.9
1, 2-Butadiene	26.8	26.8	26.8	2.4	30.7
1, 3 –Butadiene	2562	2562	2562	129.6	2773.2
1Butene	447.1	447.1	447.1	20.4	480.2
Cis 2 Butene	85.6	85.6	85.6	5.7	94.7
Trans 2 Butene	116.1	116.1	116.1	7	127.4
IsoButadiene	667	667	667	29.6	715.1

Compressor stage 2					
Steam description	Steam Suction	Steam Discharge	Outlet heat exchanger	Liquid from drum	Vapor from drum
phase	Mixed	Vapor	Mixed	Mixed	Vapor
1,3 Cyclopentane	603.7	603.7	603.7	160.8	777.7
Cyclopentane	86.1	86.1	86.1	26.8	111.7
Trans 1, 3 Pentadiene	289.1	289.1	289.1	74.3	366.2
Isoprene	113.2	113.2	113.2	20.1	137.2
1Pentene	81.5	81.5	81.5	12.7	98.1
3M1Butene	4.2	4.2	4.2	0.5	4.9
N-Pentane	5.1	5.1	5.1	0.8	6
Iso-Pentane	6.1	6.1	6.1	0.8	7.2
Benzene	7069.5	7069.5	7069.5	5635.6	7118.2
C6 non aromatic	239.9	239.9	239.9	121.6	301.1
Toluene	1335.8	1335.8	1335.8	1342.9	509
C7 non aromatic	219.6	219.6	219.6	201.1	156
Styrene	118.4	118.4	118.4	118.3	14.7
Ebenzene	1202.3	1202.3	1202.3	1557.2	216.3
C8 non aromatic	458.6	458.6	458.6	546.7	150.3
C9 and heavier	1259.3	1259.3	1259.3	1771.3	154.4
Water	1729.8	1729.8	1729.8	1484.6	718.2

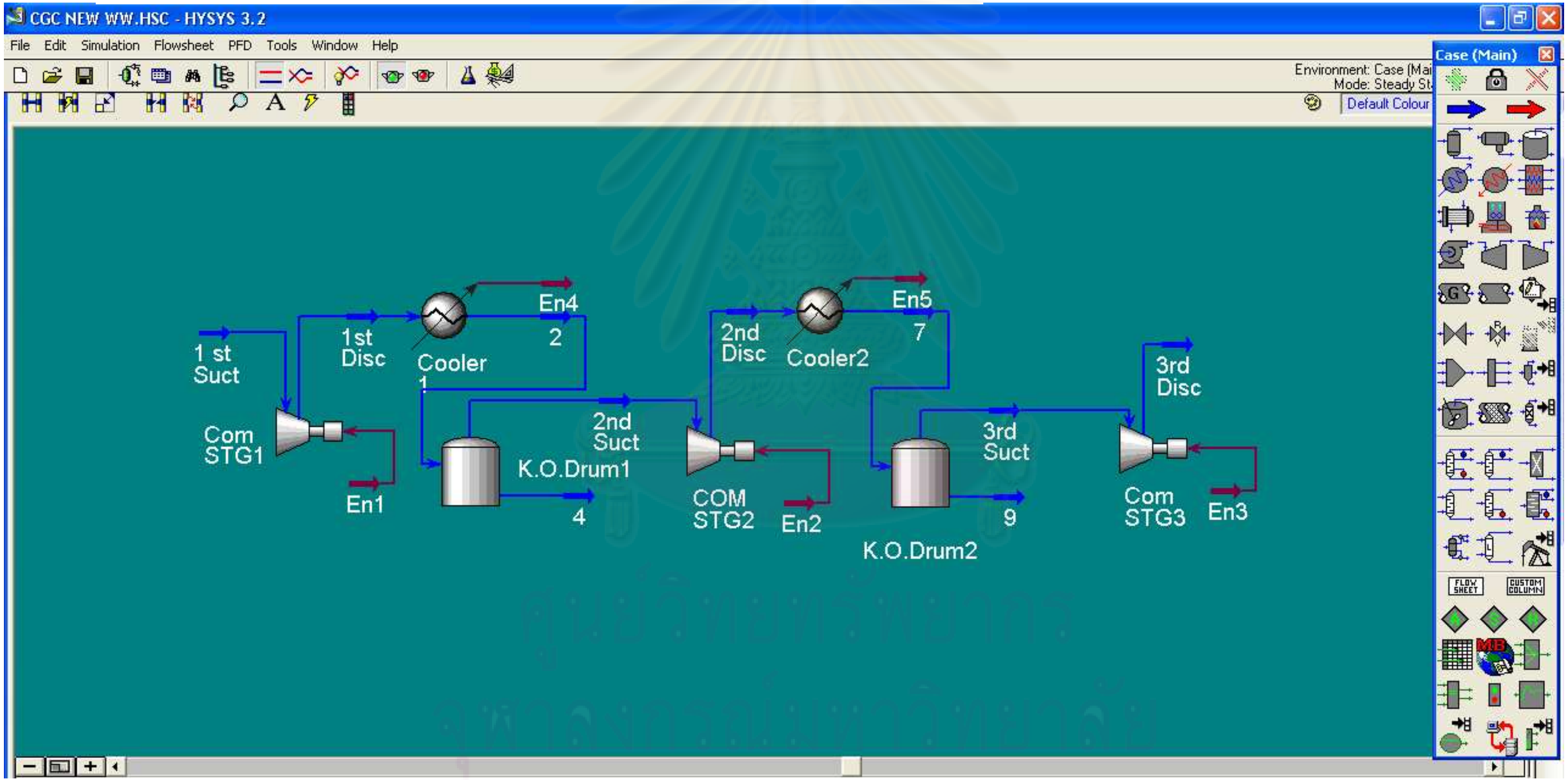
Compressor stage 3					
Steam description	Steam Suction	Steam Discharge	Outlet heat exchanger	Liquid from drum	Vapor from drum
phase	Mixed	Vapor	Mixed	Mixed	Vapor
Temperature, oC	35.76	92.64	37.9	36.4	37.83
Pressure, Kg/cm2	8.08	17.99	17.71	9.1	17.61
Total Mass Rate, Kg/hr	121176.3	121176.3	110720.3	5090.3	105630.1
Hydrogen	3539.1	3539.1	3233.7	0.3	3233.4
Carbon monoxide	29.5	29.5	27	0	27
Carbon dioxide	66.9	66.9	61.2	0.1	61
Hydrogen sulfide	10.5	10.5	9.6	0.1	9.5
Methane	15808.4	15808.4	14444.3	9.4	14434.9
Acetylene	776.9	776.9	709.8	2.7	707.1
Ethylene	53619.5	53619.5	48992.8	121	48871.8
Ethane	21248.4	21248.4	19415	67.5	19347.5
Propylene	366.2	366.2	334	7.3	327.3
Propadiene	362.1	362.1	330.8	6.2	324.7
propylene	6498.3	6498.3	5937.6	65.8	5871.8
propane	1863	1863	1702.3	19	1683.3
Vacetylene	162.2	162.2	148.2	7.7	140.6
1-Butyne	70.9	70.9	64.8	3.5	61.3
1, 2-Butadiene	30.7	30.7	28	1.7	26.4
1, 3 –Butadiene	2773.2	2773.2	2533.9	93.6	2440.3
1Butene	480.2	480.2	438.8	14.8	424
Cis 2 Butene	94.7	94.7	86.5	4	82.5
Trans 2 Butene	127.4	127.4	116.4	5	111.4
IsoButadiene	715.1	715.1	653.4	21.6	631.9
IsoButane	155.5	155.5	142.1	3.8	138.3

Compressor stage 3					
Steam description	Steam Suction	Steam Discharge	Outlet heat exchanger	Liquid from drum	Vapor from drum
phase	Mixed	Vapor	Mixed	Mixed	Vapor
Cyclopentane	111.7	111.7	102	15.9	86.2
Trans 1, 3 Pentadiene	366.2	366.2	334.6	45.4	289.3
Isoprene	137.2	137.2	125.4	12.9	112.5
1Pentene	98.1	98.1	89.7	8.3	81.4
3M1Butene	4.9	4.9	4.5	0.3	4.2
N-Pentane	6	6	5.5	0.5	5
Iso-Pentane	7.2	7.2	6.6	0.6	6.1
Benzene	7118.2	7118.2	6504	2382.6	4121.4
C6 non aromatic	301.1	301.1	275.1	62.4	212.7
Toluene	538.5	538.5	492.1	318.5	173.6
C7 non aromatic	156.5	156.5	143	67.1	75.9
Styrene	14.7	14.7	13.5	11.3	2.1
Ebenzene	576.3	576.3	526.6	435.2	91.4
C8 non aromatic	245.3	245.3	224.1	158.2	65.9
C9 and heavier	669.4	669.4	611.7	537.4	74.3
Water	918.2	918.2	839	470	368.9

APPENDIX B

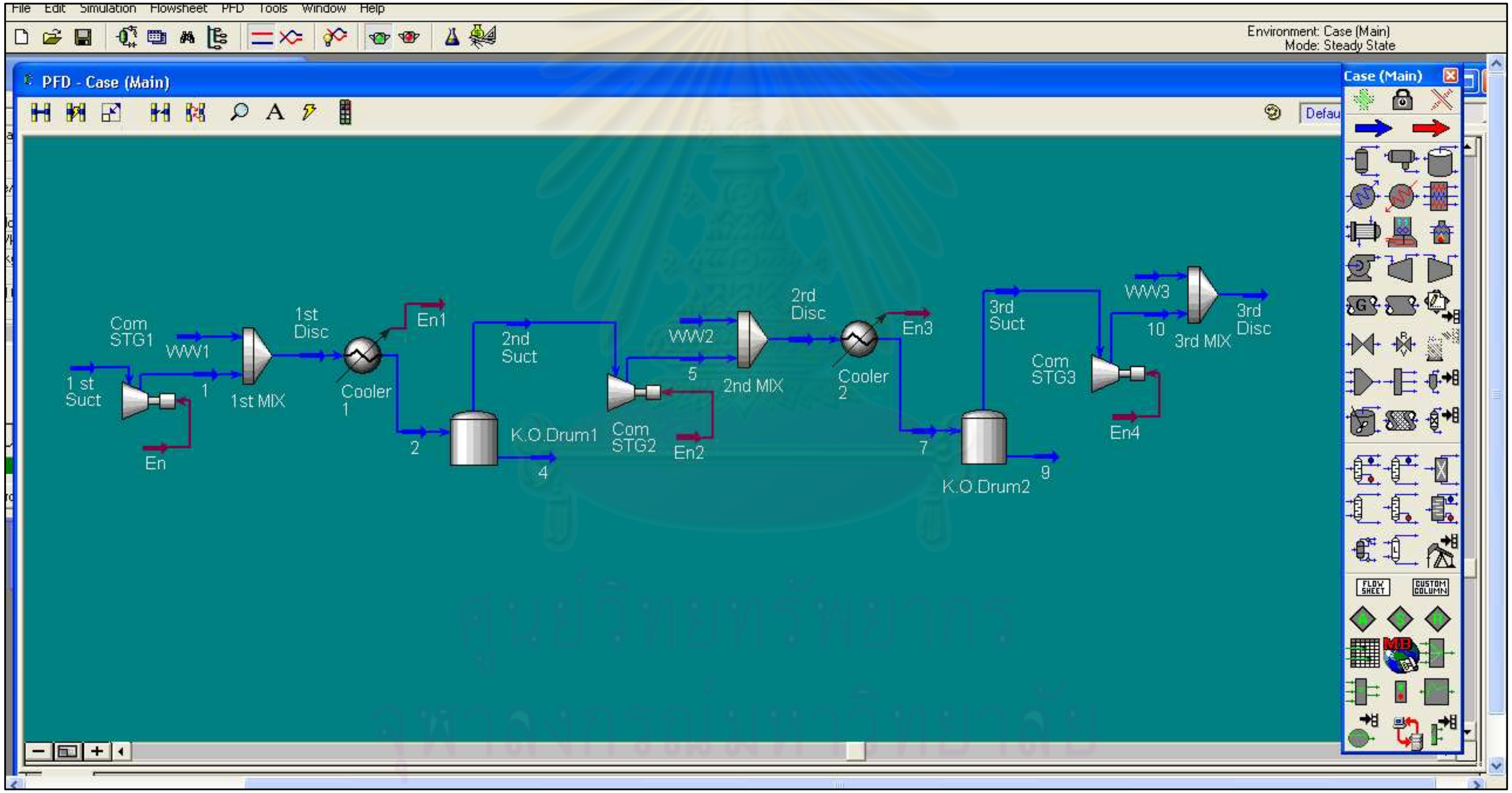
Simulation Results

Step 1: Simulation before adding wash water

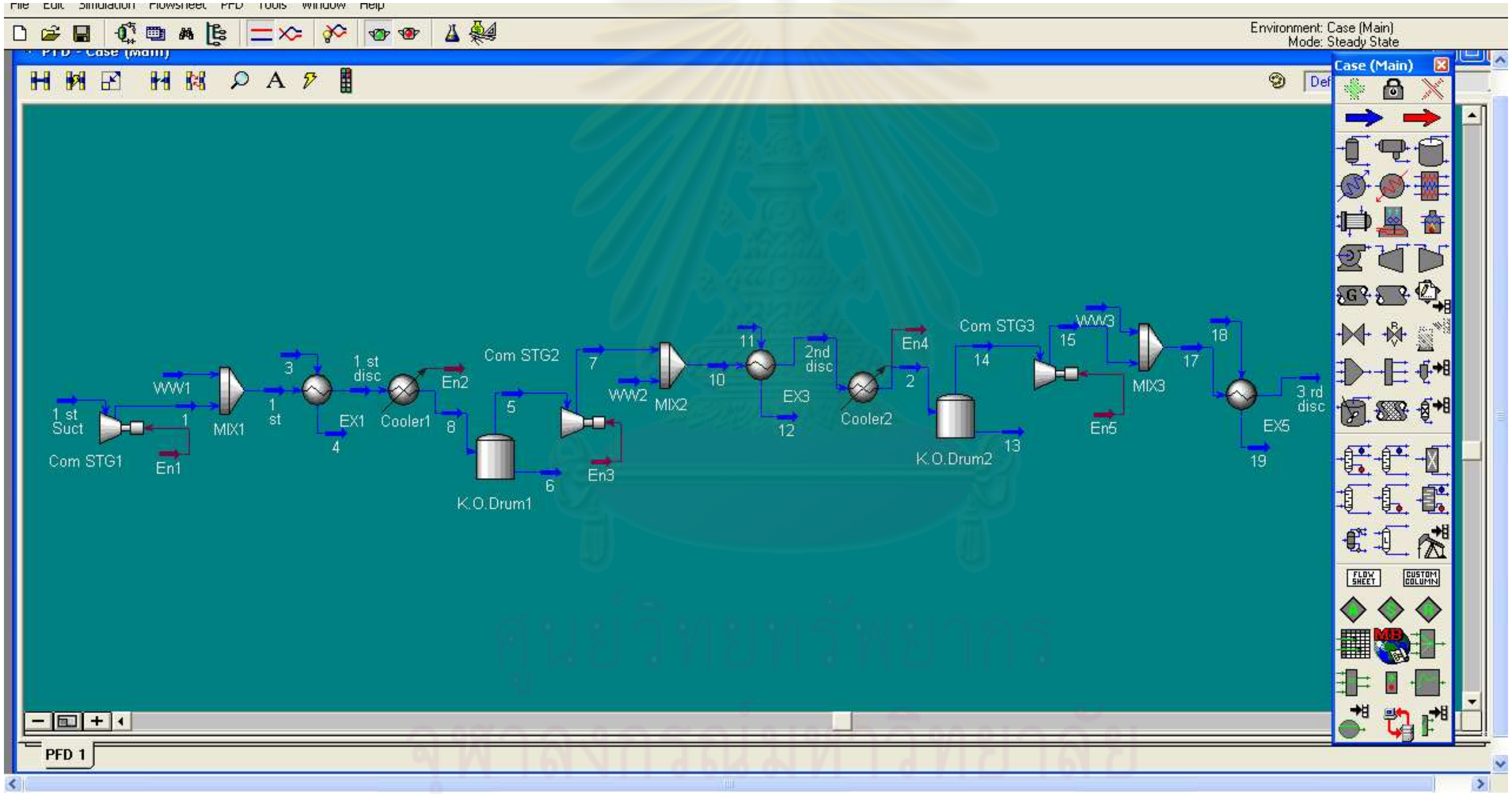


79

Step 2: Simulation after adding wash water



Step 3: Simulation after adding extra exchanger



81

VITA

Miss Piyachat Puttaraksa was born on October 25, 1983 in Rayong, Thailand, She graduated her bachelor Degree from Department of Chemical Technology in the Faculty of Science from Chulalongkorn University in 2006. In 2007, she entered the Graduated School of Chulalongkorn University to propose a Master of Engineering in Chemical Engineering and completed in 2010 with the thesis entitled “Simulation And Optimization for Compressor unit of Cracked gas in Ethylene plant“.

Publications

Piyachat P. and Paisan K., “Simulation and optimization for compressor unit of cracked gas in ethylene plant”, *The Thai Institute of Chemical Engineering and Applied Chemistry (TIChE 2010)* pp.107, Bangkok, Thailand, 2010.