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EGE UNIVERSITY
Graduate School of Applied and Natural Science

TECHNICAL FEASIBILITY OF HEAT INTEGRATION FOR THE PETROCHEMICAL ETHYLENE PLANT

MSc THESIS

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Sayın **Halil ÇELİKKAYA** tarafından YÜKSEK LİSANS TEZİ olarak sunulan "**Technical Feasibility of Heat Integration for the Petrochemical Ethylene Plant**" başlıklı bu çalışma E.Ü. Lisansüstü Eğitim ve Öğretim Yönetmeliği ile E.Ü. Fen Bilimleri Enstitüsü Eğitim ve Öğretim Yönergesi' nin ilgili hükümleri uyarınca tarafımızdan değerlendirilerek savunmaya değer bulunmuş ve 27.08.2024 tarihinde yapılan tez savunma sınavında aday oybirliği/oyçokluğu ile başarılı bulunmuştur.

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EÜ Lisansüstü Eğitim ve Öğretim Yönetmeliğinin ilgili hükümleri uyarınca Yüksek Lisans Tezi olarak sunduğum "**Technical Feasibility of Heat Integration for the Petrochemical Ethylene Plant**" başlıklı bu tezin kendi çalışmam olduğunu, sunduğum tüm sonuç, doküman, bilgi ve belgeleri bizzat ve bu tez çalışması kapsamında elde ettiğimi, bu tez çalışmasıyla elde edilmeyen bütün bilgi ve yorumlara atıf yaptığımı ve bunları kaynaklar listesinde usulüne uygun olarak verdiğimi, tez çalışması ve yazımı sırasında patent ve telif haklarını ihlal edici bir davranışımın olmadığını, bu tezin herhangi bir bölümünü bu üniversite veya diğer bir üniversitede başka bir tez çalışması içinde sunmadığımı, bu tezin planlanmasından yazımına kadar bütün safhalarda bilimsel etik kurallarına uygun olarak davrandığımı ve aksinin ortaya çıkması durumunda her türlü yasal sonucu kabul edeceğimi beyan ederim.

27/ 08 / 2024

Halil ÇELİKKAYA

ÖZET

PETROKİMYA ETİLEN FABRİKASINDA ISI ENTEGRASYONUN TEKNİK FİZİBİLİTESİ

ÇELİKKAYA, Halil

Yüksek Lisans Tezi, Kimya Mühendisliği Anabilim Dalı
Tez Danışmanı: Prof. Dr. Bikem ÖVEZ
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Plastikler, elyaf ve diğer kimyasalların üretiminde kullanılan en önemli temel hammadde olan hafif Olefinler; Etilen ve Propilendir. Endüstriyel anlamda hafif olefinlerin üretimi buharla hidrokarbonların (etan, propan, bütan, LPG, nafta ve gas oil vb.) ısıl parçalanmasıyla elde edilmektedir. 1940 'ların başında başlayan bu üretim teknolojisinde süregelen çeşitli ekonomik kısıtlar bulunmaktadır. Bu kısıtların başında gelen etkenlerden en belirgin olanı kullanılan hammadde (Etilen fabrikası için Nafta, LPG vb..) fiyatlarının dünya piyasasında talep arz dengesine bağlı olarak son çeyrekte yükselerek artmasıdır. Bu çalışmada ikincil maliyet kalemi olan ısı gereksinimi için pinch analizi dijital ikizine uygulanarak Etilen Fabrikası'ndaki darboğazların giderilmesi ve sürecin iyileştirilmesi hedeflenmiştir.

Yeni yapılanma, süreç içi eşleşmeler ve kullanılacak minimum sıcak ve soğuk utilite miktarları, minimum sıcaklık farkı (ΔT_{min}) 10 °C seçilerek oluşturulmuştur. Pinch analizi sonucunda yeni 2 farklı ısı değiştirgeç ağı tasarlanmıştır. Dizayn 1 ve 2' de sırasıyla total ısı değiştergeci sayısı 103 – 110, kapital gider 13.7 M$- 13.8 M$ olarak hesaplanmıştır Fakat bu yeni ağın uygulanabilirliği mevcut saha koşullarında maliyetli olacağından, uygun görülen iyileştirmeler paylaşılmıştır. Ayrıca endüstiride çok yaygın olan parçalama fırınlarında gaz türbini entgrasyonu çalışılmıştır. Gaz türbini entegrasyonu ile % 5' e kadar enerji tasarrufunun Olefin fabrikası dijital ikizinde gerçekleşebileceği öngörülmüştür.

Anahtar sözcükler: Etilen, propilen, olefin, süreç entegrasyonu, pinch analizi, dijital ikiz, simülasyon, Aspen Hysys, Aspen Energy Analyzer, Aspen Process Economic Analyzer, Aspen Capital Cost Estimator, gaz türbini.

ABSTRACT

TECHNICAL FEASIBILITY OF HEAT INTEGRATION FOR THE PETROCHEMICAL ETHYLENE PLANT

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Ethylene and Propylene, which are the valuable light olefins and the most important basic raw materials, are used for the production of plastics, fibres and other chemicals. The light olefins production is obtained by thermal decomposition of hydrocarbons such as ethane, propane, butane, LPG, naphtha and gas oil, etc. with the presence of steam in industrial sense. There are several economic constraints in continued production technology that began in the early 1940s. The most obvious factor of the constraints is the raw material price increment in the last quarter, depending on the demand and supply balance in the world hydrocarbon market. This study targets debottlenecking and retrofitting to digital twin of Olefin Plant by applying pinch analysis by reducing energy cost, which is second big constraint for sector.

While performing heat integration, safety and product quality requirements of the process are the first priority. Maximum heat recovery potential (or minimum energy requirement) in ethylene process is instituted by taking minimum allowable temperature difference (ΔT_{min}) as 10 °C. Result of pinch analysis, 2 different new heat exchanger networks are designed. Total shell 103 – 110, capital cost 13.7 m$-13.8 m$ are calculated in design 1 and 2 respectively. However, since the applicability of this new network would be costly in the current field conditions, the improvements deemed appropriate are shared. Also very common in industry, gas turbo generator (GTG) integration has been studied in cracking furnaces. With the integration of the GTG, up to 5 % energy saving is envisaged in DT of Olefin Plant.

Key words: Ethylene, propylene, olefin, HEN, MER, Umin design, process integration, pinch analysis, digital twin, simulation, Aspen Hysys, Aspen Energy Analyzer, Aspen Process Economic Analyzer, Aspen Capital Cost Estimator, gas turbo generator.

PREFACE

I studied Pinch Analysis of Olefin Plant for the Master of Science degree in Chemical Engineering at Ege University. My work was supported by Prof. Dr. Serap CESUR, Assoc. Prof. Dr. Zehra ÖZÇELİK, Dr. Ayça ATA, Dr. Duygu OVA ÖZCAN and Prof. Dr. Bikem ÖVEZ from Ege University and my colleagues.

In this study, we worked on what could be done by using Pinch Analysis without requiring further more investment. We realized that there could not be much return because plant has many revamps Olefin Plant that is well integrated on heat balance and there is no more space for extra heat exchangers. Therefore, we tried to offer what is being done in the new Olefin plants in the world and we tried to calculate their returns on investment with simulations and digital twins.

I hope you to enjoy reading this thesis. Study will guide you. I kindly request you to contact with me for further questions.

İZMİR

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Halil ÇELİKKAYA

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SYMBOLS AND ABBREVIATIONS

Symbols	Descriptions
A	Heat transfer area (m^2)
C_p	Specific heat capacity at constant pressure (kJ/kg°C)
h	Specific enthalpy (kJ/kg)
h_0	Specific enthalpy at reference conditions (kJ/kg)
H	Enthalpy (kW)
$\dot{m}$	Mass flow rate (kg/s)
N	Number of units
N_C	Number of cold streams
N_H	Number of hot streams
P	Pressure (kg/cm^2)
$\dot{Q}$	Heat load (kW)
$\dot{Q}_C$, $\dot{Q}_{CU}$	Heat load of cold utility (kW)
$\dot{Q}_{C,min}$	Minimum requirement of cold utility (kW)
$\dot{Q}_H$, $\dot{Q}_{HU}$	Heat load of hot utility (kW)
$\dot{Q}_{H,min}$	Minimum requirement of hot utility (kW)

SYMBOLS AND ABBREVIATIONS (continued)

Symbols	Descriptions
T	Temperature (°C)
T_0	Environment temperature (°C)
$T_{cold,pinch}$	Cold pinch temperature (°C)
$T_{hot,pinch}$	Hot pinch temperature (°C)
T_{pinch}	Pinch temperature (°C)
$\dot{W}$	Work (kW)
ΔH	Enthalpy change (kW)
ΔT	Temperature difference (°C)
ΔT_{min}	Minimum allowable temperature difference (°C)

SYMBOLS AND ABBREVIATIONS (continued)

Subscripts

c	Cold stream
h	Hot stream
in	Inlet
lm	Logaritmic
min	Minimum
out	Outlet

Greek letters

Δ	Difference

Abbreviations

C	Cold stream
CC	Composite curve
CCC	Cold composite curve
CU	Cold utility
GCC	Grand composite curve
H	Hot stream
HCC	Hot composite curve
HE	Heat exchanger

SYMBOLS AND ABBREVIATIONS (continued)

HEN	Heat exchanger network

Abbreviations (continued)

HU	Hot utility
MER	Maximum energy recovery
PA	Pinch analysis
PI	Process Integration

1. INTRODUCTION

Petrochemical sector is not the only sector that triggered the mobility of the hydrocarbon market. Today the result of human needs for energy, energy sector commitment to oil and derivatives is greater than the sector of petrochemicals. In addition to this, the oil crisis in the 1970s was accelerated the rise in all sectors-petroleum based raw material prices (Abbood et al., 2012). Nowadays the political dimension of the oil market has become the primary factor in the pricing stage of oil. Variable production costs are observed because of differences of natural resources distribution between the continents. For example, the North American continent, ethane and propane are processed as raw materials, whereas naphtha seems to be more common in the Middle East and in Europe.

In Middle East, competitors take an advantage of economies by building new mega crackers and these large-scale plants make better use of capital and resources. The other competitors except China and United States are forced to develop other tools by market economics trends. New tools, one of them is digitalization, should be used and tested in a fastest way. Nowadays these kind of tools provides analysing data and derives innovative solutions. Ethylene cracking furnace scheduling significantly impacts plant efficiency. To enhance ethylene production efficiency, this study addresses the scheduling challenges posed by variable cracking durations. Combining machine learning with a two stage optimization structure, the approach effectively manages uncertainties and optimizes production. The outer ring employs double stage programmed by stochastic way with scenario clustering, while the inner layer incorporates robust optimization. Real-world implementation in an ethylene cracking system significantly improved profit daily, demonstrating robust performance with 92.3% data coverage and a 0.826 change rate. Sensitivity analyses were conducted to refine model parameters. (Zhang and Wang, 2023)

The olefin production process uses large amount of energy in the process as required, in addition to the raw material welded constraint due to cost constraints that forces manufacturers. Therefore, resulting in too high production cost per unit of olefins decreases profitability. Today, heat integration of a process, which is one of the intense research activities, is intended to reduce the use of external resources, are called as utilities, to cool and to heat the process streams. Efforts of matching process own hot and cold streams for ensuring the required heat exchange and to reduce the use of utility is ongoing intensively upon request. To achieve energy sustainability and reduce carbon emissions, the process industry must optimize production and enhance energy efficiency. However, limited data hinders effective modelling and analysis. The study presents prediction of production and model of energy structure optimization to address challenges. The model combines a Long Short-Term Memory neural network (LSTM) with the Monte Carlo (MC) method (MC-LSTM). The MC method expands the limited real production data to train the LSTM for accurate production prediction. By analysing the model's results, inefficient samples can be identified and optimized for enhanced energy efficiency. The proposed MC-LSTM model was applied to an ethylene plant to predict production and optimize energy consumption. Results of experiments achieved an accuracy approximately 96.57%, surpassing other models. Additionally, the model identified opportunities for substantial energy savings, reaching nearly 13.22%. This approach offers a promising solution for process industries struggling with limited data to optimize production and energy efficiency, ultimately contributing to reduced carbon emissions. (Han et al., 2023)

A digital twin (DT) model consists of eight main stages as discussed in section 5. Main part of DT is classic process simulation. Overview of DT schematic is shared in Figure 1.1. Olefin plant simulation can be done easily by using any process simulation program. Some of simulation program is powerful than the others in some ways. Some of them has special models. Evaluation of process simulation program is not mentioned in this study. Aspen Hysys is chosen because main part of plant is ready for use. Utility side is integrated later. Finally, core of DT has all plant data from feed stage to products stages for purpose of describe plant behaviour well as operating conditions (Boschert et al.. 2018).

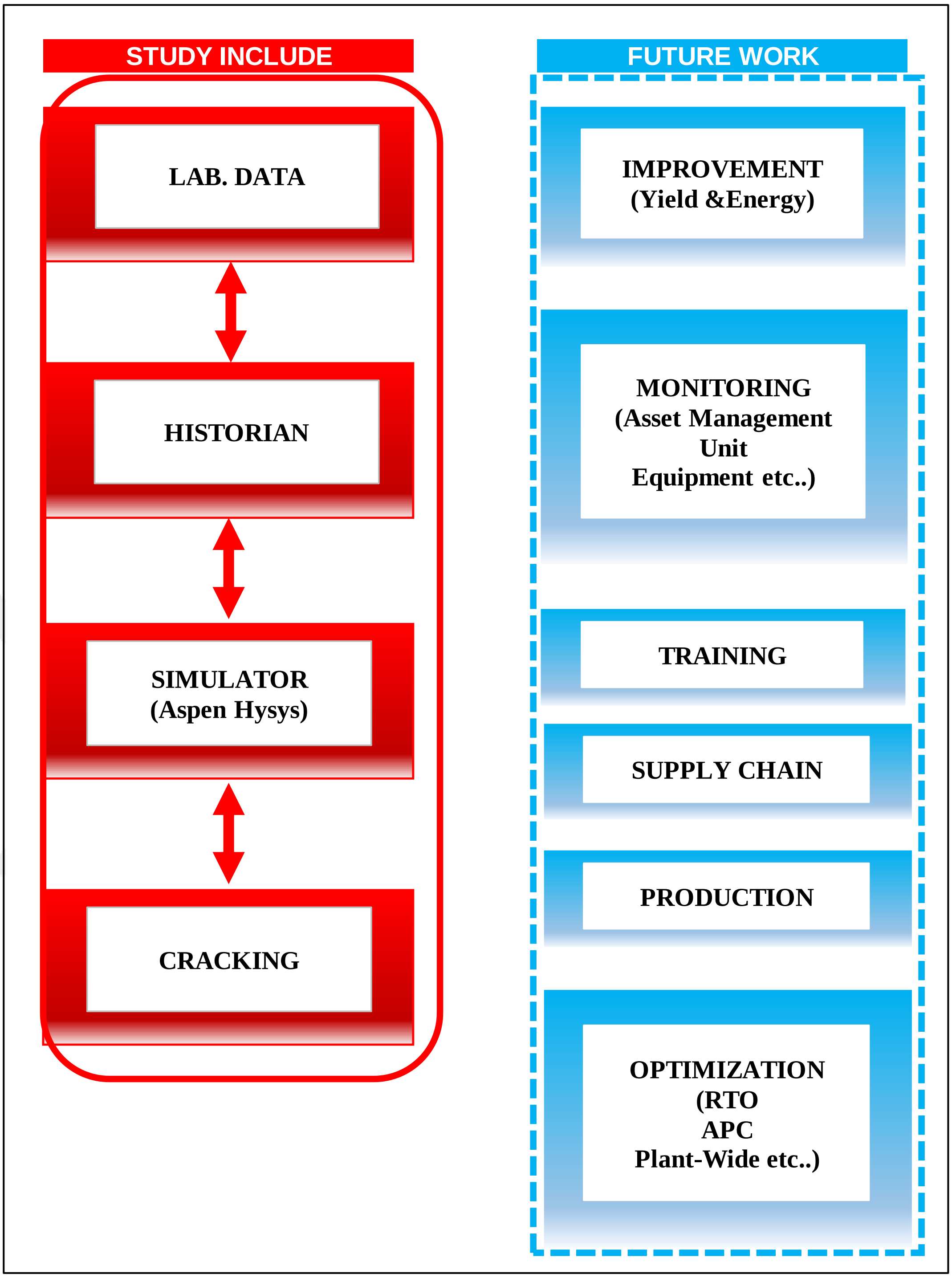


Figure 1.1 Digital Twin Schematic of Olefin Plant

To enhance energy efficiency in industrial and urban sectors, the study introduces ROSMOSE, a novel web-based tool for evaluating and comparing process integration options. Leveraging pinch analysis and optimization, ROSMOSE calculates energy requirements, identifies pinch points, and designs utility systems while minimizing costs or environmental impacts. Built on the

Quarto platform, ROSMOSE integrates various open-source tools for comprehensive analysis. A dairy case study demonstrates significant energy savings (up to 90%) and CO2 reductions (up to 96%) through optimized utility systems, heat pump integration, and renewable energy utilization. ROSMOSE empowers decision-makers to assess diverse system configurations and make informed choices for sustainable energy management. (Dardor et al., 2024)

Olefin Plant Process DT allowed Heat Exchanger Networks changes modelled and explored us improved techno-economic aspects of the Olefin plant operation, delivering improved profitability in terms of energy balance.

While performing heat integration, safety and product quality requirements of the process should be the first priority (Tuomaala et al., 2010). Maximum heat recovery potential (or minimum energy requirement) in Olefin process will be the first goal in this study. Followed by determination of process exchangers that violating pinch point which corresponding to the bottleneck of the process, the new line arrangements will be made (Akbarnia et al., 2009). Summarize, the new operating conditions at different levels with equipment changes in order to lower the energy cost that will not affect the product quality will be determined.

However, it is obviously shown that existing plant has an efficient heat integration after many revamps. Cold box system is well integrated in cold separation area. In hot area, quench oil, pan oil and quench water heat recovery circulations are reached almost their maximum limits. In addition, there is no more space for new exchangers on real site.

On the other hand, cracked gas outlet heat surplus in hot section can be useful for the entire plant reboilers, which are located in cold area, such as deethaniser, depropaniser and debutaniser. There are four restrictions on that idea. One is heating inlet feed of columns can reduce condenser’s performance, be careful about that outcome. Second is at the start-up and at the shutdown procedure there will be a process safety problem. Third concern is fouling problem can occur easily because of cracked gas impurities. The latest issue is a pressure drop (ΔP) problem due to the significant distance between the hot and cold sections in plant. As I mentioned there is no space in the middle of plant for extra heat exchangers, which are connected together with minimum pipe routes.

Dilution steam system has opportunity via Pinch Analysis to reduce Medium Pressure Steam (MPS). MPS can be used as extra heater at top of High Pressure Column of Dilution Steam System. Then this used MPS can be used as vaporizer heating media with MPS latent heat. In this application, also ΔP problem is big problem. Because if Dilution Steam (DS) pressure is decreased much more there will be safety problem for furnaces. If DS pressure is not increased, the problem may be faced. This is not a proper solution hence increasing DS system pressure will increase boiling temperature of liquid, then vaporizer first heating media quench oil circulation will be useless.

This study shows the potential for Pinch Analysis to save energy in olefin plants. Although it will not provide something new to the Pinch Technology, it will consist reasonable approach to reduce amount of energy requirement for current plant in a safe side (Beninca et al., 2011). At last, some process solutions such as increasing deaerator pressure for increasing Extra High Pressure Steam (XHS) production. Latest offer is that GTG installation for new furnaces causes reducing amount of NOx and increasing energy efficiency of furnaces.

Normally fuel gas enters radiation section of furnaces and it is burned there. Released heat is used for cracking reactions. In GTG case, large amount of fuel gas enters first GTG and outlet of exhaust gas has nearly 496 °C high preheated air as shown below Figure 1.2. This preheated air is fed to same radiation section. Thus, this preheated exhaust gas reduces the heaters' fuel requirements. On the other hand, oxygen content of GTG outlet is reduced to 15%. Although fuel gas consumption decreases nearly 2 t/h. Increasing the total flow of combustion air is necessary to maintain the same excess air to the furnace. Results in an increase in the total mass flow of stack gas passing through the furnaces. In turn, this increases the total steam production in convection. It is assumed that the power produced by the GTG is used to generate electrical energy for internal use. In addition, this electricity can be exported to other plants located in petrochemical complex easily. It is expected that there will be no ethylene loss in case of GTG upset. Furnaces can be switched into no GTG case easily.

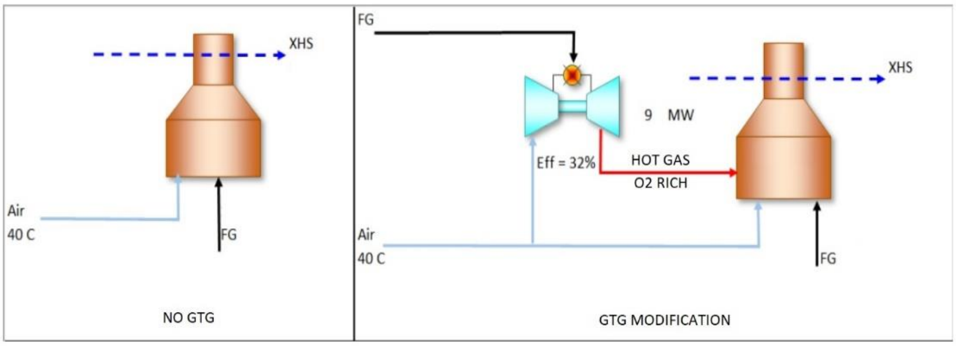


Figure 1.2 GTG Integration with Furnace

On the other hand, GTG capacity can be chosen higher capacity one up to burn all produced excess methane. Balance results of fuel gas is very important decision maker in this case. If excess fuel gas is burned in flare, this excess can be used in higher capacity GTG. In addition, heat recovery steam generation section can be added that new higher capacity GTG to produce extra steam for ethylene plant or petrochemical complex. In addition, decision makers should consider that excess steam would affect all steam balance of complex. By the way, capex will be increased with choosing higher capacity GTG.

Other good point of GTG is reducing emissions by decreasing NOx formation. Two main parameters affect NOx formation during combustion process. One of them is temperature and second is oxygen concentration of flame zone. Excess air ratio is the key to reduction of NOx. If oxygen content is higher than 8% then outlet of GTG NOx emission reduces (Tahery, 2014). Below Table 1.1 shows specifications of GTG.

Table 1.1 Specifications of GTG

Specification	**Efficiency**	**NOx emissions**	**Exhaust temperature**
Noval T12	35.3 (%)	15 (ppm)	496 (ºC)

All previous offers are simulated and optimized. The result is that there is too many constraints such an old plants. The new technological investment as gas turbine can be integrated to get better plant performance.

2. LITERATURE REVIEW

Nowadays chemical processes are very complicated. One of products are the other inputs. First one fluctuations are effecting the end one. Chemical engineers are developing tools to get efficient chemical processes. World has limited sources. So all chemical processes together must be focused as one. General optimisation must be done for total of processes. The holistic approach to chemical process for retrofitting and looking their optimising together would be numerical or graphical methods. In this field, common useful methods are process integration and pinch analysis method (Smith, 2005). Also in this study, these methods are used.

Also steam power plants constitute a significant portion of global electricity generation, necessitating performance optimization to mitigate environmental impacts and enhance economic viability. The study employs together with pinch and exergy analysis (CPEA) to evaluate in the power station under actual operating conditions. The Engineering Equation Solver (EES) software simulates the system components under varying operating parameters. Results indicate a 1% increase in high-pressure turbine network output due to pressure variations and a substantial 70% increase due to temperature fluctuations. (Radhi et al., 2024)

2.1 Process Integration

The oil crisis in the 1970s was accelerated the rise efficiency tools (Linhoff, 1998) such as process retrofitting (Kemp, 2007), emission reduction, minimizing water usage (El-Halwagi, 2006), hydrogen management (Friedler, 2010) and carbon footprint management recently. Figure 2.1 describes the relationships between the known methods of process integration.

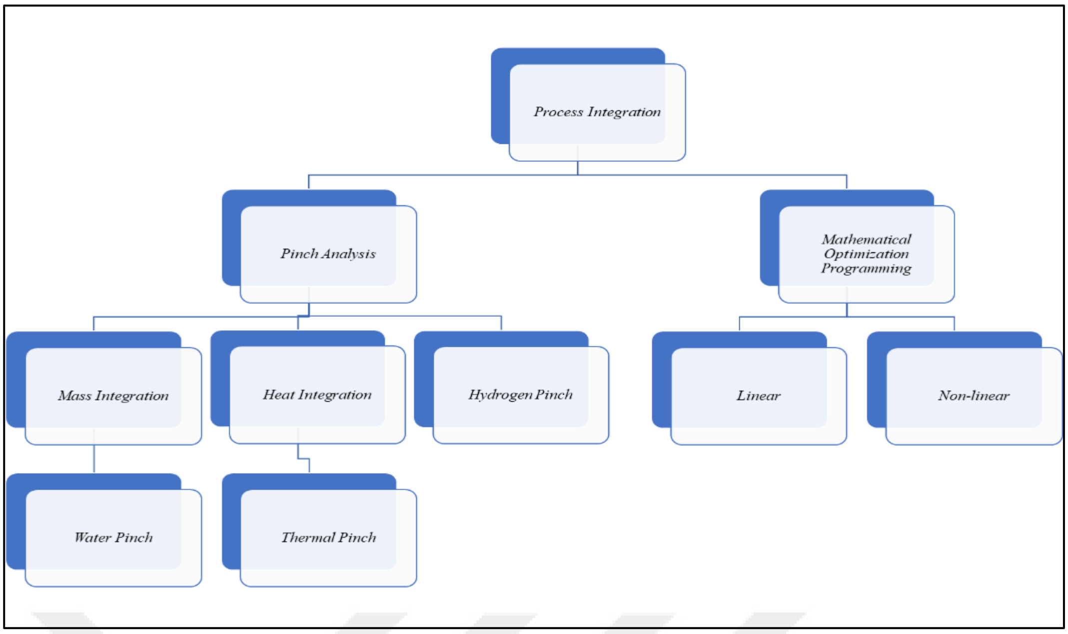


Figure 2.1 Simple Tree of Process Integration

Also as known, this tree has three basic concepts.

1. Analyse the processes, define all system units as process, utility and waste,
2. Synthesis the processes, apply thermodynamic, energy and mass balances for establish to performance targets,
3. Finalize and if required retrofit again details of process design to reach performance targets.

One of the goals is integration use of materials and energy then second is minimizing process output wastes and emissions for process integration.

2.2 Pinch Analysis

Pinch analysis has been a tool in industry for over 30 years, such as chemicals, petrochemicals, refinery, oil, paper, metallurgy, food and drink. Core of heat integrated pinch analysis has three targets;

1. Minimum unit quantity (exchangers),
2. Lowest equipment investment cost,

3. Minimum utility cost.

Alternatively, mass integrated pinch target is increasing mass transfer rate from rich process to lean process, so minimizing waste with using minimum utility. Figure 2.2 Pinch Proven Target Map for Processes shows minimum and maximum gain with applying pinch to different processes.

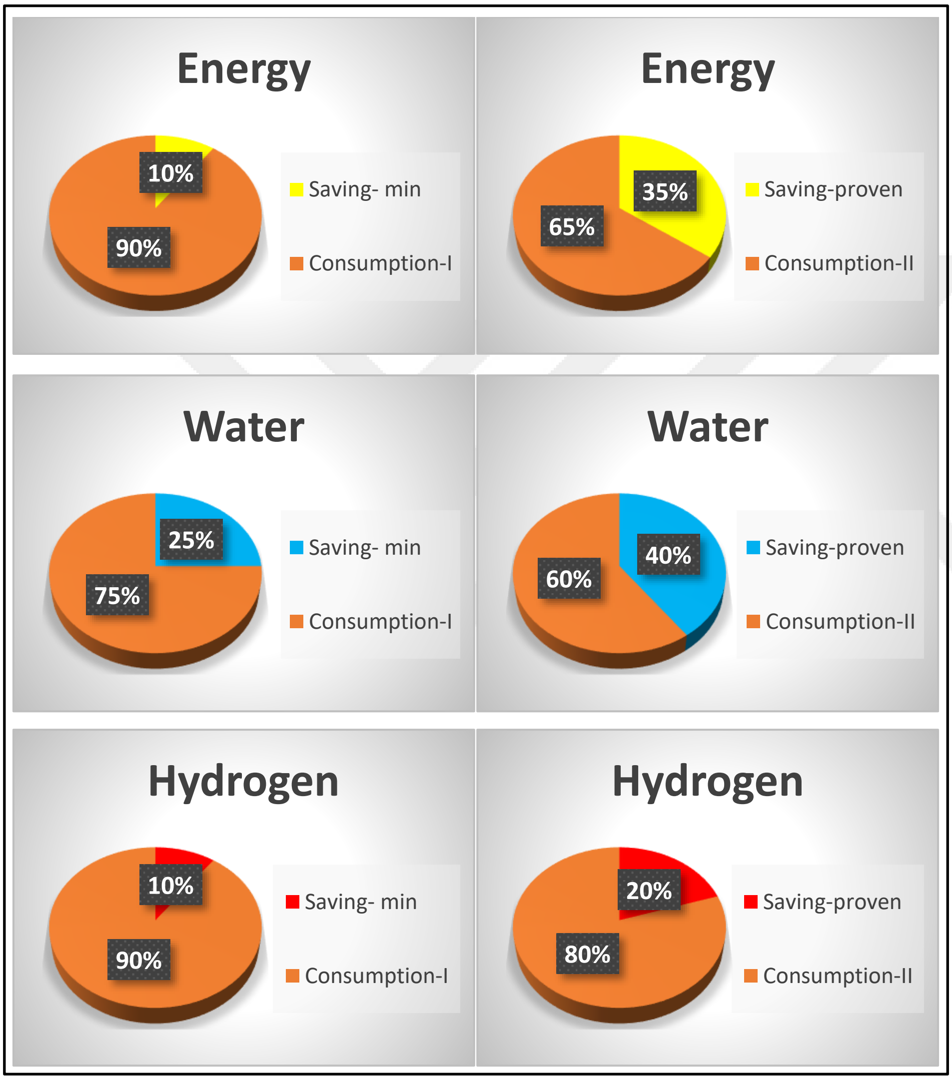


Figure 2.2 Pinch Proven Target Map for Processes (El-Halwagi, 2006)

The study of bio-based methyl ethyl ketone production evaluates the techno-economic feasibility of a biorefinery producing 2,3-butanediol (BDO) from sugarcane and subsequently upgrading it to methyl ethyl ketone (MEK). Three downstream MEK separation processes were compared: azeotropic distillation, extractive distillation of MEK, and extractive distillation of MEKisobutyraldehyde. Pinch technology integrated into the processes reduced BDO and MEK production costs by 8% and 9-10%, respectively. To enhance renewable MEK's competitiveness, future efforts should prioritize reducing feedstock costs, improving fermentation efficiency, and developing cost-effective downstream separation technologies. (Varma et al., 2024)

3. ETHYLENE PRODUCTION PROCESS

Olefin plant is analysed by pinch analysis. Improvement items of energy performance are identified through inspection of design data and process flow diagrams. Added benefits of some of pinch results modifications are that they also decrease plant energy demand. In this study GTG modification for new furnaces is worked deeply and output gain investigated for overall plant boundary (Tahery, 2014). The heart of naphtha steam cracker, also known as the furnace, accounts for nearly higher than 50% of the energy consumption against to total requirement. (Ren et al., 2006). GTG process solution is lowering energy demand of furnace.

Ethylene is the major component in Petro chemistry processes, demand drivers are mainly ethylene derivatives polymers. Every year demand is increasing 5 %. Ethylene is transparent, highly flammable, difficult to transport and store and very reactive.

Ethylene can be produced by using many technologies. If we route minor to important one, route starts with refinery cracking processes, methanol to propylene processes, methanol to olefin processes, biomass fermentation to ethanol then dehydratization to ethylene and the last dominating technology is steam cracking. The study proposes a novel, combined process for ethylene production that integrates electrified cracking and oxidative dehydrogenation (ODH). This approach offers a more economical and environmentally friendly alternative to conventional fuel-fired cracking. By replacing radiative heating with electrical heating or partial feed combustion (ODH), the process achieves higher thermal efficiency (97.1% and 98.1% for electrified cracking and ODH, respectively, compared to 89.9% for fuel-fired cracking) and significantly reduces CO2 emissions (55.4% and 49.5% for electrified cracking and ODH, respectively). Combining these technologies further enhances emission reduction (57.7%). Notably, the proposed process maintains economically competitive operating costs (0.84-0.94 € Nm^{-3} C2H4) compared to traditional methods. This research demonstrates a promising pathway for sustainable and cost-effective ethylene production with a significantly reduced carbon footprint. (Maporti et al., 2023)

Steam cracking means to all processes starts with liquid naphtha, gas LPG or gas ethane thermal cracking by using dilution steam (first step) and next step primary fractionation, at the same time compression (second step), follows product recovery section means separation processes (third step).

First step thermal cracking is called as pyrolysis. As shown in Figure 3.1 Naphtha, LPG or ethane is fed to the furnaces. All furnaces have economizer; here feedstock is heated by flue gas for getting more efficient process conditions. Hot feedstock enters convection section then slightly vaporized with superheated steam. Here steam is injected for two purposes, first is the fully vaporize feedstock and second is reduce to partial pressure of products. Then steam and feedstock mix goes to radiant section of furnace. Thermal cracking reactions occur in long (10-30 meters), narrow (30- 150 mm) single or multiple tubes which material is mostly iron and mixed nickel, chromium alloys. In the radiant section of the furnace, large molecules of feedstock are broken down by heat. The tubes in this section are heated to temperatures between 750 and 900 ^{0}C. The maximum allowable metal temperature is the limitation for heating and cracking large molecules bonds. If the outlet temperature increases, severity is increases too. The key performance indicator in this pyrolysis process is severity but not only coil outlet temperature affects yields. There is a lot of parameters such as retention time, feedstock character etc. (Karaba et al., 2012). Second is quenching hot gas for get rid of secondary reactions of free radicals which rearranges to complex molecules such as coke, mostly aromatics. Outlet of furnace hot gas mixture is quenched up to 350 ^{0}C in transfer line exchangers (TLE). TLEs are mostly two type heat exchangers (Yangkun et al., 2013). One is number of coil one tube in shell parallel heat exchangers, second is like one shell and multiple tubes heat exchangers. Heat removed from cracked gas generates high-pressure steam at the same time. This steam is used in plant for driving cracked gas and coolant utility compressors.

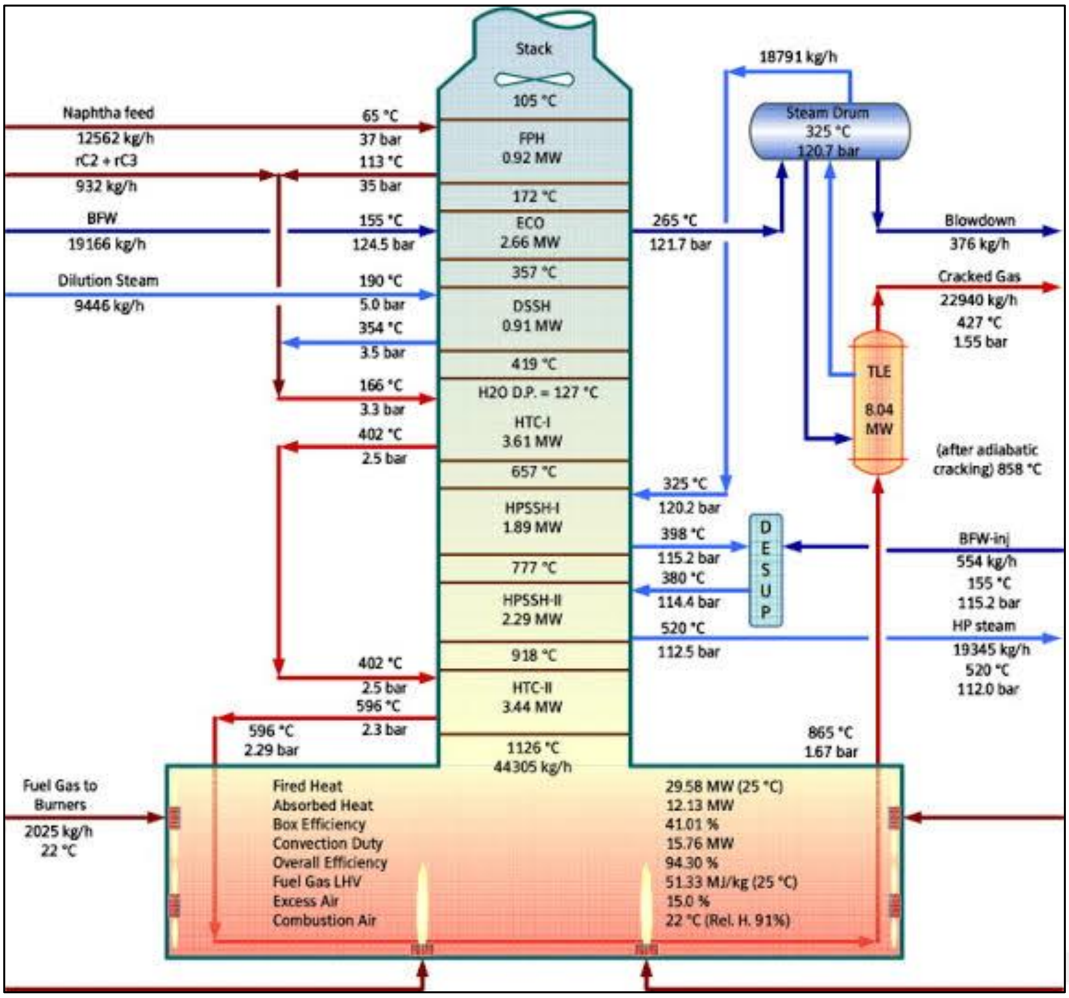


Figure 3.1 Steam Cracking Furnace Example (Ramin et al., 2009)

Second step in Olefin plant is called hot section as primary fractionation, at the same time compression. TLEs outlet goes to primary fractionator column. Mostly liquid feedstocks have fuel oil content after cracking. Fuel oil is condensed and taken as a heaviest product. Second heaviest product pyrolysis gasoline (C5-C12) is taken as a feedstock of aromatic plant. A few of pyrolysis gasoline goes to other section end. In addition, bulk content of C3, C4 mixture must be removed in condensate stripper for energy efficiency, proper use of cold utility in cold section. Secondary oil circulations recover heat. Dilution steam condenses to water and recycled to dilution steam generator system. Cracked gas is pressurized and cleaned to remove hydrogen sulphide (H_2S) and carbon dioxide (CO_2) acid gases.

Above CO_2 content is not related to net-zero CO_2 emissions industry transition. To accelerate the chemical's transition to zero CO2 values, the study investigates hybrid distillation-pervaporation processes for efficiently separating ethanol and isobutanol mixtures with water. Three process configurations, incorporating heat integration and renewable energy, were evaluated using Life Cycle Assessment, PESTLE analysis, and Multi-Criteria Decision Analysis. The D+PV+D configuration with integration of heat emerged as optimal for isobutanol-water separation, reducing CO2 emissions, environmental impacts, and total annual cost by up to 40% comparison with D+PV. Incorporating sources with renewable energy amplified the sustainable impact. This research highlights the potential of hybrid processes and sustainable energy integration for a greener chemical industry. (Thi and Toth, 2024)

Reducing carbon emissions from large ethylene plants requires cost-effective capture technologies. In the other study, heat of stack gas recovery (FHR) was investigated as a strategy to minimise energy consumption and capture costs in piperazine (PZ)-based post-combustion carbon capture (PCC) processes for 60,000 tonnes/year ethylene plants. Aspen Plus® simulations of four PCC configurations with and without FHR revealed significant improvements. The advanced flash stripper (AFS) configuration with FHR and absorber intercooling reduced the minimum regeneration energy requirement to 2.28 GJ/tonne CO2, a 37.71% improvement over the standard MEA solvent process. Economic analysis using Aspen Process Economic Analyzer® (APEA) revealed a 36.76% decrease in carbon capture costs for the FHR and inter-cooled AFS configuration, reaching \$47.27/tonne CO2. The study emphasises that FHR is a promising technology to enhance feasibility of techno economics for mega carbon capture in ethylene plants. (Ma et al., 2024)

These acid gases are poison for product recovery section in C2 acetylene and C3 methyl acetylene (MA), propadiene, (PD) hydrogenation reactors. At the end of this section all water content must be removed for proper work in next section. If there is any water and CO_2 then they can blockage cold box. Generally, main problem with compression section is fouling with polymers by the reason of increasing temperature after pressurization. This polymerization results energy losses and further step mechanical problems occurs vibration and finally like as thrust bearing damage. Majority boiling feed water and rarely wash oil (aromatic content high, non-diene C9 components) is atomized in the casing of cracked gas compressor to each section, mostly first. Simple Process Flow Diagram of Hot Section for Olefin Plant is given in Figure 3.2.

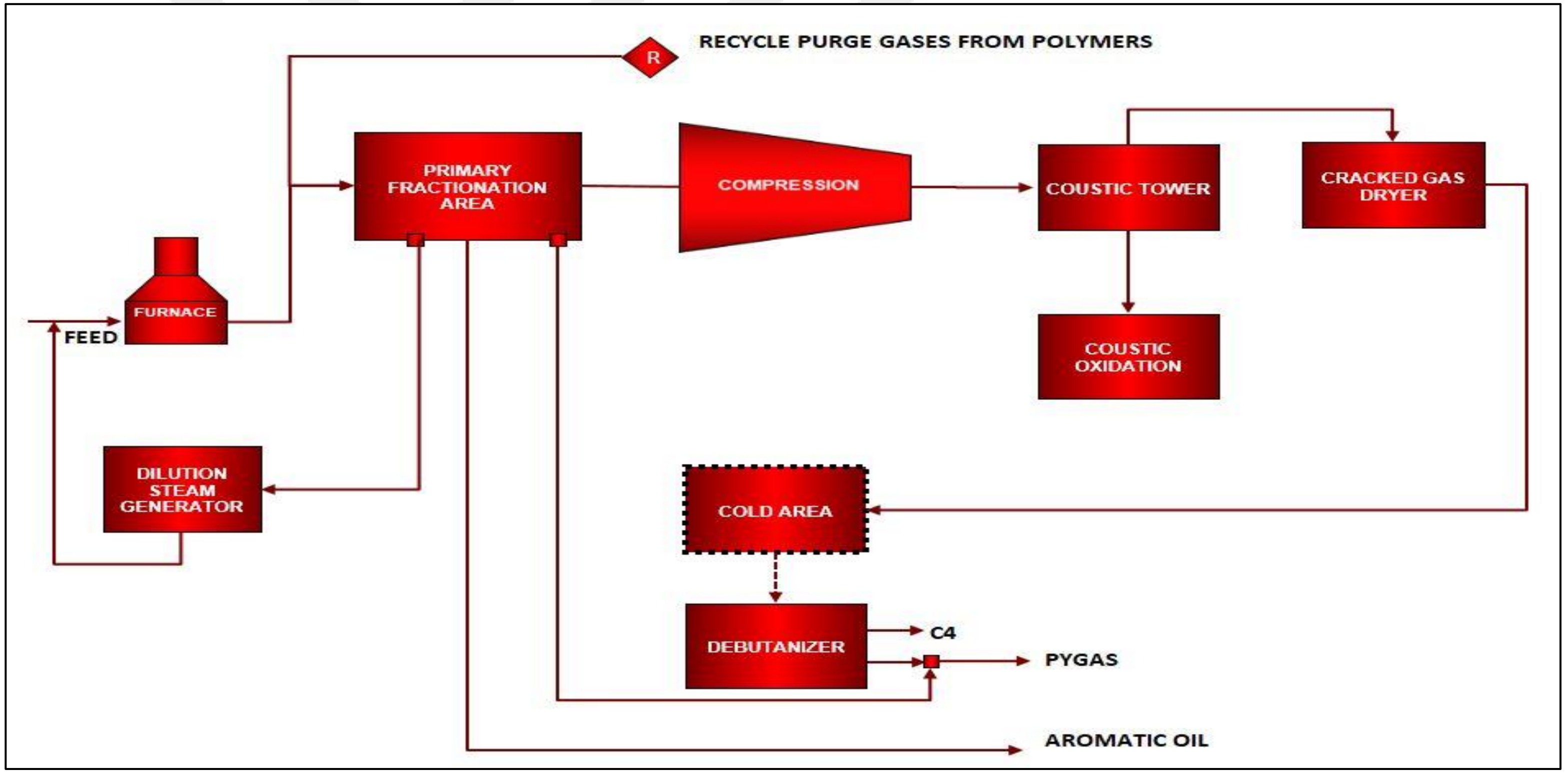


Figure 3.2 Simple Process Flow Diagram of Hot Section for Olefin Plant

Third step product recovery section is called cold section means refrigeration and extraction by flashing and distillation processes. Commonly hydrogen is produced first in this section. There are 3-separation sequence shown below in Figure 3.3, Figure 3.4 and Figure 3.5.

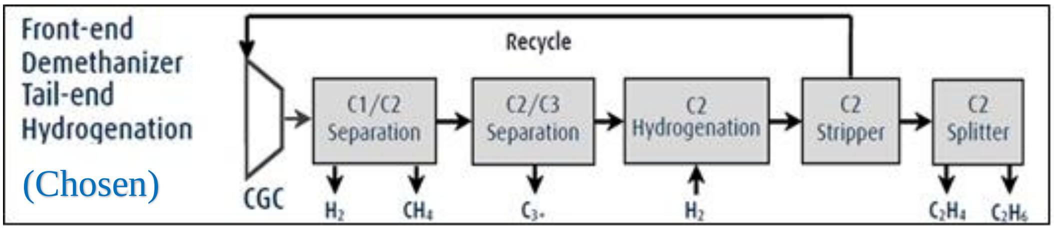


Figure 3.3 Cold Section Separation Sequence of Plant

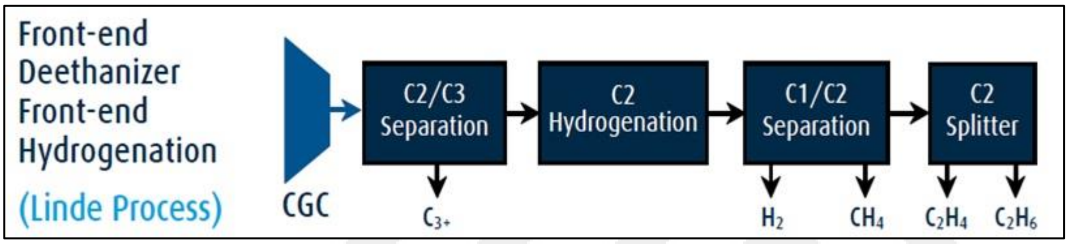


Figure 3.4 Cold Section Separation Sequence of LINDE

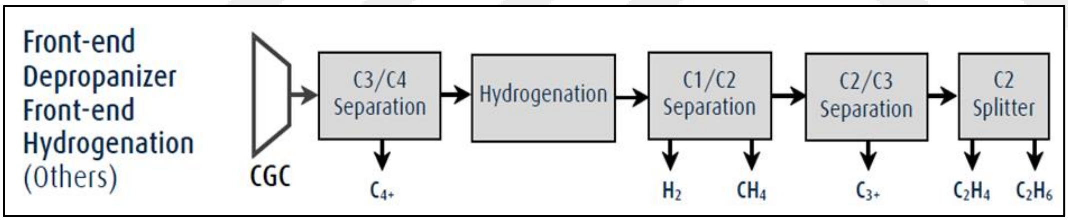


Figure 3.5 Cold Section Separation Sequence of Others

Separation sequence analysis identified best economics for front- end deethaniser layout, why the reasons are removal of heavy hydrocarbons upstream of cold train, acetylene hydrogenation in presence of hydrogen and reduced investment cost open loop ethylene refrigeration cycle (Van Geem and Marin, 2010).

In the case of front-end hydrogenation, acetylene and MA/PD are selectively hydrogenated first. The hydrogen is then removed from the hydrocarbon stream. In back-end hydrogenation, the selective hydrogenation of acetylene and MA/PD is performed at the end of the distillation train, after the hydrogen has been removed from the hydrocarbon stream. This requires two smaller reactors. Hydrogen must be supplied to support the hydrogenation reaction. Simulations show that the temperature in the overhead condenser of the demethaniser is 10°C higher in the front-end configuration, which reduces the load on the cold section. It is assumed that all the diolefins in the cracked gas are selectively hydrogenated to the corresponding mono-olefins. The energy demands of the different cold sections show that the low compressor duty required in the refrigeration section is high energy efficiency for demethaniser first, front-end hydrogenation design (Razib et al., 2012). Front-end hydrogenation design can be up to 20% lower duty than others. The design likely involves "front-end hydrogenation," a process where hydrogen is added to a starting material early in the production chain. This early addition of hydrogen might influence the final product composition, making it unsuitable for situations where C4 diolefins are desirable.

Mostly two refrigeration cycle is used for cooling requirement of cracked gaslights such as C1, C2, C3 and C4’s compounds. Cooling demand from 0 ^{0}C to -150 ^{0}C can easily supply by propylene and ethylene closed loop. These sub-systems performances are other issue for Olefin Plants. Turbines are producing work how it is within limit of sufficient vapour amount, or not (Becker and Maréchal, 2012).

In addition, demethaniser top outlet depressurized by expander and cooled stream is used as a coolant for cracked gas at the end of cold box again pressurized by compressor which is using same shaft with expander. Turbo- expander is a high valuable equipment for cold area in Olefin plant. It helps to reduce cycle duty of ethylene refrigeration (Becker and Maréchal, 2012). If fluctuation has not occurred and feed gas is clean, then turbo-expander turning up to 30000 rpm can operate without vibration, etc.

Figure 3.6 shows Simple Process Flow Diagram of Olefin Plant.

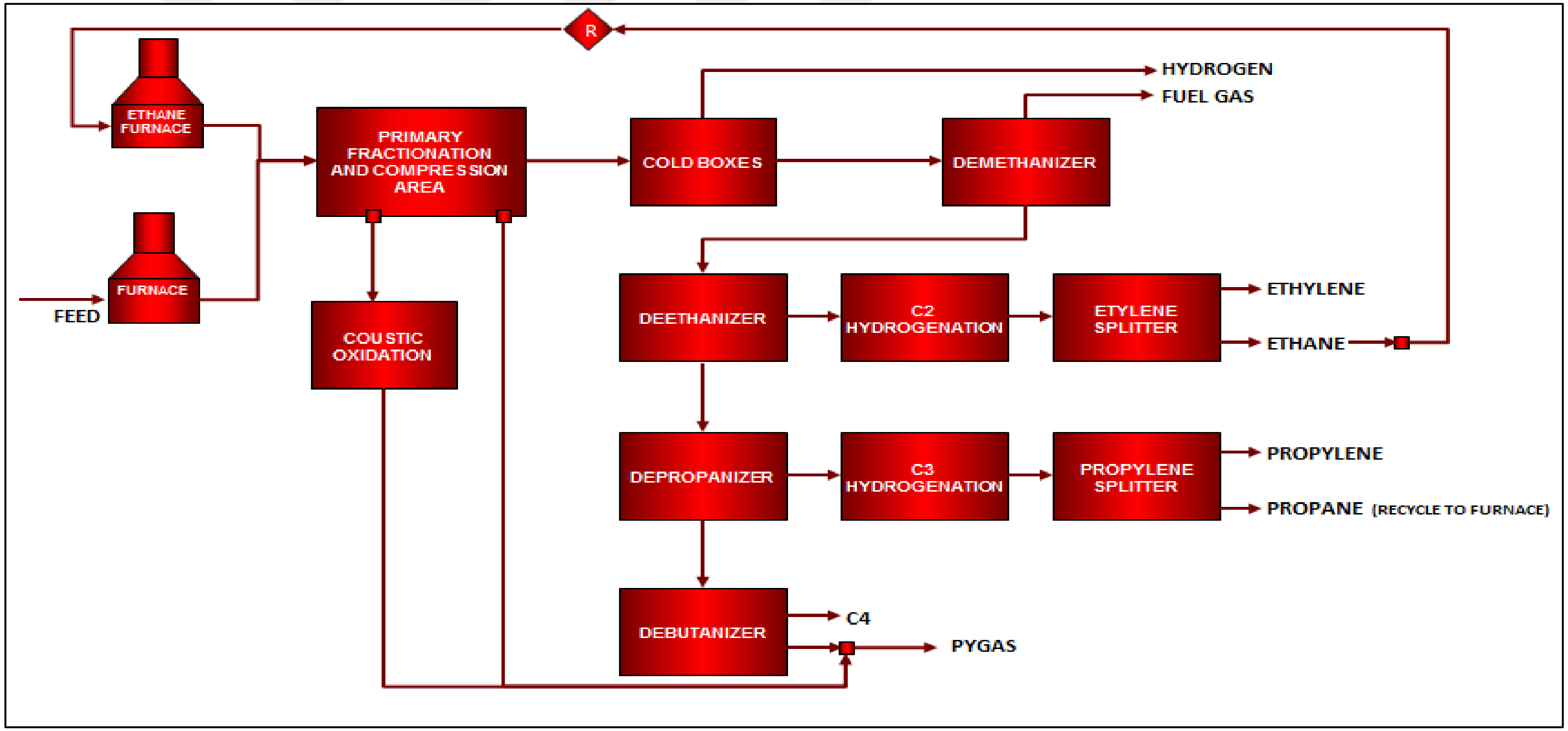


Figure 3.6 Simple Process Flow Diagram of Front- End Demethaniser Tail- End Hydrogenation Olefin Plant

There are few licensors in world for steam cracking. Table 3.1 shows four main licensors in world.

Table 3.1 Main Ethylene Licensors

Technip / Stone & Webster (PETKIM Ethylene Plant Licensor) https://www.technipfmc.com/en/about-us	Technip take it further.
KBR (Kellogg Brown & Root) https://www.kbr.com/en/about-us	KBR
CB&I Lummus http://www.lummusonline.com/global/index.htm	CB&I Lummus
LINDE https://www.linde.com/en/about-linde	THE LINDE GROUP

TECHNIP; https://www.technipfmc.com/en/about-us (Access date: 1 January 2019) Technip is a global leader in ethylene process technology with a combined experience of 110 years from Stone & Webster and Technip. The company has successfully engineered and implemented over 150 grassroots plants, producing more than 20 million metric tons per annum of ethylene through 18 steam cracking units since 2000. Technip's technology licenses account for approximately 50% of the world's total added ethylene capacity.

KBR (Kellogg Brown & Root); https://www.kbr.com/en/about-us (Access date: 1 January 2019) KBR is a global leader in the design, construction, and technology development of olefin plants. KBR has over 50 years of experience in this field, and has brought new ethylene plants with a capacity of 13 million metric tons per year online since 1990. KBR's olefins technology portfolio includes two technologies, SCORE™ and K-COT™, which can be used to produce C2, C3 and other heavies products from different type of feedstocks. The optimum selection of technology depends on the available feedstocks and the desired olefin product mix.

CB&I Lummus; http://www.lummusonline.com/global/index.htm (Access date: 1 January 2019) KBR has been a leading provider of specialty engineering, procurement, and construction (EPC) services to the petrochemical industry for over 80 years.

Linde's Engineering Division https://www.linde.com/en/about-linde (Access date: 1 January 2019) Linde is a global leader in the design, engineering, construction, and operation of industrial gas plants. The company has market segments such as olefins, production of natural gas, separation of air, hydrogen. In contrast to most of its competitors, Linde has its own extensive process engineering knowledge, which allows it to provide turnkey solutions for its customers.

These developer companies have opportunities and disadvantages relative to each other. Energy consumption of per ton ethylene produced is lower nearly 14.3 % in case of Gas Turbine Generator installed against to no GTG case. Brief advantages and disadvantages are listed in Table 3.2. As Linde process, waste heat is recovered. The study investigated the potential of supercritical carbon dioxide (sCO_2) Rankine cycles for waste heat recovery from gas turbines. Through comparative analysis and optimisation of simple, staged and split cycle configurations, the split cycle emerged as superior in terms of power generation and overall efficiency. A multi-objective optimisation approach combining genetic algorithms and machine learning can be used to determine the optimum cycle parameters. The optimised discrete cycle showed significant improvements in power output, heat recovery and overall system efficiency. The results highlight the potential of sCO_2 Rankine cycles to improve energy efficiency and reduce greenhouse gas emissions in waste heat recovery applications. (Turja et al., 2024)

Table 3.2 Brief Comparison of Main Licensors in World (Hydrocarbon-Processing. Petrochemical Processes 2003)

Licencor	***Technip S.A.***	***Lummus Technology Inc.***	***Linde A.G.***	***Stone & Webster (Purchased by Technip)***	***KBR***
Pyrolysis (Cracking) Furnace	*– SMK* gas feed to furnace coils With very large Capacity furnace – GK6* liquid feed enables a 35% yield increase*	*Feed preheated and cracked with steam in a short residence time (SRT)*	*Pyro Crack furnaces are optimizdesigned with lower residence time, temperature, and Low outlet pressure for feedstock to achieve high olefin yields*	*Ultra selective cracking (USC*) Cracking furnaces adopted within the specific feedstock and product requirements*	*– KBR combined with ExxonMobil's olefin technology – SCORE straight tube SC-design conditions to produce high olefin yields*
Licencor	***Technip S.A.***	***Lummus Technology Inc.***	***Linde A.G.***	***Stone & Webster (Purchased by Technip)***	***KBR***
Other Steam Cracker Components (than Furnace)	*Not available*	*Single refrigeration system with metathesis to increase propylene output catalytically with 15% less energy consumed*	*– Waste heat is recovered to boilers and process heaters – Process is optimized with flexibility and long intervals between major turnarounds*	*ARS technology for reduced chilling train requirements, reduced methane, partial de-ethanizer bypassing, dual feed ethylene fractionator, and reduced refrigeration demand (ca. 75%)*	*Various flow schemes based on de-ethanizer first, de- propanizer-first, and demethanizer-first configurations*

Energy Efficiency and Yield	*NA*	*– 3,300 kcal/kg for ethane cracking – 5,000 kcal/kg for naphtha cracker*	*– Ethylene yields are 25%, 35%, 45%, and 83% for gas oils, naphtha, LPG, and ethane, respectively – Specific energy consumption is ca. 6,000/5,400/4,600 and 3,800 kcal/kg of ethylene*	*– Ethylene yields from 57% (ethane, high conversion) to 28% (heavy gas oils). – Specific energy consumptions range from 3,000 kcal/kg to 6,000 kcal/kg*	*Ethylene yields up to 84% for ethane, 38% for naphtha, and 32% for gas oils.*
Gas Turbine	*NA*	*3 GJ/t of ethylene saved*	*NA*	*NA*	*NA*
Ethylene Yield, wt%	*35%*	*34.40%*	*35%*	*NA*	*38%*
SEC, GJ/t of Ethylene	*18.8 (best) or 21.6 to 25.2 (typical)*	*18 (with gas turbine) 21 (typical)*	*21 (best)*	*20 to 25*	*No data*
Licencor	***Technip S.A.***	***Lummus Technology Inc.***	***Linde A.G.***	***Stone & Webster (Purchased by Technip)***	***KBR***
Global Installed Capacity and/or Market Share	*– Production capacity the last 10 years of 7 mty – 25% the last 10 years*	*~ 40% of the world's ethylene plants*	*Over 15 million tons of ethylene produced in more than 40 plants worldwide*	*– Over 120 ethylene units built – Expansion techniques based on ARS technology have increased original capacities by over 100%*	*Collaborated in over 140 ethylene projects with capacities up to 1.3 mty since 1990*

4. EVALUATION OF ETHYLENE PRODUCTION PROCESS BY PINCH ANALYSIS

This study final aim is to optimize an olefin plant by identifying bottlenecks, implementing process improvements, and exploring the integration of novel technologies. Pinch analysis will be employed to enhance energy efficiency by maximizing heat recovery and minimizing external energy consumption.

4.1 Application of Pinch Analysis on Olefin Plant

The path followed in the using of pinch analysis is given below:

Data Extraction: Olefin plant hot and cold streams were extracted from DT of plant. Considering the process flow sheets, the problem table was created as shown in Table 4.1. The terms “Hot streams” (i.e. heat sources) and "cold streams" (i.e. heat sinks) are relative, and the temperature at which a stream is considered "hot" or "cold" will vary depending on the specific process. Hot streams typically requires to be cooled before they sould be processed further, while cold streams requires to be heated before they can be used. Supply temperatures are represented by Ts, while target temperatures are designated as Tt.

Data for pinch analysis was extracted from the olefin plant's process flow diagrams. A problem table was constructed, categorizing process streams as hot or cold based on their temperature relative to the desired process conditions. Among the 49 identified hot and 34 cold streams, only two potential heat exchanges were found within the process itself.

Essential stream data, including mass flow rates, inlet and outlet temperatures, were collected. Existing utility streams, such as cooling water and steam, were also characterized. Heat loads for all streams were calculated using ASPEN HYSYS based on stream properties and entered into ASPEN Energy Analyzer for further analysis. The resulting heat capacity flow rates and enthalpy changes are presented in Table 4.1. Heat capacity flow rate (GJ/ ºC h) and enthalpy change per unit temperature (GJ/h) all calculated values for utilities are shown in Table 4.1.

Heat capacity flow rates for streams undergoing no phase change were calculated by multiplying mass flow rates with specific heat values obtained from ASPEN Hysys, as outlined in Equation 4.1:

$$\mathrm{CP} = \dot{\mathrm{m}}\mathrm{Cp} \qquad (\mathit{Equation}\ 4.1)$$

Due to the significant impact of latent heat on process streams undergoing phase changes, CP values for these streams were determined using Equation 4.2:

$$\mathrm{MCP} = \frac{\dot{\mathrm{Q}}}{\Delta \mathrm{T}} \qquad (\mathit{Equation}\ 4.2)$$

The heat load of each stream was determined by calculating the difference between inlet and outlet enthalpy values, as obtained from ASPEN Hysys, using Equation 4.3:

$$\dot{\mathrm{Q}} = \Delta\mathrm{H} = (\mathrm{H}_{\mathrm{in}} - \mathrm{H}_{\mathrm{out}})\dot{\mathrm{m}} \qquad (\mathit{Equation}\ 4.3)$$

During the calculation of heat loads, two assumptions were made. To illustrate;

1. First condition is no restriction for location of new exchangers in plant

2. Utilities, propylene and ethylene system heat loads were closed systems will be operated as real state; there is no change for optimisation of closed loop cooling media. Therefore, the heat loads of the respective process streams were retained as initially calculated. (Bakhtiari and Bedard, 2013).

Having calculated heat capacity and loads of heat for both process with utility streams, the subsequent heat exchanger network design focused solely on process streams to minimize utility consumption.

Choosing Minimum Allowable Temperature Difference: Following the calculation of heat capacity flow rates CP and $\dot{Q}$, the minimum allowable temperature difference, ΔTmin, was determined. Selection of ΔT_{min} is very important and may optimization be done for reducing CAPEX and OPEX (Figure 6.10). ΔTmin represents the smallest temperature difference allowed between hot and cold streams within a heat exchanger. Increasing ΔTmin leads to higher hot and cold utility demands, as more heat will need to be transferred to attain the required temperature difference. However, a lower ΔT_{min} will result in a larger and more costly heat exchanger, as a larger heat transfer surface area is required.

Accurate heat exchanger (HE) modeling is crucial for efficient system optimization. While pinch point temperature difference (ΔTpinch) methods are commonly used, their inherent inaccuracies can hinder optimization algorithms. The study investigates the performance of high-order interpolation methods for calculating ΔTpinch in HE system optimization. Through case studies of heat pumps with varying complexity, we demonstrate that gradient-based optimization algorithms outperform particle swarm and genetic algorithms. Crucially, employing hybrid high and low-order interpolation methods for ΔTpinch calculation significantly accelerates optimization by a percent of 5-10 against to conventional methods. (Brodal et al., 2024)

The study investigates the optimization of heat pump integration in industrial processes to enhance energy efficiency and decarbonization. A novel methodology combining Process Change Analysis (PCA) and exergy analysis is developed to identify process modifications that maximize heat pump performance. The approach is demonstrated in biodiesel and vinyl chloride monomer purification case studies. Results indicate that strategic process changes, such as replacing wet water washers with membrane separation units, can significantly improve heat pump efficiency and reduce energy consumption. The proposed framework provides a valuable tool for optimizing heat pump integration in industrial settings, leading to substantial energy savings and reduced greenhouse gas emissions. (Raad et al., 2024)

The equation for calculating the required heat exchange surface area is presented in Equation 4.4:

$$A = \frac{\dot{Q}}{\Delta T_{lm}} \qquad (Equation\ 4.4)$$

So if ΔT_{lm} approaches to zero, means the surface increases to infinite and increases the capital cost too.

In petrochemical and refinery processes, ΔT_{min} is generally choosen between 10 ºC and 20 ºC (Figure 6.10). In this study ΔT_{min} is choosed as 10 ºC.

Algorithm: The cascade algorithm involves adjusting hot and cold stream temperatures to achieve the minimum approach temperature (ΔTmin). For instance, temperatures are shifted by ΔTmin/2 in opposite directions. These adjusted temperatures are then used to calculate enthalpy differences between hot and cold streams.

According to equation 4.5, ΔHCFR is calculated as:

$$\Delta HCFR = \Sigma CP_{cold} - \Sigma CP_{hot} \qquad (Equation\ 4.5)$$

A negative ΔH indicates a heat deficit, while a positive value represents a heat surplus. The smallest absolute ΔH value is identified, and its corresponding heat load is distributed across all temperature intervals. Once the cascade diagram is completed, a ΔH value of zero signifies the pinch point, where no heat transfer occurs.

4.2 Application of Pinch Analysis on Olefin Plant by Using Aspen Energy Analyzer

All data of streams except cracked gas compressor interim hot streams are added to simulation Heat Integration Manager Tab. In addition, utilities are added as shown below Table 4.1. After DT_{min} is entered, recommended design tab is activated. Automated New Design feature generates optimum designs using three steps.

Step 1: A linear programming (LP) model is used to screen out very poor matches. This model simultaneously optimizes the heat exchange area and heat load for each utility, and it is used to eliminate matches that are not likely to be economically feasible.

Step 2: A mixed-integer linear programming (MILP) model is employed to optimize heat exchanger network design. By building upon initial near-optimal matches and linearizing the log-mean temperature difference (LMTD), the model effectively minimizes annual costs while ensuring heat balance. The model identifies optimal heat exchanger pairings and utility loads, leading to improved network performance.

Step 3: Another MILP is formulated that identifies a heat exchanger network which satisfies the optimal heat load distribution determined from the initial model's solution. This model is solved iteratively, and it is terminated when the maximum number of designs is reached. The model results are visually represented through Composite Curves (Figure 4.1) and the Grand Composite Curve (Figure 4.2). Additional findings are detailed in the RESULTS and DISCUSSION part of the report.

Table.4.1 Aspen Energy Analyzer Utility Table

HI Project: Case 1 (Utility Streams)

Name		Inlet T [C]	Outlet T [C]	Cost Index [Cost/kJ]	Segm.	HTC [kJ/h-m2-C]	Target Load [kJ/h]	Effective Cp [kJ/kg-C]	Target FlowRate [kg/h]	DT Cont. [C]
LP		148.0	147.0	2.200e-006		21600.00	0.0000	1981	0.00	Global
MP		250.0	249.0	2.500e-006		21600.00	2.438e+008	1703	143131.76	Global
Cooling Water		26.00	32.00	2.125e-007		13500.00	6.716e+007	4.183	2675825.24	Global
REE-48		-48.00	-47.00	3.364e-006		4680.00	1.240e+008	1.341	92432373.12	Global
REE-69		-69.00	-68.00	5.876e-006		4680.00	1.050e+007	1.341	7829045.82	Global
REE-101		-102.0	-100.0	8.531e-006		4680.00	4.726e+006	1.341	1761963.82	Global
D306T-H2		-172.0	-161.0	1.890e-005		4680.00	1.062e+006	1.341	71981.25	Global
REP+23		23.00	24.00	2.739e-006		4680.00	0.0000	4.000	0.00	Global
REP+6		6.000	7.000	2.839e-006		4680.00	0.0000	4.000	0.00	Global
REP-6		-6.000	-5.000	2.990e-006		4680.00	0.0000	4.000	0.00	Global
REP-19		-19.00	-18.00	3.100e-006		4680.00	0.0000	4.000	0.00	Global
REP-34		-34.00	-33.00	3.364e-006		4680.00	8.099e+006	1.341	6039755.82	Global
Air		36.00	40.00	1.000e-009		399.60	5.598e+008	1.000	139938749.7C	Global
XHS		478.0	476.0	3.500e-006		21600.00	0.0000	1703	0.00	Global
<empty>										

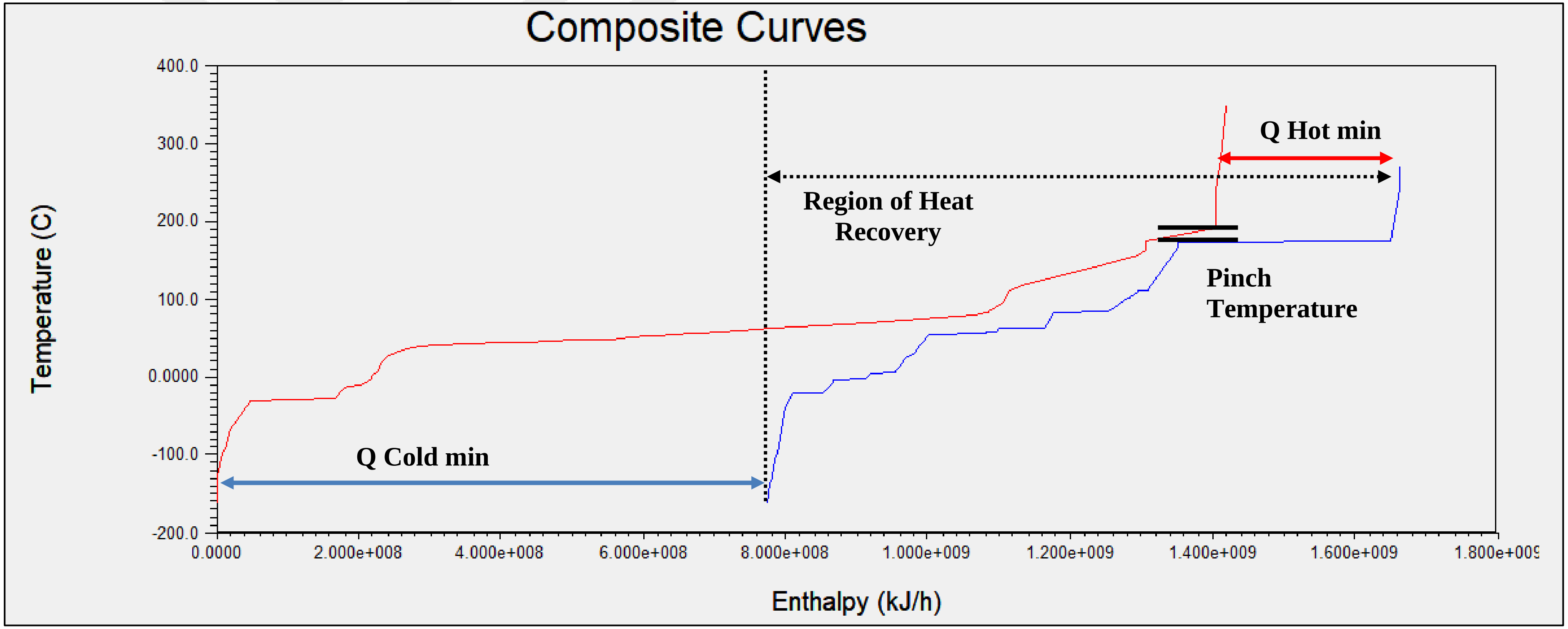


Figure 4.1 Hot & Cold Composite Curves

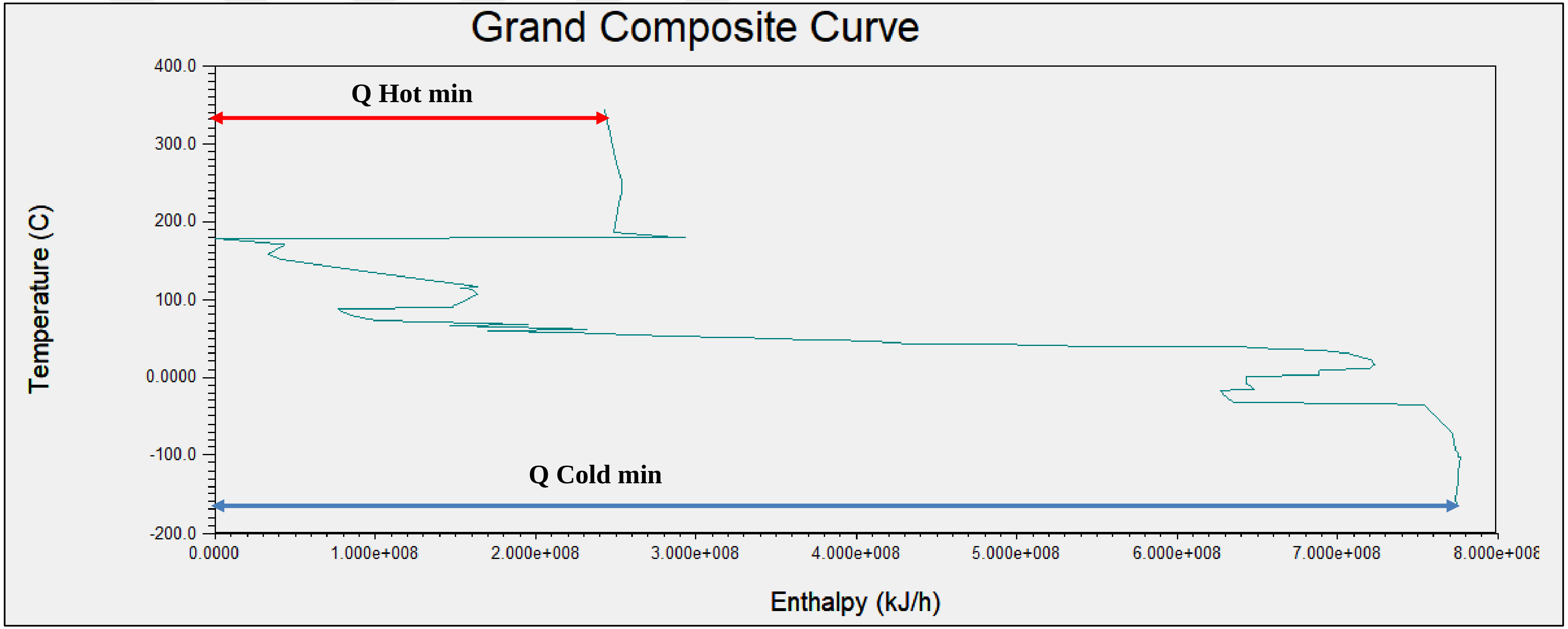


Figure 4.2 Grand Composite Curve

5. DIGITAL TWIN of OLEFIN PLANT

DT of olefin plant is created by integrated combination of many softwares. With the help of DT, all possible HEN scenarios are easily tested for the real site without any modifications. Thus, the plant continues to operate without any shut down. It is not recommend stopping the plant and applying the all results of the study by one by. There will be a cost of about between 5000 $ and 7500 $ per ton capacity of ethylene due to any plant shut down or upset position. The DT allowed us to determine forecasting with a question “What next?”, prediction with a question “What if?” and optimisation with a question “What’s best?” scenarios applying with pinch analyse that maximize profitability, enabling study to make better decisions without facing any shut down position.

The DT model consists of these following stages;

1. Feed Properties Collection (Sample Manager)
2. Process History Data (Process Studio)
3. Feed to Cracked Gas (SPYRO)
4. Simulation Model (Aspen Hysys)
5. Checking Products Quality and Quantity
6. Virtual Key Process Parameters Decisions
7. Alternative HEN Design Stage
8. Comparing Performances

Overview of DT algorithm is shared in Figure 5.1;

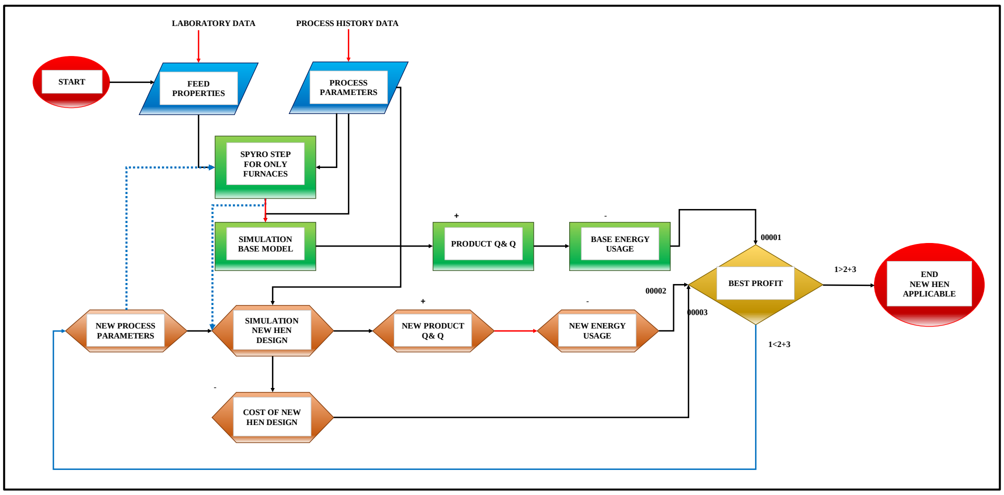


Figure 5.1 Digital Twin Algorithm of Olefin Plant

1. Feed Properties Collection (Sample Manager) step consists data collection for feeds such as naphtha and LPG. Specifications of each feed is very important. While production planning, feedstock properties can be readily determined based on product specifications and market demands. The required feedstock quality can be achieved with reverse engineering. Sample of feed property is shared below Figure 5.2. If naphtha is fed then PINA analysis can be inserted easily. If LPG is feed then components of LPG can be inserted by using composition tab.

Another media is fuel gas for using as a heating media. In addition, fuel gas composition must be entered.

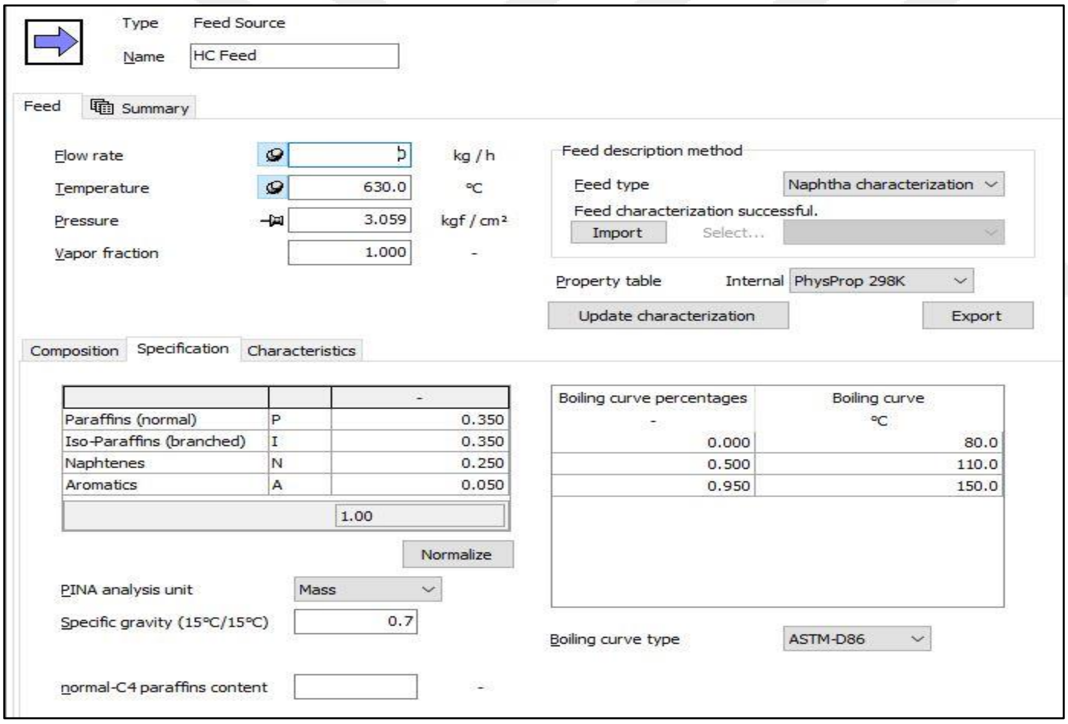


Figure 5.2 One of Sample Feed Property Screen of Digital Twin

2. Process History Data (Process Studio) usage is for collection real plant data. Verification of real data has complexity. Invalid data removed from real steady state value of plant. 1-hour average data is the best such as this kind of study. Sample of feed Process History Data is shared below Figure 5.3.

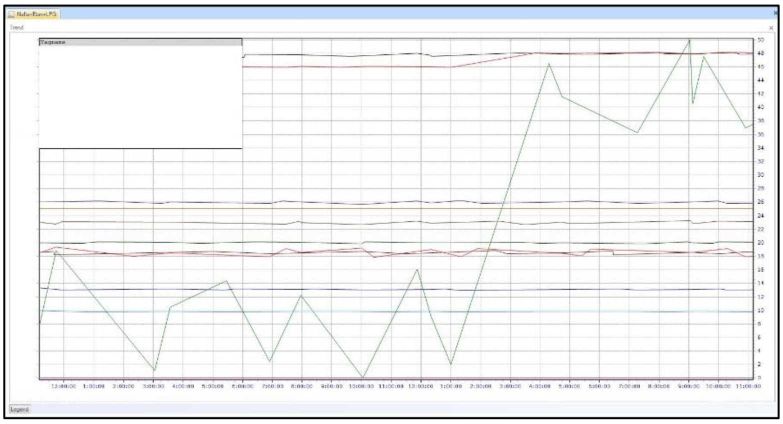
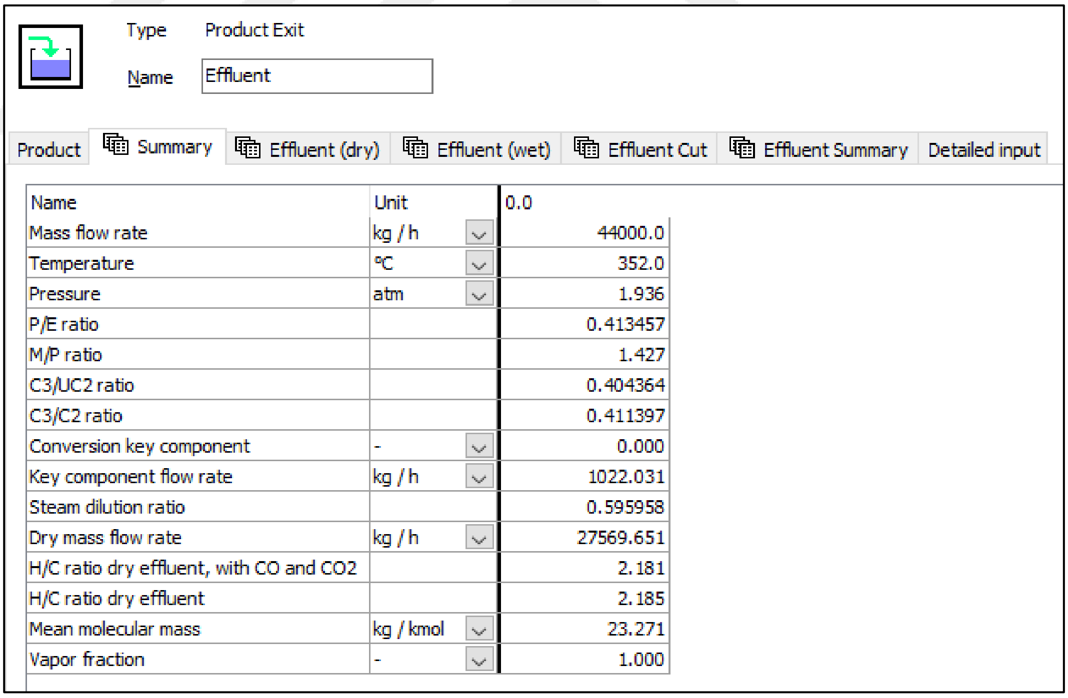


Figure 5.3 Process History Data Screen of Digital Twin

3. Feed to Cracked Gas (SPYRO) step is used for only base of furnaces. DT converts feeds to products using via SPYRO. Calculations done one time manually but SPYRO has excel interface. This Excel interface will facilitate effortless conversion for future applications.



Type Product Exit

Name Effluent

Product | Summary | Effluent (dry) | Effluent (wet) | Effluent Cut | Effluent Summary | Detailed input

Name	Unit	0.0
Mass flow rate	kg / h	44000.0
Temperature	°C	352.0
Pressure	atm	1.936
P/E ratio		0.413457
M/P ratio		1.427
C3/UC2 ratio		0.404364
C3/C2 ratio		0.411397
Conversion key component	-	0.000
Key component flow rate	kg / h	1022.031
Steam dilution ratio		0.595958
Dry mass flow rate	kg / h	27569.651
H/C ratio dry effluent, with CO and CO2		2.181
H/C ratio dry effluent		2.185
Mean molecular mass	kg / kmol	23.271
Vapor fraction	-	1.000

Figure 5.4 Typical Effluent Summary Screen of Spyro

4. Simulation Model (Aspen Hysys) step is classic simulation results for all plant including utilities REP, REE, steam, cooling water, fuel gas, hydrogen and power demand. Screen of plant simulation is shared below Figure 5.4.

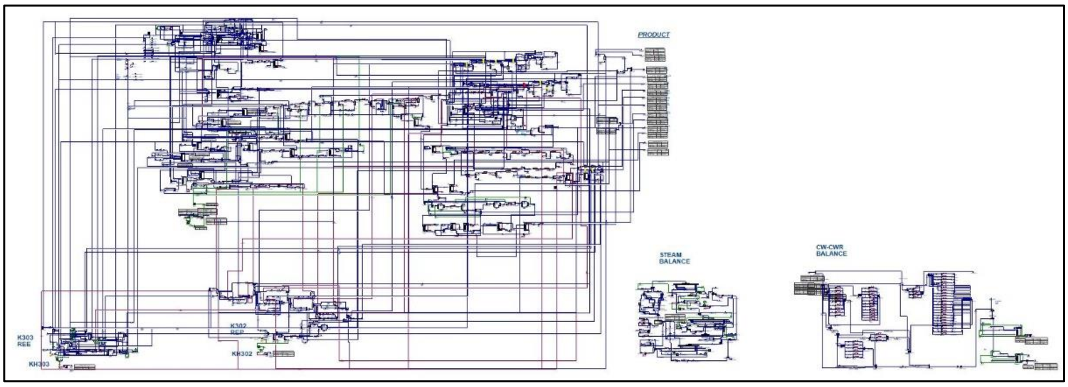


Figure 5.5 Simulation Screen of Digital Twin

5. Checking Products Quality and Quantity is outputs of previous simulation results. All products qualities are set as base case. Impurities are same for all cases. Only differences are quantities of products.

6. Virtual Key Process Parameters Decisions step is that some of equipment is forced over design at maximum 10%. Sample screen of plant column internal check is shared below Figure 5.5.

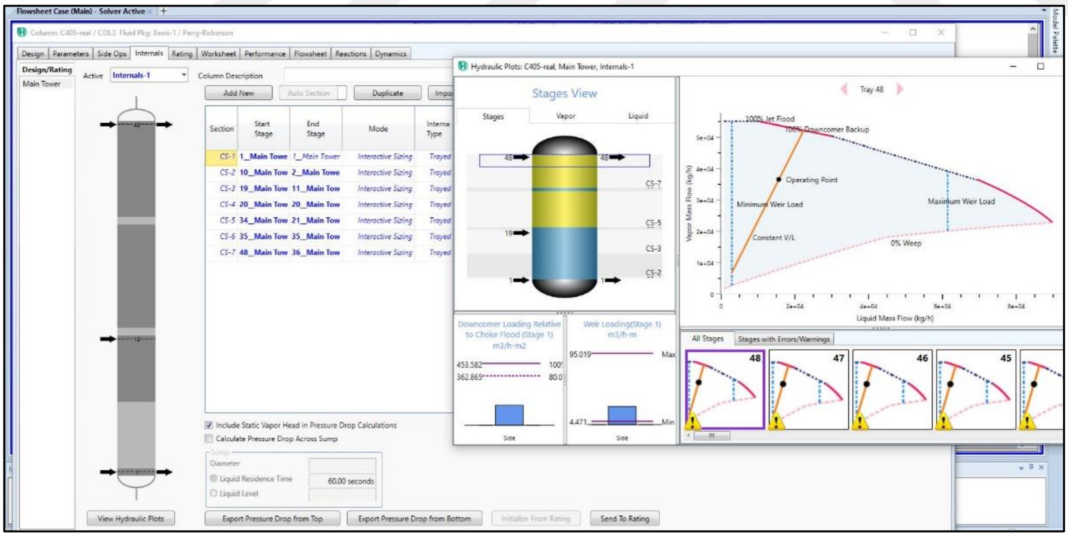


Figure 5.6 Sample Screen of Plant Column Internal Check

7. The core focus of this study is the development of alternative heat exchanger network (HEN) designs. A detailed exploration of these designs can be found in the Results and Discussion section.

8. Comparing Performances is evaluation the differences in terms of gains minus the cost. It is a simple calculation step to define what better performance is chosen.

6. RESULTS AND DISCUSSION

Furnace streams are not taken into consideration because thermal cracking is very complicated chemical reactions, any operation failure causes coking in radiant coils. In literature, the study investigates the failure of finned tubes in the convection section of an ethylene cracking furnace. Metallographic, chemical, and mechanical analyses were employed to identify the root cause. The fracture surface revealed columnar crystal cracks with fatigue striations, indicative of corrosion fatigue. Additionally, dimple features at the crack tip suggested plastic deformation. The presence of chromium carbide precipitates at grain boundaries pointed towards intergranular corrosion. Considering the operating environment with thermal stress and corrosive media, the failure is caused by a combination of corrosion fatigue and intergranular corrosion. (Chen et al., 2023)

In addition, XHS producing in TLX is effecting secondary thermal cracking reactions. On the other hand, cracked gas compressor intermediate hot streams are also not included in the pinch analysis. Consequently, furnace and cracked gas compressor intermediate hot streams should be included in subsequent analyses.

In classical way by using excel, Ethylene process streams are divided two section based on temperature below and above of 0 ˚C. One is called hot section and the other is called cold section. The pinch temperature is calculated for hot section 178˚C; the hot and cold streams pinch temperatures are 183˚C and 173˚C, respectively. In addition, the minimum required utilities are calculated 62478 kW and 151260 kW, respectively (Table 6.1).

The pinch temperature is calculated for cold section -12.2˚C; the hot and cold streams pinch temperatures are -7.2˚C and -17.2˚C, respectively. In addition, the minimum required utilities are calculated 16815 kW and 41301 kW, respectively (Table 6.2).

Table.6.1 Hot Pinch Summary

Hot Side	Q (kW)
Minimum Hot Utulity	62478 (2.2 E+08 Kj/h)
Minimum Cold Utulity	151260 (5.4 E+08 Kj/h)
Hot pinch for hot side	183 ˚C
Thot_side pinch	178 ˚C
Cold pinch for hot side	173 ˚C

Table.6.2 Cold Pinch Summary

Hot Side	Q (kW)
Minimum Hot Utulity	16815 (0.6 E+08 Kj/h)
Minimum Cold Utulity	41301 (1.4 E+08 Kj/h)
Hot pinch for cold side	-7.2 ˚C
Thot_side pinch	-12.2 ˚C
Cold pinch for cold side	-17.2 ˚C

Looking by side of biggest energy user distillation columns have new design features. Dividing- wall and high capacity tray columns has no stagnant area so no loss of efficiency or capacity. If required reflux ratio is decreasing, then OPEX is decreasing too. Tray cost increases but column shell reduces and the net margin is saving overall CAPEX.

The other application is using heat pumps in Ethylene crackers for reducing heat requirement. Heat pump modifications can be obtaining to C2 ethylene- ethane splitter, C3 propylene- propane splitter and depropanizer columns in case of working lower pressures. Generally, if the propylene system is part of an ethylene plant, in which there is typicality a huge heat source of quench water that can be used as heating media for the C3 Splitter reboiler, then heat pump system may be not applicable. In refineries Fluid Catalytic Cracking (FCC) unit or propane dehydrogenation unit, there is no source of sufficient low-grade heat is available, and then applying heat pump is typically the better choice (Linhoff, 1998).

On the other hand as industrial process heat demands rise, heat pumps become less efficient, and investment decisions more complex. The study proposes a methodology combining mixed-integer linear programming and pinch analysis to optimize heat pump integration into existing systems. A case study demonstrates that even under unfavorable economic conditions (high electricity vs gas price ratio), heat pump integration can yield marginal economic benefits of up to 6%. The developed approach supports companies in making informed decisions about sustainable and cost-effective heat generation strategies. (Knorr et al., 2024)

Olefin plant turbines are using XHS steam is a main energy carrier. While this is a great opportunity, accurate performance monitoring is very challenging (Yu et al., 2020). The primary source is a byproduct of cracked gas cooling furnaces. Advantages of steam are easy to generate, easy to transport, suitable for heat release, transfer, large energy release benefit from latent heat, low toxicity and suitable for transformation into mechanical energy as said. It can be applicable for Olefin plant to increase deaerator pressure like energy plants. Typical Olefin plants deaerators working pressure 0.5 kg/cm^2 to 3 kg/cm^2, it can be increased 5 kg/cm^2 to 8 kg/cm^2. Here is limitation is coming boiling water preheater outlet temperature of furnace. Small amount of heat sink can be afforded. Hence, total XHS produced will increase up to 3%.

Last integration is applied by using GTG. That modification to produce power while utilizing typically 496 $^{\circ}C$ hot and 15% O_2 rich exhaust gas as combustion air for cracking furnaces. Turbine exhaust gas is providing lowering furnaces fuel requirement. In addition, steam production in convection of furnaces section is increasing too (Tahery, 2014).

For Plant GTG modification is very complicated because in generally there is insufficient space for a GTG in the old plant layout. Also, if we install such equipment applying turbine exhaust will be more challenging to furnace radiant section for old type furnaces. Mostly new type of furnaces can be chosen for this modification. Burner type is the key here; also, best solution may be next to new type furnaces.

On the other hand, power integration of GTG will be a problem via electricity connection for an old plant. This problem is not included scope of work.

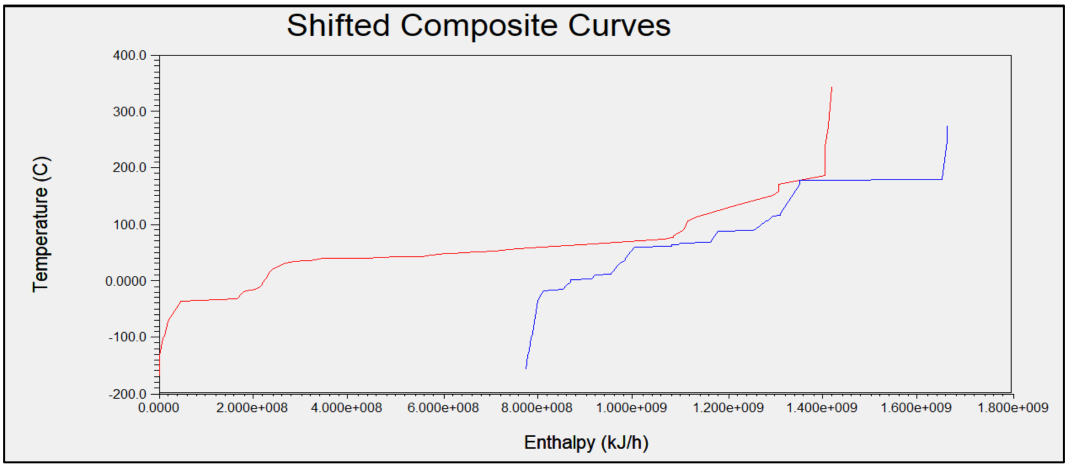


Figure 6.1 Shifted Composite Curves

In Figure 6.1 The Shifted Composite Curve is a tool used in heat exchanger network (HEN) design to identify the pinch point, which is the location in the network where the maximum heat recovery can be achieved. The Shifted Composite Curve (Linhoff, 1998) is constructed by shifting the Hot Composite Curve (HCC) down by $\Delta Tmin/2$ and the Cold Composite Curve (CCC) up by $\Delta Tmin/2$, where $\Delta Tmin$ is the minimum approach temperature.

The pinch point occurs where the shifted hot and cold curves intersect. At this point, no further heat transfer is possible without violating the minimum temperature difference requirement.

The Shifted Composite Curve is a valuable tool for determining the pinch point because it is easier to visualize than the original composite curves. The shift also simplifies the identification of the matches that are most beneficial for heat recovery.

There are two main pinch points. Because of the nature of Olefin process there are too hot and extremely cold streams. One of the pinch belongs to cold pinch point, second belongs to hot pinch point.

The figure (Figure 6.1) unshifted composite curve visually represents the temperature and heat load profiles of all hot and cold process streams. The shifted composite curve is a similar graph, but the hot and cold streams have been shifted vertically by half the minimum approach temperature.

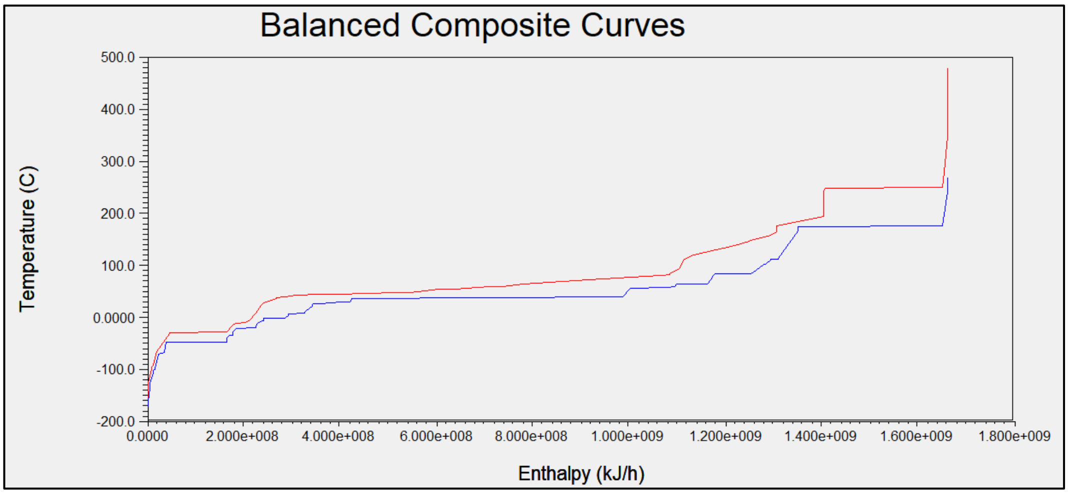


Figure 6.2 Balanced Composite Curves

In Figure 6.2 The Balanced Composite Curve (BCC) graphically represents the combined temperature and heat load profiles of both process and utility streams within a system. The BCC is constructed by combining the Hot Composite Curve (HCC) and the Cold Composite Curve (CCC).

The HCC is a composite of both the hot process and utility streams. The HCC encompasses the temperature and heat load data for all hot process and utility streams.

The CCC is a composite of both the cold process and utility streams. The CCC encompasses the temperature and heat load data for all cold process and utility streams.

The BCC is constructed by superimposing the HCC and the CCC on the same graph. The vertical gap between the two curves indicates the heat duty required from utility streams.

The BCC can be utilized to estimate the total heat transfer area required for the heat exchanger network. This is achieved by dividing the total process heat load by the LMTD derived from the BCC. The BCC provides a comprehensive overview of process heat loads, enabling efficient identification of potential heat matches and accurate estimation of required heat transfer area.

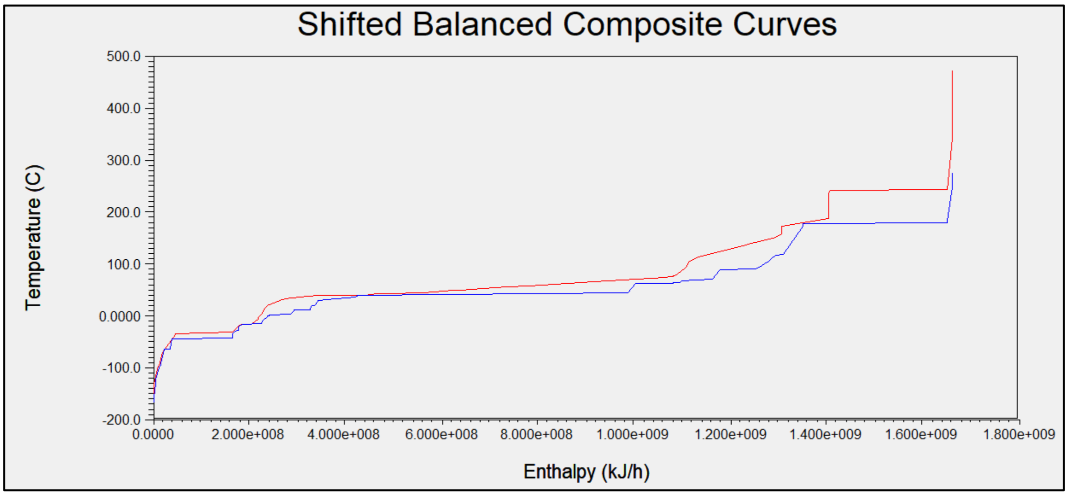


Figure 6.3 Shifted Balanced Composite Curves

In Figure 6.3 The Shifted Balanced Composite Curve (SBCC) visually represents the temperature and heat load profiles of all process and utility streams after applying the cascade algorithm, including utility streams, after the hot and cold streams have been shifted vertically by half the minimum approach temperature. The SBCC is constructed by combining the Shifted Hot Composite Curve (SHCC) and the Shifted Cold Composite Curve (SCC).

The SHCC is a composite of both the hot process and utility streams after they have been shifted vertically by half the minimum approach temperature. The SHCC incorporates the temperature and heat load data for all hot process and utility streams after applying a temperature shift equal to half the minimum approach temperature.

The SCC is a composite of both the cold process and utility streams after they have been shifted vertically by half the minimum approach temperature. This means that the SCC includes the temperature and heat duty of all the cold streams in the process, as well as the temperature and heat duty of the cold utility streams after they have been shifted by half the minimum approach temperature.

The SBCC is constructed by superimposing the SHCC and the SCC on the same graph. The vertical gap between the two curves represents the heat duty of the utility streams after they have been shifted by half the minimum approach temperature.

The SBCC enables the identification of pinch points, where hot and cold streams exhibit their maximum temperature difference. Additionally, the SBCC facilitates the calculation of necessary heat transfer area within the heat exchanger network. Also, all pinch points are listed in Table 6.1 and Table 6.2.

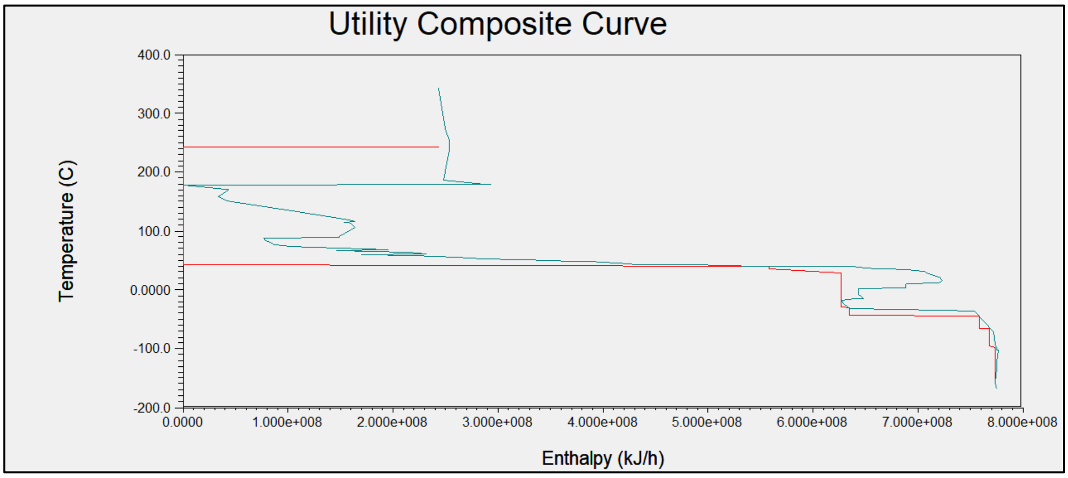


Figure 6.4 Utility Composite Curves

In Figure 6.4 A Utility Composite Curve (UCC) is a graphical depiction of temperature and heat load of all the hot and cold utility streams in a process. The UCC is constructed likewise as a Grand Composite Curve (GCC), but it includes the hot and cold utility streams in addition to the process streams.

The UCC can be utilized to establish the minimum hot and cold utility requirements for the process. This is achieved by locating the point on the UCC where the two curves intersect. The point of intersection represents the minimum temperature gap between the hot and cold fluids, and the heat duty at this point represents the minimum hot and cold utility requirements.

The UCC can also be used to visually inspect how much of each utility contributes to the total utility target. This is done by looking at the slope of the UCC. The steeper the slope, the more of that utility is required. For example, if the UCC has a steep slope in the hot utility region, then this indicates that a lot of hot utility is required.

The UCC provides a comprehensive overview of utility heat loads within the process. This information aids in identifying optimal heat integration opportunities and determining overall utility requirements.

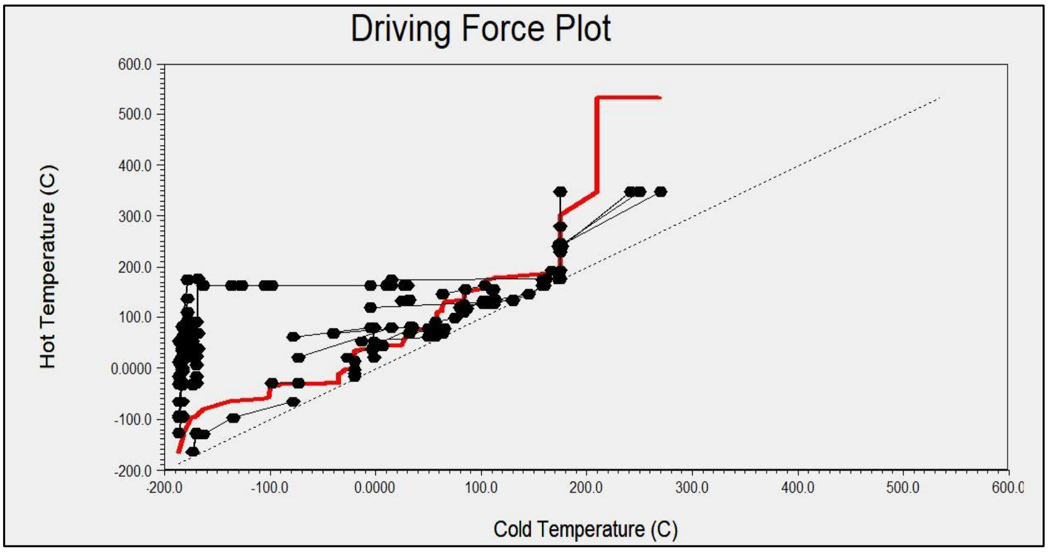


Figure 6.5 Driving Force Plot

In Figure 6.5 The Driving Force Plot visually represents the temperature differential between hot and cold streams within a heat exchanger network. This temperature difference, or driving force, is a critical factor influencing heat exchanger efficiency.

The Driving Force Plot view is typically used in conjunction with the Grid Diagram view. The Grid Diagram view shows the physical layout of the heat exchanger network, and the Driving Force Plot view shows the driving force within the hot and cold streams at each location in the network.

The Driving Force Plot view is a valuable tool for heat exchanger network design because it allows you to visualize the driving force within the hot and cold streams. This information aims to identify potential matches within the hot and cold streams, as well as to optimize the layout of the heat exchanger network.

The Driving Force Plot view will stay open and "float" above the Grid Diagram, so that you can view the information as you are building the network of heat exchangers. This allows you to see how the driving force changes as you make changes to the network, and it helps you to ensure that the network is optimized.

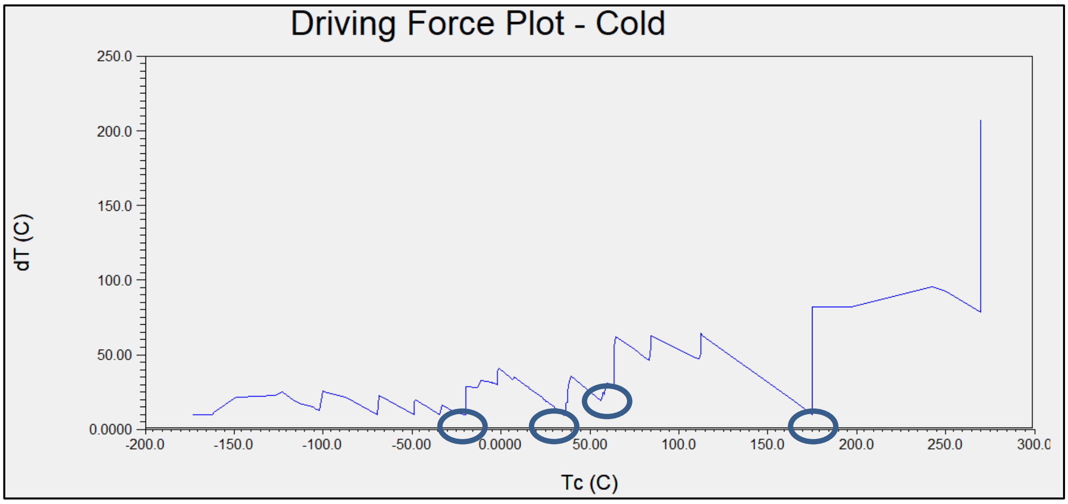


Figure 6.6 Driving Force Plot Cold

In Figure 6.6 The Cold Driving Force Curve graphically illustrates the temperature difference between corresponding points on the hot and cold composite curves. The hot composite curve (HCC) is a plot of the temperature of the hot streams in a process, and the cold composite curve (CCC) is a graphical representation of cold stream temperatures within a process. The cold driving force curve plots the difference between the two curves against the cold stream temperature.

The cold driving force curve can be used to identify the pinch point in a heat exchanger network. The pinch point is the point in the network where the maximum heat recovery can be achieved. The cold driving force curve will show a sharp increase in the minimum temperature difference.

The cold driving force curve can also be used to determine the minimum cold utility requirements for a process. The minimum cold utility requirements are the amount of cold utility that is required to cool the cold streams to their target temperature. The cold driving force curve will show a plateau at the target temperature, and the height of the plateau will represent the minimum cold utility requirements.

The cold driving force curve is a useful tool for heat exchanger network design because it provides a visual representation the temperature differential within the hot and cold fluids. This knowledge facilitates the discovery of the pinch point and the minimum cold utility requirements. The pinch of the heat exchanger network can be observed as the lowest point of the curve by blue circles, closest to the Temperature cold axis.

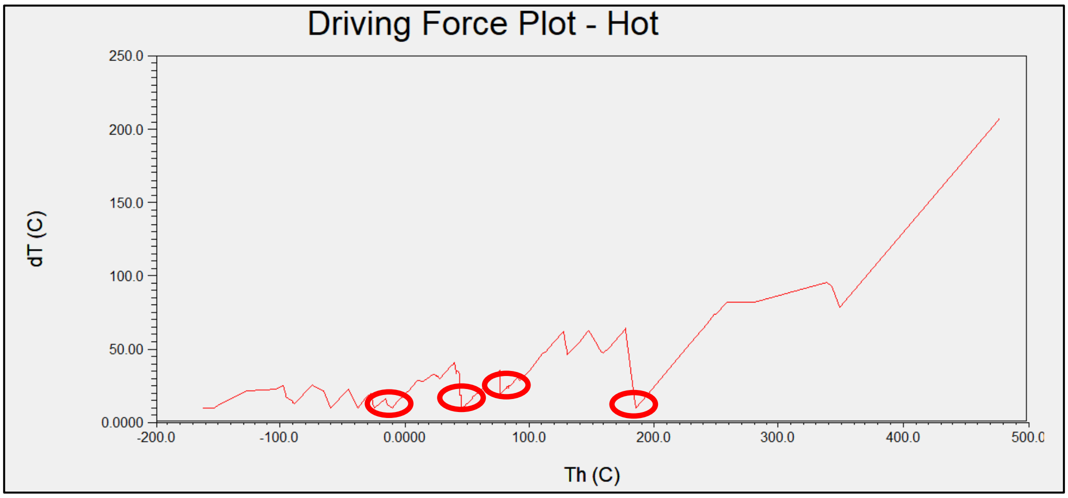


Figure 6.7 Driving Force Plot Hot

In Figure 6.7 The Hot Driving Force Curve is a graphical representation of the temperature difference between the hot and cold composite curves. The hot composite curve (HCC) is a plot of the temperature of the hot streams in a process, and the cold composite curve (CCC) is a plot of the temperature of the cold streams in a process. The hot driving force curve plots the difference between the two curves against the hot stream temperature.

The hot driving force curve can be used to identify the pinch point in a heat exchanger network. The pinch point is the point in the network where the maximum heat recovery can be achieved. The hot driving force curve will show a sharp increase in the temperature difference at the pinch point.

The hot driving force curve can also be used to determine the minimum hot utility requirements for a process. The minimum hot utility requirements are the amount of hot utility that is required to heat the hot streams to their target temperature. The hot driving force curve will show a plateau at the target temperature, and the height of the plateau will represent the minimum hot utility requirements. The hot driving force curve is a useful tool for heat exchanger network design because it provides a visual representation of the temperature difference between the hot and cold streams. This information can be used to identify the pinch point and the minimum hot utility requirements.

The hot driving force curve is similar to the cold driving force curve, but it plots the temperature difference against the hot stream temperature instead of the cold stream temperature. This difference in perspective can be helpful for identifying pinch points in heat exchanger networks with multiple hot streams. The pinch of the heat exchanger network can be observed as the lowest point of the curve by red circles, closest to the Temperature hot axis.

HENs are a complex and challenging problem to solve. However, there are techniques that can be used to systematically compare different alternatives and to identify the optimum network configuration. Aspen Energy Analyzer can be a valuable tool for non-experts who want to design HENs. In Figure 6.8 Aspen Energy Analyzer that generated best retrofitted new HEN.

In addition, Pinch Points of new HEN are included in Figure 6.9.

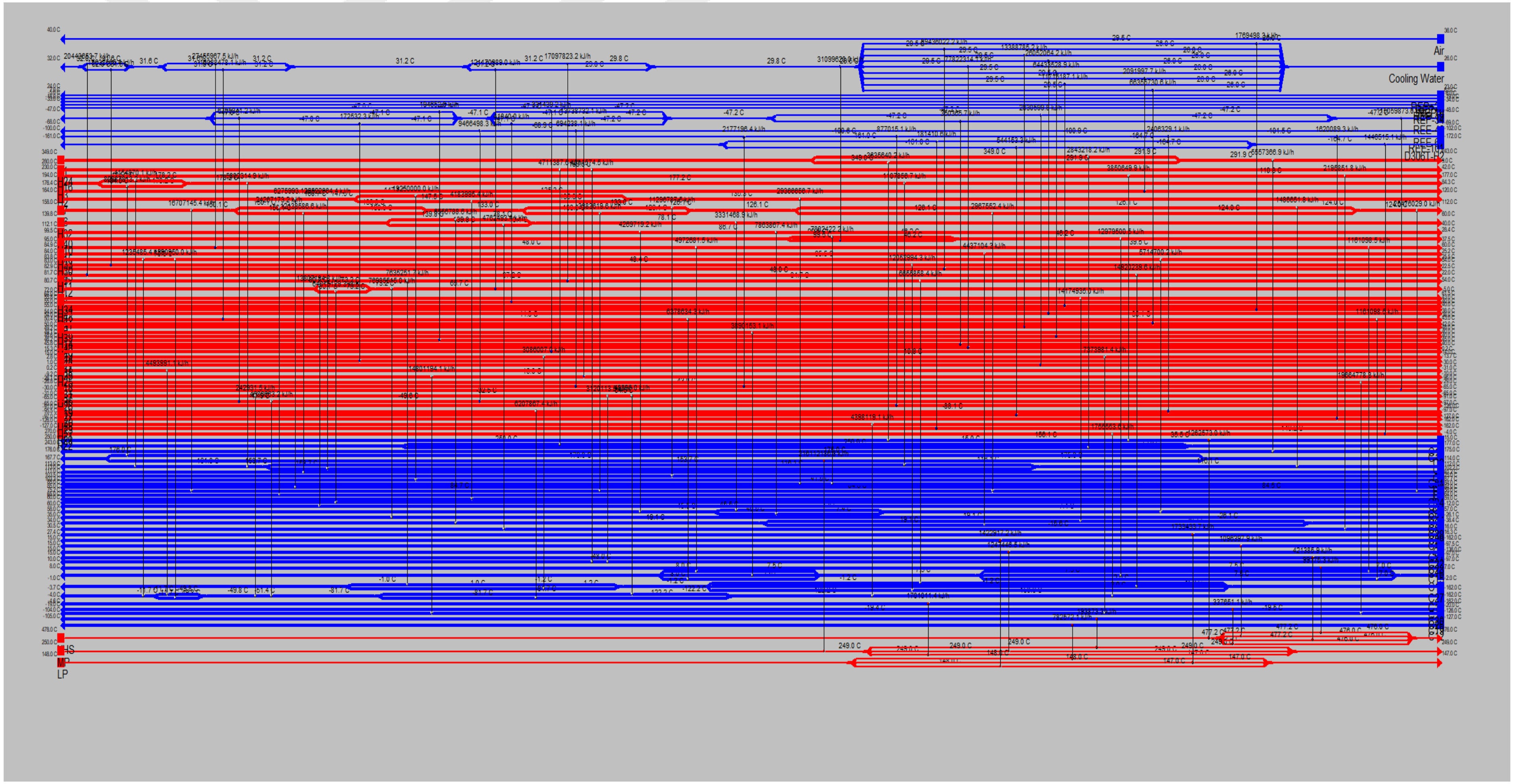


Figure 6.8 Cascade Algorithm of New Best Design

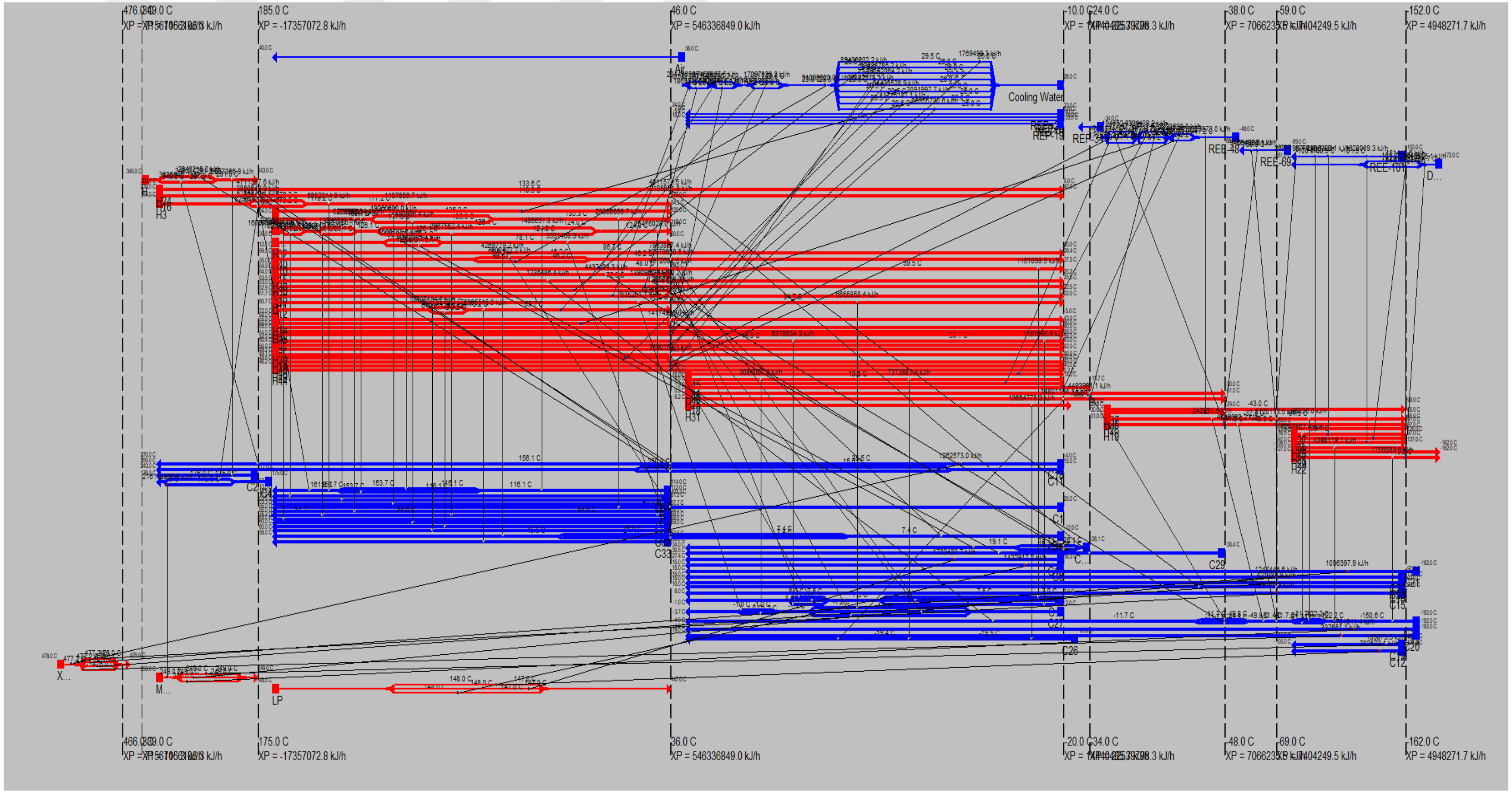


Figure 6.9 Cascade Algorithm of First New Best Design with Pinch Points

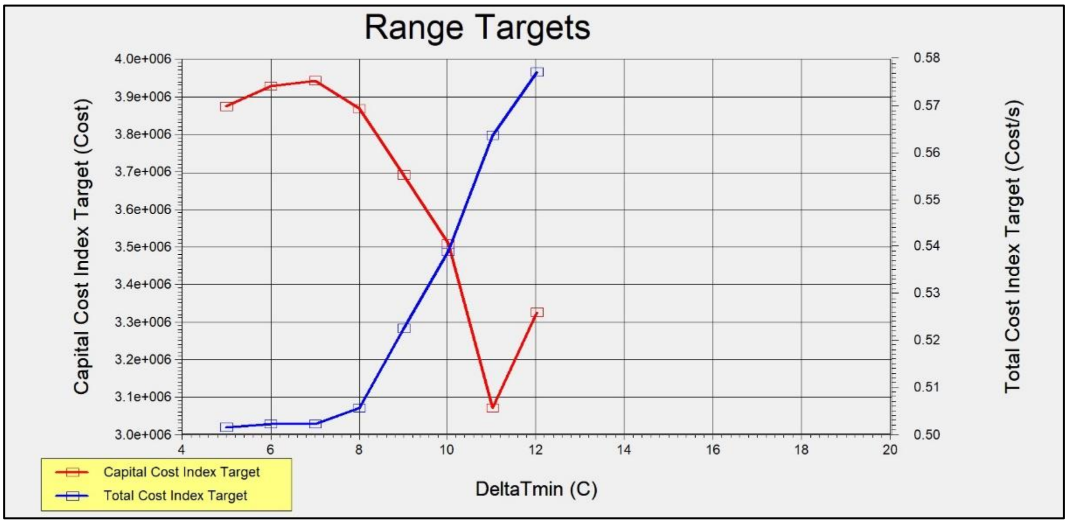


Figure 6.10 Capital Cost Index and Total Cost Index vs ΔT_{min}

Industrial acceptance for petrochemical and refinery application ΔT min is selected 10 °C for in this study. Exchanger types are simple shell and tube exchangers. Nowadays new core in kettle type exchanger let us obtain ΔT min less than 10 °C. Advantages will be increasing heat flux, reduced equipment size and removable cores benefit to clean easy. Therefore, new results have a little bit higher CAPEX because core in kettle is new technology but much lower OPEX. Figure 6.10 Capital Cost Index and Total Cost Index vs ΔT_{min} shows how temperature difference is effecting CAPEX and OPEX.

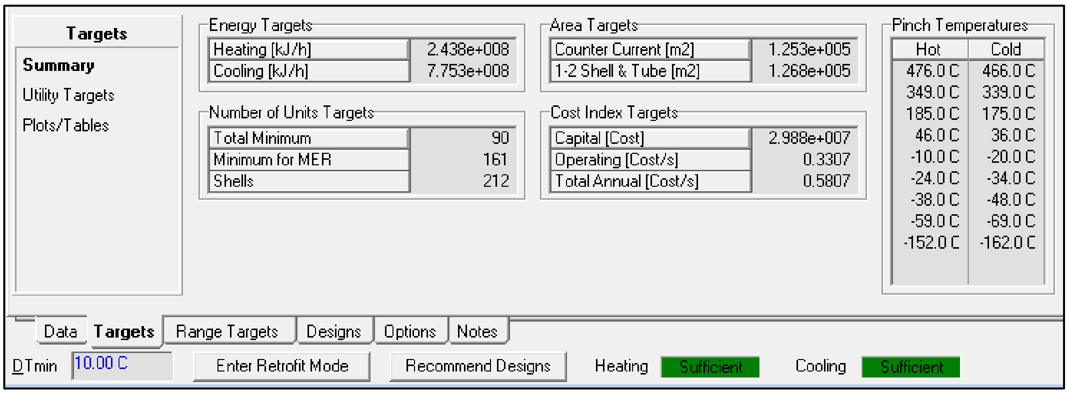
When no range values for ΔTmin are chosen, Aspen Energy Analyzer will use the cost data for all ΔTmin values ranging from zero to the calculated maximum ΔTmin. This will allow for an economic evaluation of all possible heat exchanger networks.

The default interval for which the ΔTmin range is divided is determined using the Golden Search method. The Golden Section Search is a straightforward yet effective technique for locating the optimum of a unimodal function. It works by repeatedly dividing the search interval in half and then choosing the half that has a higher probability of encompassing the optimal value. http://mathforcollege.com/nm/mws/gen/09opt/mws_gen_opt_txt_goldensearch.pdf (Access date: January 2019).

In Figure 6.10 Capital Cost by red line and Total Cost by blue line vs various ΔT_{min} are plotted. Once this information is provided, Aspen Energy Analyzer can calculate the Total Annual Cost of the heat exchanger network. This information can then be used to compare different heat exchanger networks and to select the one that is most cost-effective.

- Values for each of the utility costs: This includes the cost of each utility, such as steam, electricity, and cooling water. Utility costs fluctuate based on plant location and prevailing market conditions.
- Cost law for the heat exchanger network: This equation quantifies the heat exchanger network's cost. Several cost models exist that can be used, and the choice of cost law will impact the overall cost of the heat exchanger network.
- Values for the annualization factor parameters: These parameters include the Rate of Return (ROR) of the plant and the plant life (PL). The ROR is the percentage of the investment that is returned each year, and the PL is the number of years that the plant is expected to operate. The values of these parameters will affect the annualization factor, which determines the total annual cost of the heat exchanger network.

Table 6.3 Targets Summary & Pinch Points for Aspen Energy Analyzer



There are four new HEN design offered for Olefin Plant using Aspen Energy Analyzer. Targets are summarized and Pinch Points are shown in Table 6.3 information is split into five groups:

The Energy Targets group displays the minimum amount of additional heating and cooling energy required for the process streams to achieve their outlet temperature. The additional energy value is based on the total energy required (by the process streams) minus the energy available for transfer (in the process streams).

- Total Minimum: This is the minimum total amount of energy required for the heat exchanger network.
- Minimum for MER: This is the minimum amount of energy required for the heat exchanger network if the design takes into account the pinch temperature. The pinch temperature is the minimum temperature difference between the hot and cold streams where the maximum heat recovery can be achieved.
- Shells: This is the minimum number of shells required for the heat exchanger network. A shell is a unit of a heat exchanger that contains multiple tubes. The Area Targets group displays the minimum heat transfer area required if all the exchangers were Counter Current type and if all the exchangers were 1-2 Shell & Tube type. The Cost Index Targets group displays the minimum capital, operating and total annual cost values.

The Area Targets group displays the minimum heat transfer area required for the heat exchanger network if all the exchangers were Counter Current type and if all the exchangers were 1-2 Shell & Tube type.

The Cost Index Targets group displays the minimum capital, operating and total annual cost values.

The Pinch Temperatures group displays the possible Hot and Cold temperature values.

Below Table 6.4 consists four recommended new designs by Aspen. Best design is shown in yellow. Design 1 has ΔT_{min} 10 °C, the other Designs 4, 5 and 6 has same ΔT_{min} 2 °C each other. Table shows us there is no big difference in capital costs for ΔT_{min} between 10 °C and 2 °C, if approach is same for both cases. In A_Design5, increasing both units and shells increases capital cost as predicted.

Alternatively, decreasing units and increasing shells in A_Design6 effects capital cost negatively, too.

Table 6.4 Properties of Recommended New Designs by Aspen Energy Analyzer

HI Project: Case 1 (Designs)

Design		Total Cost Index [Cost/s]	Area [m2]	Units	Shells	Cap. Cost Index [Cost]	Heating [kJ/h]	Cooling [kJ/h]	Op. Cost Index [Cost/s]
A_Design5		0.4625	3.931e+00	110	166	1.128e+007	2.264e+008	7.579e+008	0.3682
A_Design6		0.4509	3.404e+00	103	145	9.885e+006	2.264e+008	7.579e+008	0.3682
A_Design4		0.4460	3.181e+00	106	138	9.295e+006	2.264e+008	7.579e+008	0.3682
A_Design1		0.4460	3.181e+00	106	138	9.295e+006	2.264e+008	7.579e+008	0.3682

7. CONCLUSION

DT is a digital monitoring of the actual olefin process that allows for better operation and understanding of the history of actual data. Even traditional simulation is accurate representation of a particular operating case; DT is representation of the plant over its full range of operation. Thus, the full history and future of the asset is captured by using DT. Therefore, DT outputs that allows monitoring reality versus model results in an easy way. DT is sensitive to changes in feeds such as naphtha, LPG and recycled ethane. More LPG processing as a feed is highly recommended for olefin plant depends on economics. Ethylene and propylene are more valuable products rather than heavier products. In addition, DT is centralized data collection and reporting for typical olefin plant utilities such as REP, REE, steam, cooling water, fuel gas, hydrogen, power demand and emissions beside process data.

One of DT output is for utilities REP and REE cooling temperature levels. The cold end of an olefin plant is where the extremely cold temperatures are used to separate the olefins from the other components of the crude oil. This section of the plant typically uses refrigeration processes to achieve the required cold temperatures. The minimum approach temperature (ΔT_{min}) is the smallest allowable temperature difference between hot and cold process streams within a heat exchanger. The smaller the ΔT_{min}, the more efficient the heat exchanger will be. However, the smaller the ΔT_{min}, resulting in a higher heat exchanger cost. The cold end of an olefin plant, the ΔT_{min} can be very small because the cold temperatures are already very low. This means that the heat exchangers in plant cold end can be very efficient, resulting in substantial energy cost reductions. The layout of the cold end units is also important for energy recovery. The cold end units should be located close together to facilitate heat transfer. The refrigeration systems for cold end are key for energy savings. The refrigeration systems should be efficient and well-maintained. One of the best technology criteria in the heat exchanger networks in cold end is to choose a minimum temperature approach of within 2 °C and 5 °C. This will allow for efficient heat transfer while still prediction of the cost reasonable. In a quick run, ΔT_{min} of 2 °C was done but results of DT was not better than ΔT_{min} of 10

°C. Hence, exergy changes in each temperature level of the REP and REE cycles should be computed to define which cycles are precisely same as each other for further work in a detail way.

DT's predictive power answers what happens if feedstock or conditions change. DT is not just a model. DT includes also accurate data generation, gap analytics and energy benchmarking. What-if scenarios represents mass, energy and utility balance for an achievable optimized scenario. DT solutions give an idea for aligning with high priorities. For example, GTG can be studied for old furnaces too, even if the Return of Investment is low, the benefit may be valuable and to force get this GTG investment to all old furnaces. Nearly 2 t/h fuel gas saving is observed in DT by integration of 9.4 MW GTG to new furnaces. Thus, carbon emissions could be decreased with this simple solution. Adding higher capacity GTG or two twin GTG could be more efficient and safely. If any shut down happens in steam and power utility generation supplier, olefin plant can easily keep production without any fluctuations in its own process. Therefore, sustainability of olefin plant increases by doing that GTG modification.

By the way, all economic investigations are done by making cost analysis. In this study, Aspen Process Economic Analyzer is used for two new HEN design for Olefin Plant. These results are near the close 40 % in real world. There is another better estimation option for cost of utilities called as operating and equipment cost called as capital cost. Therefore, future research should be for the heat exchanger network with using Aspen Capital Cost Estimator. Those new results will be nearly same configuration with chosen better HEN design, but accuracy will increase up to 10 %.

Economic evaluation alone is insufficient for real-world applications. There are too much limitations for old sites had many revamped in past. New HEN designs need to inspect better locations for new heat exchangers. Then civil work, additional valves with instrumentation's, control philosophy, piping routes and pipe racks will be required. All these items must be considered carefully (Akbarnia et al., 2009).

On the contrary, if new U_{min} job opportunities arise in section result and discussion are implemented, OPEX (Operational Expenses) will be minimized by using together new digital technologies such as Advanced Process Control System (APC) and Real Time Optimization (RTO).

Finally, the study can be conducted again by using multistage optimization algorithms or combined with exergy analysis for more effective analysis (Abbood et al., 2012).

Combined exergy and pinch analysis must be target in next step. Especially including steam driver turbines in Olefin Plant must be considered carefully (Bandpy et al., 2011).

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Finally, dear reader thanks for your time and I hope you to enjoy reading this thesis. I kindly request you to contact with me for further questions, I am really willing to learn better ideas for topic of this study with your new questions and your intelligent suggestions.

Mr. Halil ÇELİKKAYA

Education

2022 - 2023	Sabancı University Executive MBA
2014- Cont.	Ege University Department of Chemical Engineering- Msc
2003- 2008	Ege University Department of Chemical Engineering- Bsc

Job Experience

(2023- Now)	Socar Turkey Technical Services- Ethylene, EO and ACN Process Manager
(2022- 2023)	Socar Turkey Technical Services- Ethylene Process Manager
(2019- 2022)	Socar Turkey Technical Services- Process Engineering Department- Lead Engineer

Refinery and Petrochemical Complex (Monomers and Polymers) Projects Management

- ROG integration project
- Cogeneration project
- Refinery Boiler optimization
- Ethylene furnace EOR optimization (6 sigma)
- Ethylene GTG integration project
- Ethylene recycle gas optimization

(2018- 2019) Petkim Petrochemical- Production Management Office-Supervisor (PMO SI)

Monomer Plants Optimization, Value Creation and Digitalization Projects Management

- Ethylene recovery in cold box and turbo- expander system
- Exchanger placement to utilize dsg blow down heat
- Ethylene air coolers blades change
- Ethylene C2 reactor optimization (6 sigma)

- Ethylene C3 reactor optimization (6 sigma)
- Cooling water temperature reduction
- Cooling tower efficiency improvement
- Ethylene plant furnace coil renewal
- Ethylene steam system optimization
- Install improved O2/CO analyzers and Combustion One advanced controllers to ensure stack excess O2 controlled continuously
- Prosteam energy trading/ Petkim energy optimization
- Aromatic plant sulfolane non-aromatic cycle optimization
- Ethylene plant's column optimization

(2013- 2019) Petkim Petrochemical Ethylene Plant (Stone & Webster- Technip Licensor)- Process Senior Engineer

- Responsible for APC Honeywell revamp project
- Responsible for efficiency and reliability monitoring of turbo machinery (expander & compressors- cracked gas, refrigerant ethylene & propylene)
- Responsible for efficiency of heat exchangers
- RBI for pressure relief devices of NSC plant
- Support NSC plant Hazop study project (SEVESO- III directive)
- IIIrd revamped project of NSC plant (%13 capacity increase)

-Provide technical support for executing revamp project

- Support pre-commissioning after Revamp III

-Develop Aspen Hysys process simulation models for revamp process design of columns, thermal design of heat exchangers in terms of energy analysis

- Improve and optimize manufacturing processes to maximize productivity by using SPYRO in Liquid Naphtha Cracker, LPG Cracker and Ethane Cracker

- FAT of III. Revamp DCS YOKOGAWA Centum CS 3000 to Vp and ESD YOKOGAWA Prosafe **(Yokogawa/ ITALY)**

(2010- 2013) Petkim Petrochemical Ethylene Plant- Process Engineer

- Ethylene pinch analyze
- Identify bottlenecks and troubleshooting by analyzing operating data

(2009 – 2010) BatıSöke Cement Factory- Shift Engineer (11 month)

(2007) Trainee in PETKİM LDPE-T (Low Density Polyethylene-Tubular) Plant

(2006) Trainee in ER-BAKIR Electrolytic Copper Wire Production

<u>Language</u>

English (fluent), German & Japanese (3-month course- beginner)

Knowledge of Computer Programs

Aspen Certified User in Aspen Hysys (**Certification ID: 00004097**)
https://www.credential.net/876b43b8-6ef0-44c4-93e1-3b23a9a5190d
Aspen Plus, Aspen Energy Analyzer, Aspen Flare System Analyzer, Unisim, ChemCad, Pro II, Unisim, Prosteam, Pipenet, DCS Yokogawa (CS 3000, Centum Vp), ESD Yokogawa Prosafe, Microsoft Visio, Matlab, Comsol Multiphysics, MS Project, Microsoft Office, Visual Basic, AutoCAD, Navisworks, IntelliCAD

Participate In Summits& Seminars & Projects & Voluntary Works & Interests

- Shell Geotem 2023 Conference- **Lisboa/ Sep 2023**
- Speaker "Petkim Petrochemical Plant Ethylene Plant Energy Review with KBC"- Asian Downstream Summit- **World Refining Association- Singapore/ Oct 2019**
- Speaker "Case Study: Star Refinery- Petkim Petrochemical Plant Integration"- Mediterranean Refining& Petrochemicals Summit- **S&P Global Platts- SPAIN/ June 2019**
- Perform the Console FAT of Ethylene Plant DCS- ESD Upgrade Project- **Gesab- SPAIN/ July 2018**
- Petkim Ethylene Plant Energy Study Project- **KBC- UK/ March 2018**
- Introduction to Pinch Analysis and Super Target- **KBC- UK/ March 2018**
- Petro-SIM Basics and Utility Modelling- **KBC- UK/ March 2018**
- EEPC 2017 (European Ethylene Producers Committee)- **EEPC- GERMANY/ Oct 2017**
- Mitsubishi Heavy Industries Turbo Compressor Operation Training- **MHI- JAPAN/ Apr 2017**
- Franklin Covey The 7 Habits of Highly Effective People® Signature Edition 4.0- Izmir/ March 2017
- Turbo Expander Compressor Full Training Course (MTC 200/90)- **Cryostar- FRANCE/ Oct 2016**
- Ethylene Middle East Technology Conference- **EMET- BAHRAIN/ Oct 2016**
- CCPS' Overview of Risk Based Process Safety- **AICHE Academy- BAHRAIN/ Oct 2016**
- Aspen Tech APC User Conference- **Aspen- NETHERLANDS/ May 2016**
- Linde Olefin Academy- **Linde- GERMANY/ April 2016**
- Seveso III Training- Turkey Chamber of Chemical Engineers- Istanbul/ Dec 2015
- Linde Ethylene Licensor Conference- **Linde- GERMANY/ June 2015**
- Yokogawa User Conference- **Yokogawa- GERMANY/ July 2014**
- ADR 1.3.2.1 TUGEM- Izmir/ Apr 2014
- FAT of Naphtha Steam Cracker IIIrd Revamp Project- **Yokogawa- ITALY/ 2013**
- "Training of Trainers" certificate from The Istanbul University (for providing engineering training courses)/ 2012
- Cracking Furnace Conference/ İstanbul TURKEY/ 2012
- Honeywell Unisim Design and Process Modeling Training Course/ Izmir/ Feb 2011
- Rotary Kiln Furnace Simulation Training /ANKARA TÇMB/ March 2010

- I. st Viticulture and Winemaking Symposium- Member of the Executive Committee/ DENİZLİ /Sept.- October 2008
- Total Quality Management (Ege Uni. Chemical Eng.)/ April 2008
- Communication In Family (ÜEMTEM)/ Feb. 2008
- Business and Human Relations (ÜEMTEM)/ Jan. 2008
- Communication Skills- How to Read Body Language (ÜEMTEM)/ Sep. 2007
- KAIZEN- (ÜEMTEM)/ Aug. 2007
- Cv Writing and Interview Techniques (ÜEMTEM)/ July 2007
- NLP- (ÜEMTEM)/ April 2007
- Time Management (ÜEMTEM)/ April 2007
- Project of Child Training (İzmir Metropolitan Municipality)/ June 2007- 2008
- Quality Management Systems- ISO 9000: 2000 (İzmir Academy)/ Dec. 2006
- Starfish Project- Training of Future Project Leaders- İzmir/ 2005- 2006

References
- Available upon request

APPENDICES

Appendix 1. GTG NovalT12 Modification in Aspen Hysys

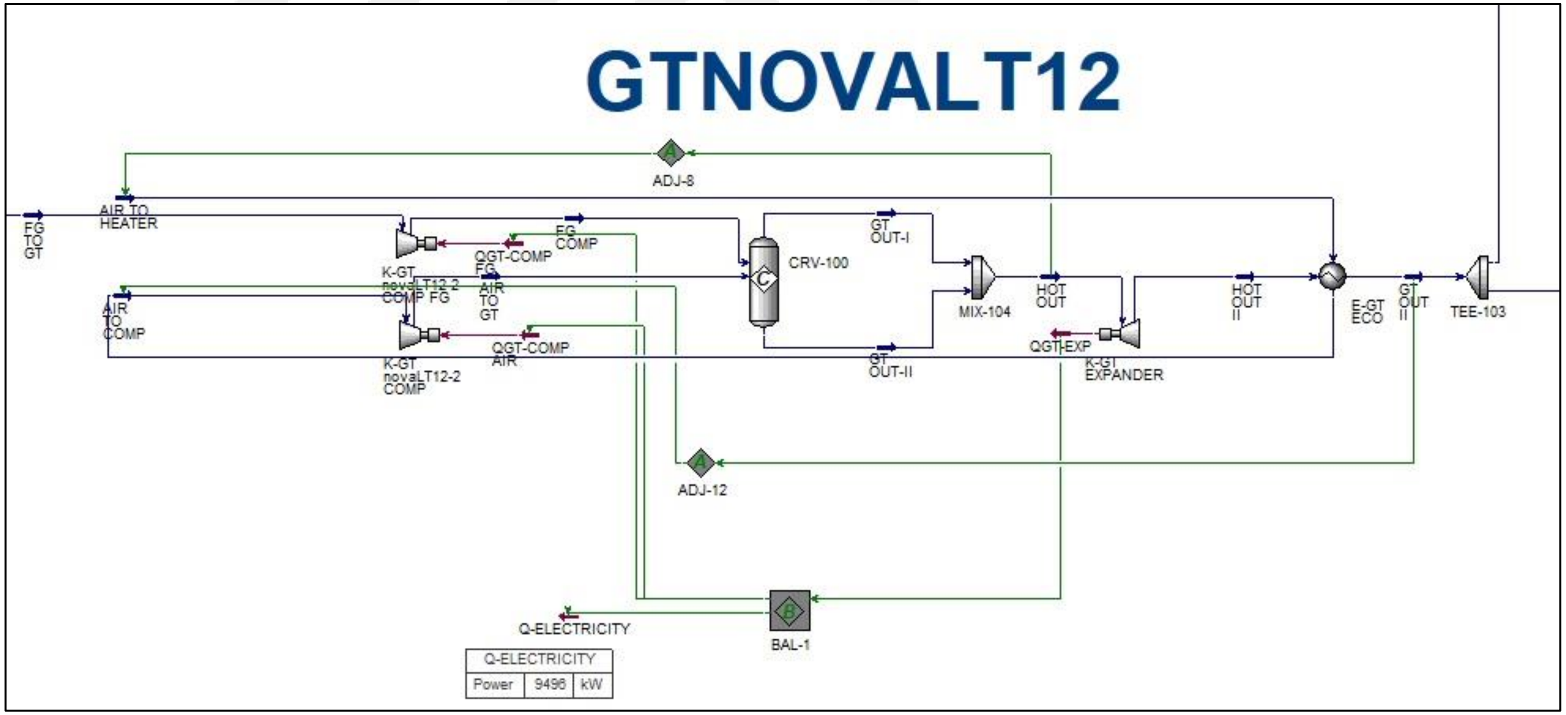

Figure A.1 GTG NovalT12 Modification in Aspen Hysys

Appendix 2. Furnace Simulation in Aspen Hysys

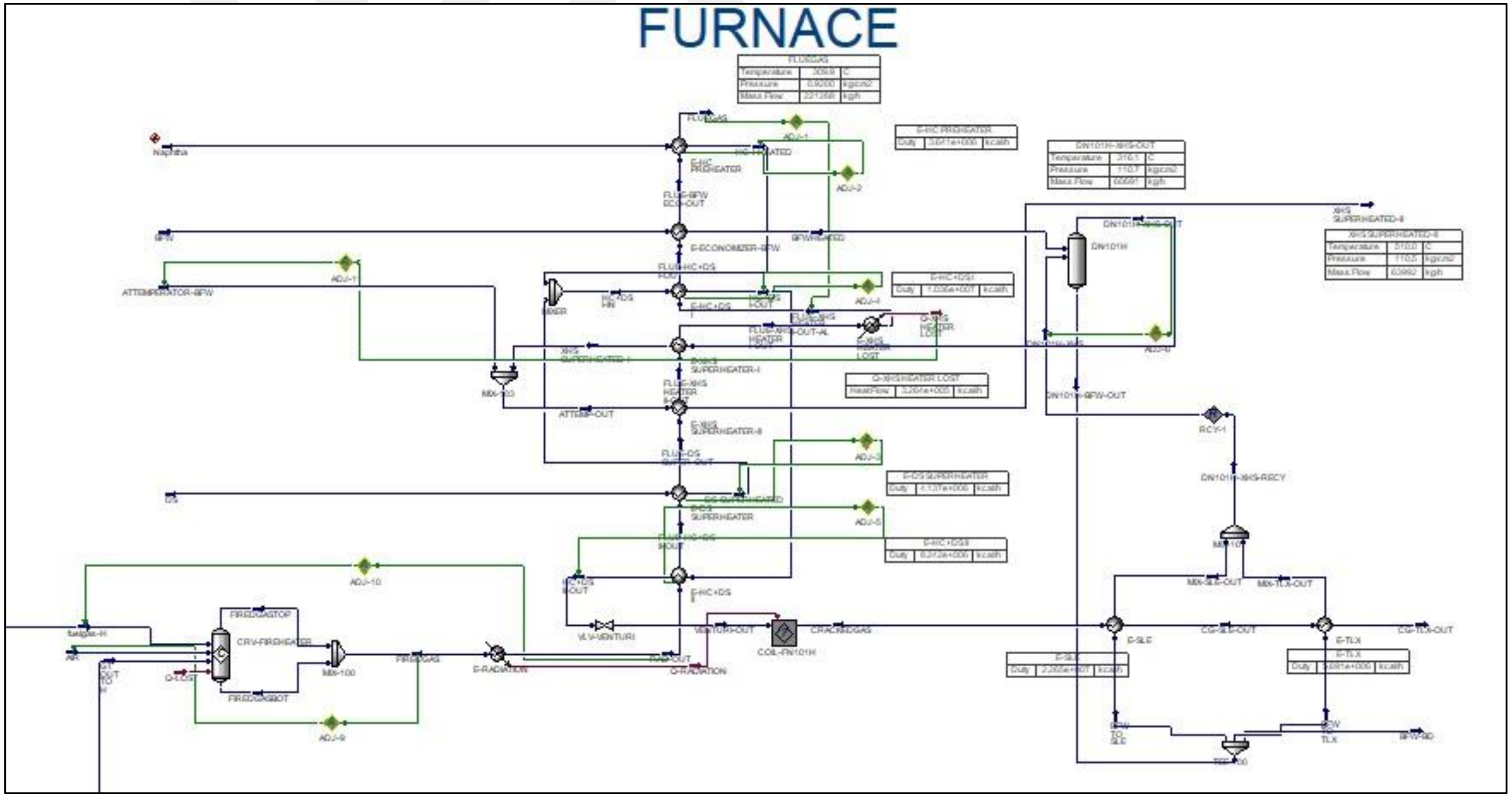

Figure A.2 Furnace Simulation in Aspen Hysys

Appendix 3. GTG Modification to Furnace Simulation in Aspen Hysys

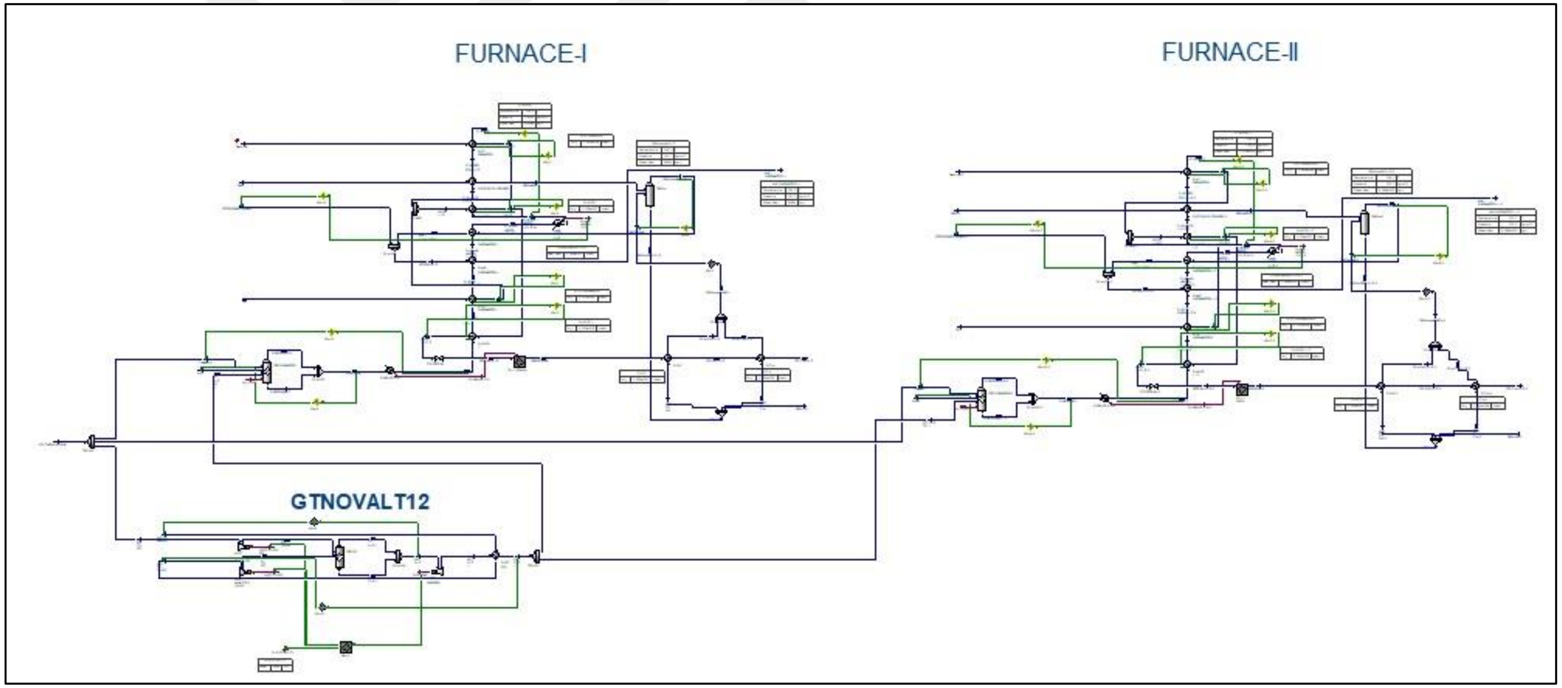

Figure A.3 GTG Modification to Furnaces Simulation in Aspen Hysys