



UNIVERSITY OF BAHRAIN

College of Engineering

Mechanical Engineering Department

Optimizing product yield and quality of a local NGL Plant using Model Predictive Control

A Thesis Submitted in Partial Fulfillment of the Requirements for the
Degree of M.Sc. in Engineering Management Program

Submitted by:
Yusuf Ahmed Al-Nafaiei

ID number: 20160021

Supervised by
Dr. Mohamed Bin Shams

University of Bahrain

May 2021



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Thesis Discussion Committee Decision

Student Name: YUSUF AHMED ALNAFAIEI	Student ID: 20160021
Department: Mechanical Engineering	College: Engineering
Program: Masters in Engineering Management	
Thesis Title: Optimizing product yield and quality of a local NGL plant using Model Predictive Control	
Supervisor Name: Dr. Mohamed Bin Shams	Student GPA: 3.77

The Discussion Committee met on 29-Jun-21 In Virtual Microsoft Teams. Upon discussing the student whose details are stated above in public, the discussion committee recommends the following:

The student is awarded a grade A in the thesis subject.

- ☐ Without making any changes to the thesis.
- ☒ After making the modifications attached with this decision.

Discussion Committee Members Approval:

Name

- | | |
|-----------------------------|---------------------|
| 1. Dr. Mohamed Bin Shams | (Advisor) |
| 2. Dr. Ebrahim Mattar | (Internal Examiner) |
| 3. Prof. Mohamed Bashammakh | (External Examiner) |

Signature

Mohamed Bin Shams
Digitally signed by Mohamed Bin Shams
DN: cn=Mohamed Bin Shams, o, ou, email=new_day_410@hotmail.com, c=BH
Date: 2021.07.06 00:56:40 +03'00'

Abstract

The objective of any plant whether operating in the oil, gas or petrochemical industry is to add value in the process chain through converting raw materials into more valuable products while subject to many constraints including but not limited to raw material availability, product prices, safety and environmental constraints.

To meet these constraints in an optimal manner, operators of these plants resort to control strategies and philosophies that would help them to better enhance production and meet quality constraints.

A Natural Gas Liquid Plant (NGL) located in the Kingdom of Bahrain with its recently commissioned Train – III has a total combined processing capacity of 660 Million Standard Cubic Feet per day (MMSCFD). Currently, the plant uses a traditional regulatory control strategy, namely, the Distributed Control System (DCS) to control its key process parameters. However, such control system lacks the ability to handle multiple interactive constraints explicitly as well as the ability to address the dynamic process interactions and hence provides sub-optimal solution to the constrained control problem.

This research aims to explore the potential of Model Predictive Control (MPC) as an advanced optimal control strategy to address some of the quality and production challenges currently encountered in operating the local NGL plant. In order to provide a realistic basis for this study, the studied NGL plant was simulated on three different levels which were necessary to achieve the objective of the work. Firstly, a steady state simulation of the plant has been built while incorporating design details of the plant and validated with real plant design data. Secondly, a dynamic simulation of the plant was later built on top of the steady state simulation in order to capture the transient behavior of the plant's process in response to process changes. The dynamic simulation built was also verified by conducting test simulations with scenarios derived from actual plant data and the response of which have been compared and verified. The validated dynamic simulation was then used to generate a model of the plant which is necessary for the design of the MPC. Lastly, a supervisory MPC was built above the dynamic simulation layer in order to achieve the several objective requirements of the different processing units. A framework to formulate an optimal control problem has been presented. This study demonstrates the superiority of the MPC in

achieving the different economic objectives of a train of distillation columns in the considered NGL plant namely – Deethanizer, Depropanizer and the Debutanizer units. Several operational scenarios – similar to what the real plant experiences, have been simulated in order to evaluate the robustness of the MPC to sustain its optimization performance. Several simulation scenarios have been considered such as changes in the feed rates and compositions. Comparative analysis showed the ability and superiority of the MPC system to constantly optimize the different objective requirements defined for each of the units. The study also presents an estimate of the quantitative benefits to be achieved for the subject plant. An annual incremental revenue of \$1,971,565 has been estimated when the production of the three main products of the NGL plant i.e., Propane, Butane and Naphtha is optimized. Further economic analysis into the investment required for implementation of the MPC system for the subject plant yields a payback period of 1.25 years.

Dedication

This thesis work is dedicated to my father and mother – Ahmed and Maria, who have supported and guided me my entire life. I will forever be grateful.

Acknowledgement

Alhamdulillah, I praise and thank Allah, the All Mighty for the many blessings I have received in my life including the opportunity and strength to complete my master's degree.

Firstly, I would like to thank my parents for the continuous support and guidance they have showed me throughout my education – without which none of this would have been possible. They have encouraged and provided me with strength from my first day in school until the day I submitted my thesis. I am extremely grateful for their persistent encouragement especially at times when I felt overwhelmed.

I would also like to thank my thesis supervisor Dr. Mohamed Bin Shams for his utmost and unlimited support he has shown me throughout the course of completing the thesis. I thank him for the time he has taken, for the countless nights out of his personal time we have spent discussing the technicalities of the thesis. He has pulled me out of technical difficulties several times. I am truly grateful to have had him supervising my thesis. I have started the journey of my thesis viewing him as my supervisor and ending the journey having made a friend.

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List of abbreviations

Abbreviation	Definition
ARX	Auto Regressive with Extra Input
BTU	British Thermal unit
C1	Methane
C2	Ethane
C3	Propane
C4	Butane
CS	Compressor Station
CV	Controlled Variable
CVP	Control Vector Parameterization
DCS	Distributed Control system
DEC2	Deethanizer
DEC3	Depropanizer
DEC4	Debutanizer
DMC	Dynamic Matrix Control
ESD	Emergency shutdown System
HIECON	Hierarchical Constraint Control
HMB	Heat and Mass Balance
IDCOM	Identification and Command
KPI	Key Performance Indicator
LIC	Level Controller
LP	Linear Programming
LPG	Liquified Petroleum Gas
LQG	Linear Quadratic Gaussian
LV	Level control Valve
MARR	Minimum Attractive Rate of Return
MIMO	Multiple Input Multiple Output
MMSCFD	Million Standard Cubic feet Per Day
MPC	Model Predictive Control
MPHC	Model Predictive Heuristic Control
MT	Metric Ton
MV	Manipulated Variable
NGL	Natural gas Liquid
NLP	Non Linear Programming
PFC	Predictive Functional Control
PFD	Process Flow Diagram

PIC	Pressure Controller
PID	Proportional, Integral, Derivative
PSIA	Pounds per Square Inch Absolute
PSIG	Pounds per Square Inch Gauge
PV	Pressure Control Valve
QDMC	Quadratic Dynamic Matrix Control
QP	Quadratic programming
RMPCT	Robust Multivariable Predictive Control Technology
ROG	Refinery Off Gas
RTO	Real Time Optimization
RVP	Reid Vapor Pressure
SCF	Standard Cubic Feet
SISO	Single Input Single Output
SMOC	Shell Multivariable Optimizing Control
SSTO	Steady State Target Optimizer
SQP	Successive Quadratic Programming
TEG	Triethylene Glycol
TIC	Temperature Controller
TV	Temperature Control Valve

Chapter 1 - Introduction

1.1 Introduction

1.2 Research problem

1.3 Research questions

1.4 Research objectives

1.5 Research significance

1.6 The subject plant

1.7 Lack of product quality optimization

1.8 Scope of work

1.9 Current plant control philosophy

1.1. Introduction

According to the Economic development Board of Bahrain, the oil and gas sector accounts for 13.35% of the GDP. A local Natural Gas Liquid (NGL) facility in the Kingdom of Bahrain with its recently commissioned Train-III has a total processing capacity of 660 MMSCFD. The plant feed consists of associated gas from 31 well manifolds, compressed gas from upstream facility and Refinery off Gas (ROG) from a nearby refinery. The plant recovers and refines propane, butane and naphtha and exports these products on a commercial basis.

The reduction of production costs has often been the driving force behind many capital improvements projects particularly with automation resources (Rebeiro et al., 2013). A proper process control system is very critical to ensure smooth operability while maximizing economic benefit and has the potential of either reaping benefits or causing significant losses to any processing facility.

Control system applications have tremendous impact on the overall performance of any facility. In the oil and gas field, control systems have been improving ever since the discovery of the earliest commercial oil well in the mid-nineteenth century (Habashi, 2000). With economic growth and development, the demand for a fuel that is cheap, has high energy intensity and relatively abundant meant that fossil fuel-based energy stocks have been increasingly becoming important in the past decades - succeeding coal as the primary source of energy in the 1960s (Kinderman et al., 2011). The oil and gas sector today is divided into three main categories: upstream (also referred to as exploration and production E&P), mid-stream (oil transportation and gas

distribution) and downstream (refineries, gas processing plants and petrochemical industries).

Majority of the plants operating today heavily utilize the ubiquitous Proportional-Integral-Derivative (PID) controllers to manage and maintain its daily processes. As will be shown later, these controllers implement control strategies that render the overall performance of the plant to suboptimal. PID control systems are robust and easy to tune controllers. It is usually implemented within a multi-loop Single Input Single Output (SISO) structures. This inherently decentralized structure allows each controller to work independently toward achieving its objective (i.e., keeping the controlled variable at a desired set point) while ignoring the effect of other loops. Each controller runs the risk of lack of coordination amongst the different control loops.

Apart from this, the complex nature of refinery, gas and petrochemical plant operations and given the fact that PID control systems are SISO systems, this will only result in hundreds and in some cases thousands of individual control systems running suboptimal. In addition, through experience, PID controllers have been known to go out of tuning with time and hence deteriorate its performance even more resulting in extra operational costs (Christofides et al., 2013). Although, there are many techniques for the tuning of multi-loop PID controllers, it still remains a difficult task mainly due to the ignored loop interactions (Johansson et al., 1988).

The aforementioned deficiencies necessitate the need for a more comprehensive approach that can explicitly handle process interactions, difficult dynamics e.g., inverse response and constraints in a more structured

and optimal way. Apart from the above-mentioned problems, there are additional issues pertaining to the way plants are operated that aggravates the control problem in operated plants, which can be summarized in the following points:

1. The impact of operator's performance on operating the plant – Plants are normally operated 24 hours a day. Organizational structures of operation teams differ amongst companies. However, the basic setup remains the same. Plants are generally operated by a team of operators. Outside or field operators are the personnel responsible for attending to the daily operational requirements on the field such as manually adjusting process parameters that are not actuated, maintaining of liquid levels by regularly draining vessels, switching of standby vessels during changeover and other daily tasks. Control panel operators (also known as control man) are the personnel generally responsible for overseeing the performance of the entire plant through the Distributed Control System (DCS). It is known within the industry, that operator skills impact plant performance and production. The task of control operators becomes particularly tedious and stressful when the process variables are highly interactive during transient behavior or during the occurrence of failure of critical equipment. Consequently, control operators are required to deal with events that are happening simultaneously. Pietilä, Remes and Haavisto (2011) conducted a study to identify the set of skills and technical requirements of plant operators that will lead to high operational performance. They have found that different operation-shift groups tend to differ in terms of production quality. Their analysis of several significant relevant performance indicators e.g., the economic

efficiency index, quality of feed concentrates and product recovery rates for different operator shift groups have showed that different operator groups performed differently. Li, McKee, Horberry and Powell (2011) conducted another study to examine the human operator element and their impact on the less-than-optimal performance of control systems. Their field study had concluded that significant difference in control performance and knowledge existed amongst different operators, and this made a real impact on the stability and performance of the production processes.

McKee (1999) examined several important aspects related to successful control systems such as instrumentation, process dynamics, process behavior, control strategies and overall maintenance practices. The study also highlighted the importance of skilled workers to operate the control systems. The report explicitly states, “There was some suggestion that control room operator’s performance at times imposes a constraint on the performance of the control system.”

Therefore, it is quite evident from the above that plant performances are heavily influenced by the operator’s performance in monitoring and controlling multivariate interactive processes and this in turn results in different overall production profiles.

2. Handling of disturbances

The periods of operational disturbances are the most challenging for operating workers in downstream plants. Disturbances can render plants to operate in a random manner while violating product and safety limits. This is especially true for disturbances originating at an arbitrary

upstream point and propagating to subsequent process units. The ability to compensate for these upsets and to stabilize the process depends – apart from the operator’s capabilities – on the deployed control system. In a survey conducted by Thwaites, it was found that 75% of physical industrial assets are under some process control worldwide – however over 60% of the control loops are found under performing (Thwaites, 2008).

Further exploring, 30% of the PID control loops are set on manual thereby laying the responsibility of deciding of controller actions on the control operator (Starr, Petersen and Bauer, 2016). Such an environment can hamper the decision-making process of the operators and hence rendering sub optimal performance of the plants.

The plants under study are being controlled using decentralized multi-loop PID controllers implemented within a conventional DCS system. As mentioned, the biggest disadvantage of this system is the inability to explicitly address process interactions and handling constraints is usually not straightforward. Another significant shortcoming of the popular DCS system is the inability to maximize economic benefits while satisfying hard operational constraints. The aforementioned deficiencies lead to loss of opportunities in terms of production yield. With pressure arising from changing demands and increased need for economic competitiveness, industries are increasingly forced to adopt better advanced operation/control strategies in order to remain profitable and competitive.

Accordingly, there is a need to evaluate a complementary solution that can boost the capabilities of the current regulatory system while taking into

account, within an optimization framework, market product specification, operational constraints and global NGL prices.

1.2. Research Problem

Downstream processing facilities in the oil & gas industry are continuously seeking to reduce operational costs and to enhance production profitability. Plants regularly seek a systematic and structured way to enhance production on a constant basis while satisfying product specification, minimizing energy costs while meeting environmental and safety constraints.

The current control system of the NGL plant under study lacks the ability to maximize economic benefits while satisfying quality product constraints. This results in financial losses in terms of production rates. In addition, the current regulatory system does not contain any provision for product-mix optimization based on current global NGL prices and hence presents a wasted opportunity.

1.3. Research Questions

- Does the implementation of Model Predictive Control (MPC) strategy increase production efficiency in the subject NGL plant i.e., production rate and reducing off-specification products while satisfying operational constraints?
- What are the tangible and feasible financial benefits that can be attained by adopting MPC for the local NGL under study?

1.4. Research Objectives

- To formulate the main NGL control problem within the MPC optimization framework to maximize benefit such as increasing the

yield of the valuable product e.g., optimizing product quality blends while respecting market specification constraints.

- To quantify the financial benefits of applying MPC to the fractionation section of NGL facility under study.

1.5. Research Significance

The research addresses a tangible problem that is currently being experienced in the NGL facility under study, that is, the loss of significant opportunity due to the inability of the current control strategy to address the aforementioned problems. The research will delve the benefits and the feasibility of implementing MPC to present concrete findings and recommendations. The research findings and recommendations will highlight the potential benefits of implementing MPC on the subject NGL facility and will serve as a guideline to the decision makers in the subject NGL facility and to any other party contemplating the benefits or suitability of implementing MPC.

1.6. The subject plant

The study will be conducted on a Natural Gas Liquid Plant (NGL) located in the Kingdom of Bahrain. A brief description of the plant, its process and current control system will be given.

The NGL facility has three processing trains with a combined capacity of 660 MMSCFD, which processes associated gas from well manifolds and Refinery Off Gas (ROG) coming from a nearby refinery. It extracts and markets mainly three products: Propane, Butane and Naphtha.

Train I & II operates at 310 MMSCFD and 440 psig while Train – III receives 350 MMSCFD of gas at a much higher-pressure of 1300 psig. The associated gas feeding Train II & I is separated from oil at 31 well manifolds. It is

compressed and dehydrated in 7 compressor stations (CS) located all around Al-Sakhir desert in the Southern Governate. The feed gas is transmitted from each compressor station at around 440 PSIG to both Train II & I. All of the ROG is processed in Train I&II through tie-ins up stream of the CS. Train I & II are interconnected for common services at different points such as the lean oil absorption system and the glycol dehydration unit for operational flexibility. NGL products are recovered using the chilled absorption system. The plants consist of the following units: Gas Dehydration, Condensate Stabilization, NGL Absorption, Deethanizer, Rich Oil Still, Depropanizer, Debutanizer and Desulphurization units. Liquefied propane and butane are transferred to refrigerated storage tanks located in one of the ports in the Kingdom of Bahrain where it is subsequently shipped. Naphtha is sent to the nearby refinery by pipeline, which is subsequently exported.

A huge amount of the processed gas leaves the plant as residue gas (mainly Methane at 76 mol% and ethane at 3.1 mol%) which is sent to a distribution center where the major consumers of this gas are an Aluminum Smelting facility, Oil refinery and Power Generation Stations.

Train-III which was commissioned in November 2018, uses different process technology. The plant has a total processing capacity of 350 MMSCFD of associated gas. The plant is also designed to handle 9 MMSCFD of ROG.

A block flow diagram of Train – I/II is given in the next page.

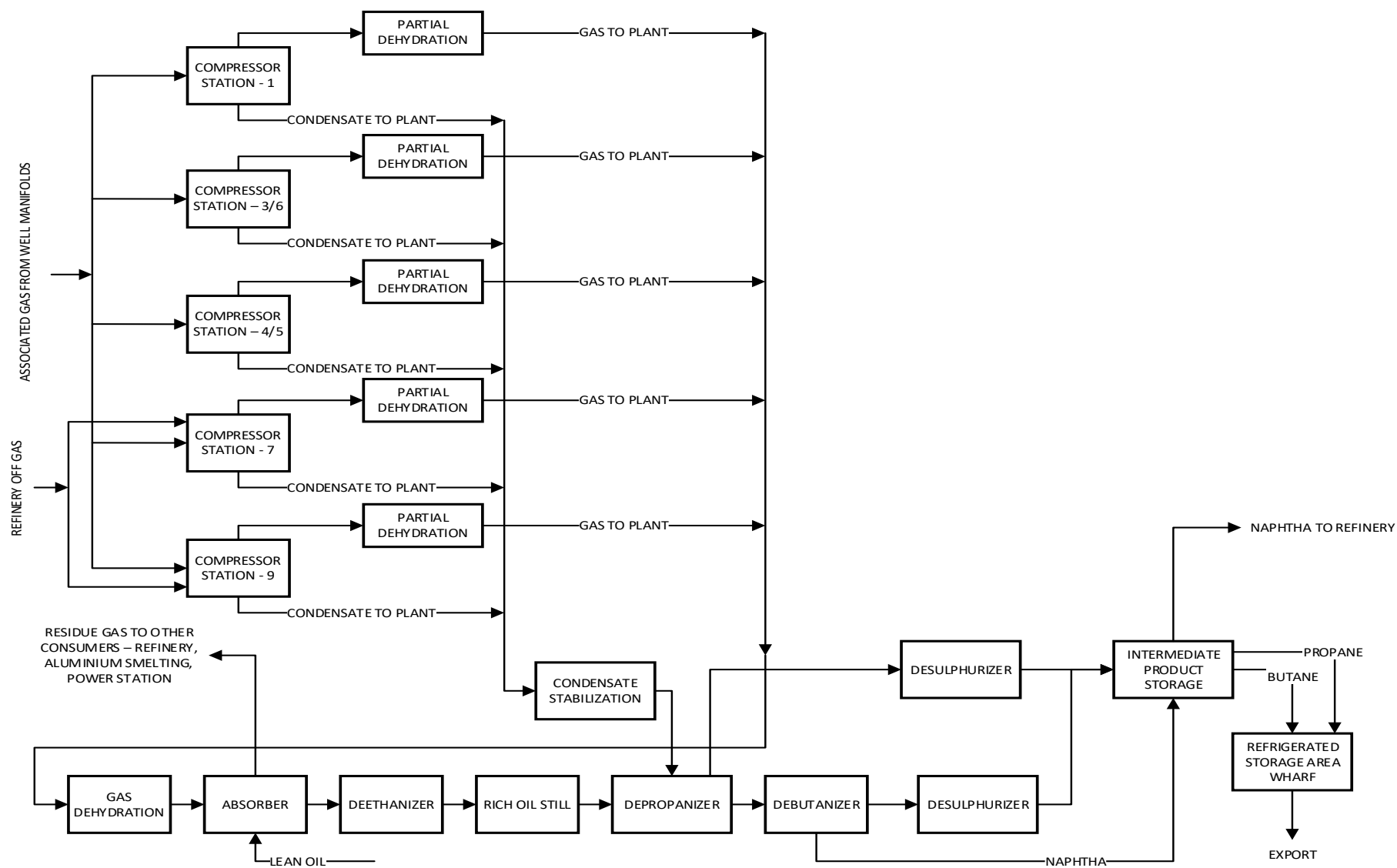


Figure 1 Block flow diagram for Train -I/II of the facility under study

1.7. Lack of product quality optimization

The current control configuration of the subject plant provides no optimization in terms of production quality. This presents an economic opportunity wastage in terms of optimizing production quality in order to maximize profitability. An analysis of the quality of the different products of the plant – propane, butane and naphtha, showed that significant financial losses is being incurred by the operator due to suboptimal production quality. To cite an example, the average Reid Vapor Pressure (RVP) of the naphtha product from the subject plant was found to be 10.12 psia with the limit being 12 psia. This represents a suboptimal mode of operation since increasing the RVP closer to the specification limit would entail optimal slippage of butane along with the naphtha product in the Debutanizer section and hence increasing profitability due to the fact that naphtha prices are higher than that of butane.

1.8. Scope of work

The scope of research work will be limited to Train-I. Train-I has a total processing capacity of 160 MMSCFD at 440 PSIG and is identical to Train – II in all aspects except in capacity. The scope of work will envelope the NGL fractionation units, which can be identified, from experience, as spots with substantial financial losses. Since both Train – I & II are identical in design, configuration and the feed gas quality, the results that will be concluded from the study of Train-I can be applied to Train – II.

1.9. Current plant control strategy

The control system of the plant under study consists of a Distributed Control System (DCS), which employs a large number of multi-loop (PID) controllers.

The DCS is generally a process-oriented system and is centered on the operator control station. Figure 2 shows the functional representation of the control levels of the system.

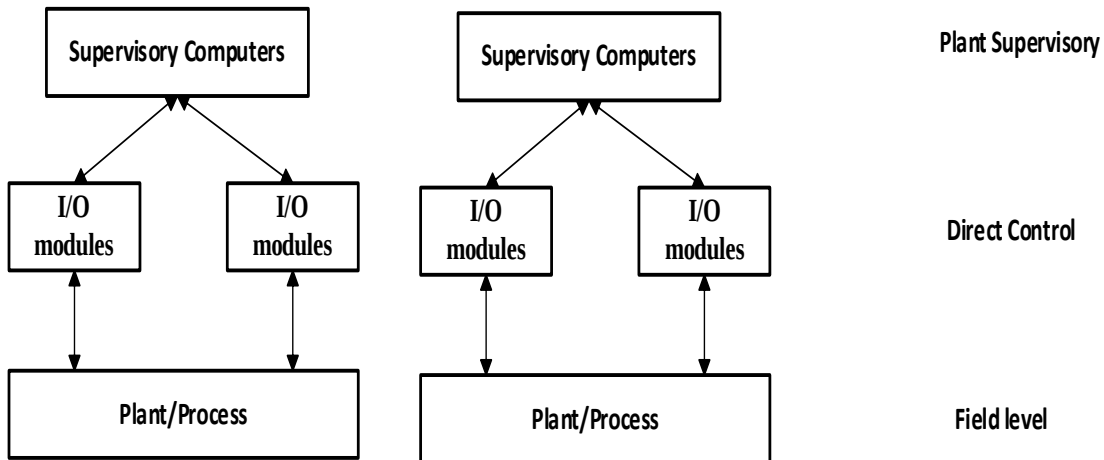


Figure 2 Functional schematic of the existing control system of the plant

The first level in the control philosophy consists of field instruments such as temperature sensors, flow meters, transmitters, and control elements such as valves. This layer is responsible for transmitting actual process parameters to the next level. The direct control level generally consists of the Input / Output modules along with the auxiliary electronic processors, which are distributed and not centralized. At this level, the controllers process and execute its set logic to provide output signals in order to achieve its individual controller objective. The next level is the plant supervisory level, which gathers data from all the processor nodes distributed in the system and provide information in the graphical interface available to the operator controlling the plant. It is

important to mention here that the MPC proposed in this thesis as an advanced process control strategy to the local NGL is not meant to replace or compete with the current conventional multi-loop PID controller implemented in the DCS system, but rather implemented in a hierarchal manner above the DCS layer as shown in figure 3.

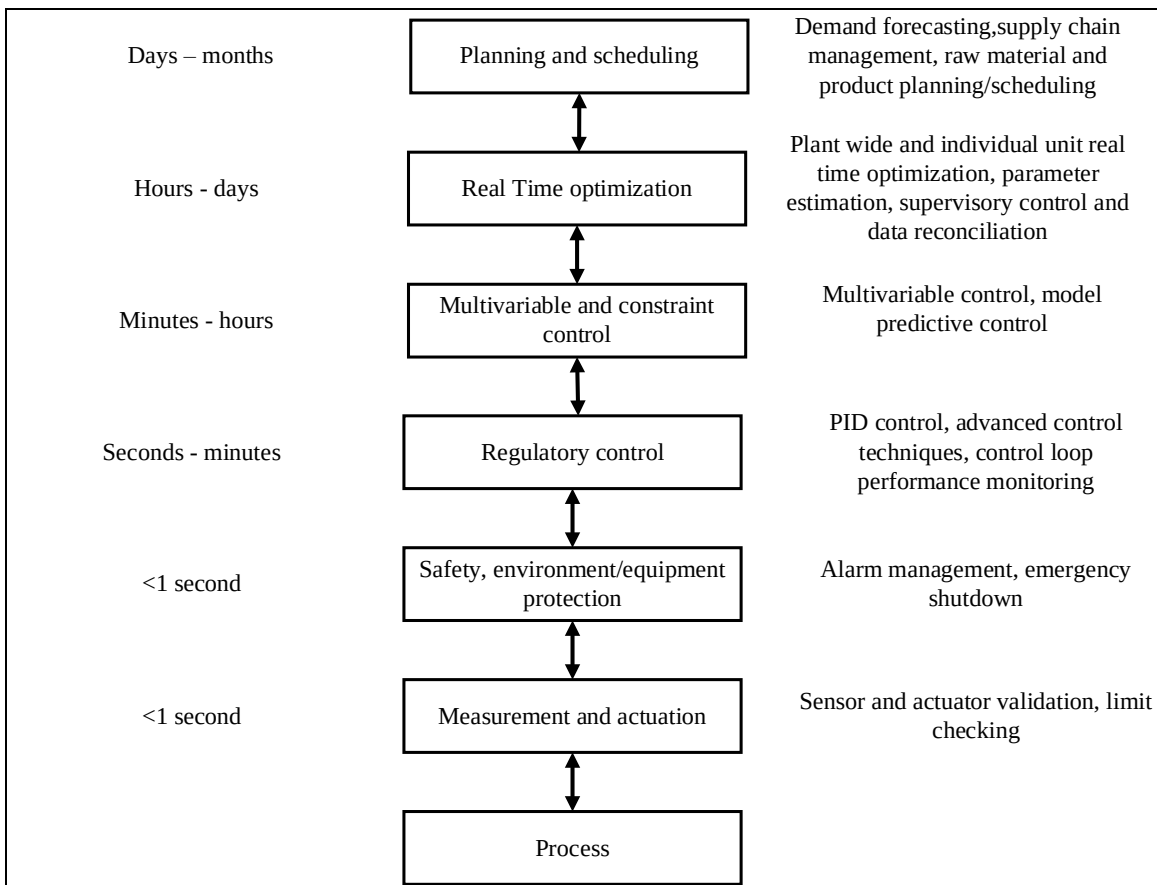


**Figure 3 The five levels of process optimization in manufacturing
(Seborg et al., 2004)**

Chapter 2 - Literature Review

- 2.1 Model Predictive Control MPC – Overview
- 2.2 Main components of MPC
- 2.3 Role of MPC in the plant optimization hierarchy
- 2.4 Identification of potential benefits
- 2.5 Benefit assessment methodologies – MPC
- 2.6 Base case identification
- 2.7 Commercially available technologies
- 2.8 Basis of prediction

Since its inception, Model predictive control (MPC) had received significant interest by academia and industries. Several reasons contribute to the success of MPC in industry: (1) the internal model that captures the important and difficult dynamics between the controlled, manipulated and disturbance variables, (2) the ability to handle input and output constraints in a structured manner, (3) the control calculation from MPC algorithm can be aligned with the economic optimum set-points of key process variables, (4) while used as a part of the control calculations, the internal model can also play a preventive role against potential failures. The aforementioned advantages enable elongated periods of operation with minimal operator's intervention.

Tuan et al., 2016 studied the effect of controlling the Depropanizer column using MPC on the product quality variation of propane. The study utilized a steady state simulation model converted to a dynamic model in Aspen HYSYS to generate the data required for model identification through ARX. The model parameters estimated through the ARX showed appreciable fit with the simulated dynamic model of the column. Results indicated superior setpoint tracking capabilities for propane concentration in both the top and bottom product. Even though the study stated that financial benefits and reduced time are expected (due to avoiding of real plant testing), the study did not conduct a cross validation of the dynamic model with actual plant data. Conducting a validation of the dynamic model is quite important in order to be able to verify model accuracy and hence would serve better benefit estimation. Also, the study lacked aggressive simulation testing scenarios that would have enabled to properly assess the advantage of operating an MPC on a Depropanizer column. The study was also limited to a single distillation column. This would

not enable a complete appreciation of the effect of MPC on a NGL fractionation train in terms of benefit estimation.

Luyben, 2013 presented a paper whereby effective control structures were developed and tested for a train of distillation columns in a NGL plant. A steady state simulation model was developed which was later converted to a dynamic model using Aspen HYSYS dynamics. The paper presented effective control structures in each of the Demethanizer, Deethanizer, Depropanizer, Debutanizer and De-isobutanizer columns. Effective production quality control was demonstrated using conventional PI controllers which enhanced stable regulatory dynamic responses. The simulation models developed were subjected to rather aggressive test scenarios mainly involving changes in both feed and composition for controller setpoint tracking was studied. However, both the steady state and dynamic models were not subjected to validation with actual plant data. The study did not consider the impact of application of MPC on the operational and profitability performance of the NGL fractionation train. Hence no economic benefit estimation was presented that could have given an idea of the amount and extent of the performance improvement by the MPC.

Fikar et al., 2000, conducted a study where an industrial Depropanizer multi component model was analyzed in order to optimally determine output trajectories that would minimize the transient time to reach a steady state. They have utilized a model developed in 1992 that uses a system of differential equations to simulate the thermodynamic and hydraulic performance of the column. The objective function defined was the reduction of the time from one steady state to another in case of changeover. The optimization method

used is the control vector parameterization (CVP) that involves converting an optimization problem into a nonlinear programming (NLP) problem that then is solved using successive quadratic programming (SQP) to find the optimal profiles of the manipulated variables. The study, however, does not provide a structured approach to decide the optimal controlled variables for the changing scenarios. In all cases, the objective function incorporates a desired point of the controlled variable and assumes this value to be the most optimum. Also, the study does not quantify the benefits nor does it provide an approach to validate the dynamic model used in the study.

There are papers that have explored the adoption of advanced control strategies for columns. But the author has not been able to find a study that explores the economic impact of operating MPC on a complete NGL fractionation train taking into account cross validation of the steady state and dynamic state models developed with real plant data. The author has also not found any paper that aggressively conducts test simulation based on real plant experiences. The only study that was found with the approach of studying a NGL fractionation train was that of Luyben, 2013. However, the study was done with the intent of presenting a control strategy without addressing MPC nor the economic benefits.

As far as literature records, the development of MPC can be traced back to the early works of Kalman in the beginning of 1960s (Kalman, 1960) with linear quadratic Gaussian (LQG). His work at the time, assuming a discrete-time linear state space model, was designed to minimize a quadratic objective function consisting of state deviations from the origin and squared input. The LQG algorithm adopted an infinite prediction horizon strategy and was

observed to have powerful stabilizing properties for any reasonable linear process (Qin and Badgwell, 2003). However, it had little impact in the industrial and academic world for several reasons.

The formulation had no provisions to incorporate constraint handling. A successful implementation of such technologies would require a system that can maintain parameters as close as possible to constraints but without violation. In addition, the process models required had to be developed from fundamental physics and chemistry principles, which is a tedious task and is very difficult to economically justify such investments. This quintessentially meant that the technology would receive little success and acceptance amongst control experts (Richalet, Rault, Testud and Papon, 1977). The conditions prevailing in the industry led to an environment that would require a development of a more general based control approach in which a dynamic optimization problem is solved at each control execution and from which plant dynamics can be described as an explicit model taking any mathematical structure (Qin and Badgwell, 2003). All these factors led to the development of more flexible and robust control strategies that are known today as MPC. In the 1970s Richalet et al (1978) reported successful application of model predictive control. The approach that was described at that time known as Model Predictive Heuristic Control (MPHC). The approach had the distinguishing feature of utilizing a dynamic model for the plant (impulse response) to predict the effect of the plant for a suggested future control action that was obtained by solving a quadratic performance objective formulation function on a finite prediction horizon. Another distinguishing feature of this approach was the ability to include constraints for both the input and output parameters. The quadratic objective function was solved using a heuristic iterative algorithm.

Later in the 1970's a different MPC technology approach was developed by engineers at Shell Oil. Cutler and Ramaker in a paper presented in the 1979 National AIChE meeting revealed details on an unconstrained multivariable algorithm, which was called Dynamic Matrix Control (DMC) (Cutler and Ramaker, 1979). Soon after in 1980 Prett and Gillette presented details of modified DMC application to a Fluid Catalytic Cracking Unit (FCCU) (Prett and Gillette, 1980). The modification to the algorithm allowed the system to handle nonlinearities and constraints. The DMC uses a linear step response of the plant to predict the effect of a certain future input obtained after solving a quadratic performance objective on a finite prediction horizon. Multiple step response outputs were combined using superposition and the link that ties the inputs to the predicted output was called the *Dynamic Matrix*. The aim of the quadratic objective function was to bring the output as close as possible to the setpoint utilizing a least square method and penalized the manipulated variables MV movement thereby resulting in a controller that was less aggressive.

The initial MPHC and DMC algorithms represent the first generation of MPC technology,

The second generation of MPC came with the introduction of Quadratic Dynamic Matrix Control QDMC. The previous MPHC and DMC lacked in terms of constraint handling. Modifications by engineers at Shell Oil on the DMC provided a way to deal with constraints. However, it was a heuristic approach in which the input and output constraints appeared explicitly (Qin and Badgwell, 2003). The QDMC algorithm was first presented by Cutler et al. in a 1983 paper at the AIChE conference in a paper titled "*An industrial*

perspective on advanced control” (Cutler, Morshedi and Haydel, 1983). Years later Garcia and Morshedi 1986 presented details on a more comprehensive methodology. The main features of this approach involved the use of a quadratic programming to solve for a constrained control problem utilizing a linear step response model, a finite prediction horizon with control and state constraints defined as linear inequalities.

With time and with increasing popularity of MPC in controlling different challenging applications, different variants of MPC algorithms were developed. The QDMC approach was praised in a way because of its ability to incorporate constraints; however, there was no clear way to ensure the feasibility of the solutions. Certain feedforward disturbances can render the infeasibility of the solution for which QDMC does not provide a way to deal with (Qin and Badgwell, 2003). In real applications, process input signals (MVs) and output control signals (CVs) can be lost for several reasons including electronic hardware failure, suboptimal performance of valves and manual intervention from operators. This can cause a system to move from an optimization problem with conditions producing a unique solution to over-determined or under determined systems – which all depend on the number of MVs compared to the CVs. In situations where the MVs match the CVs, the system is called a square plant and has unique solution – which are rare cases in real plant applications. The common fat plant case is where the number of MVs available are more than the CVs also called over-defined systems. Here the extra degree of freedom can be employed to achieve additional control objectives. As mentioned earlier, several different reasons can result in the CVs being more than the MVs (thin plant case), in which it is not possible to meet all of the control objectives. The control specifications will need to be

relaxed in order for the controller strategy to work. In addition, increasing difficulty has been faced when trying to convert control requirements into relative weights in a single objective function.

All these reasons paved the way for the third generation of MPC algorithms such as the Shell Multivariable Optimizing Control (SMOC) that utilizes state space model, general disturbance models and state estimation Kalman filters (Mayne, Rawlings, Rao and Scokaert, 2000).

Other third generation MPC algorithms examples include the Identification and Command (IDCOM) and Hierarchical Constraint Control (HIECON). The third generation of algorithms have a common feature of distinguishing between different levels of constraints. It also provides the ability to recover from an infeasible solution

2.1. Model Predictive Control – Overview

Several MPC algorithms started appearing in literature since 1960s with each of the algorithm providing certain advantages in issues faced in the prior algorithms. Nevertheless, MPC in its core consists of 5 different elements which are – Predictions, Receding Horizon, Modelling, Performance Index, and Constraint Handling. The MPC can be described as an anticipation strategy where it executes control moves that are expected to deliver the most optimal solution over a fixed or limited future time horizon.

One of the most common reason cited for the success of MPC in industries is that the most profitable operation of a plant is commonly found when the process is operating at or near to the constraint boundary.

The basic principle of MPC is to determine the most optimal future sequence of control inputs, over a future time horizon to achieve an optimized performance. Determining the predicted future sequences is based on an

internal mathematical model, which can be described as the heart of the system. Given the most recent process/plant measurements, the algorithm develops predictions of the future behavior of the system. The system then solves a particular minimization objective function and hence gives the initial control input, disregarding the rest of the future control moves – that is required to achieve the most optimal performance identified from the solution. The same process is continuously repeated at predefined intervals and at every interval; the horizon is moved one time step into the future. Given that no model will be completely accurate in predicting outputs, MPC algorithms also have a mechanism to reduce the discrepancy between the internal model and the actual system behavior, hence involves a feedback provision.

MPC derives its success from the ability to include operational and engineering constraints in a structured way. However, it also presents other advantages such as ability to handle non-minimum phase processes (i.e., system with significant time delay or inverse response), changes in system parameters and its suitability for application to large multivariable processes (Richalet, 1993).

The MPC usually contains the following concepts:

1. Usage of an explicit model of the process to predict the process output over a fixed future time horizon.
2. Determination of a specific control sequence derived from solving an optimization problem (minimization of an objective performance index).
3. The concept of receding horizon, which involves application of the first control input previously calculated and thereby moving the horizon at each instant towards the future. The aforementioned concepts ensure an

efficient and profitable operating approach, which usually involves operating the system close to its constraints.

Furthermore, the MPC strategy approaches present a strong and robust way of directly reflecting many performance criteria, which are relevant to the process and can utilize a variety of process models. Hence there is no restriction in terms of the model, the objective function or constraint application (García, Prett and Morari, 1989).

Due to the above, MPC has become a popular approach in industries to control system with complex dynamics and interactive multivariable control systems (Darby and Nikolaou, 2012).

The MPC approach can be characterized by the strategy shown in figure 4. The figure illustrates the case of a Single Input Single Output (SISO) with the horizontal axis representing a discrete time. The current sampling time is given as k and a decision about process input $u(k) = u(k|k)$ has to be determined. The variables that are required for all of the calculations is presented in as trajectories (future) of the process control input and output. The figure presents two output responses and two input trajectories with a reference setpoint for comparison.

The model description is done with the following in consideration:

1. The controller sampling interval is given as T_p
2. There is a dynamic model and a model for the constraints.
3. Current and past process measurements are available together with the past control inputs.

4. Setpoints are known and should be defined by the operator beforehand for an assumed prediction horizon (or sometime identified by upper real time optimization layer).

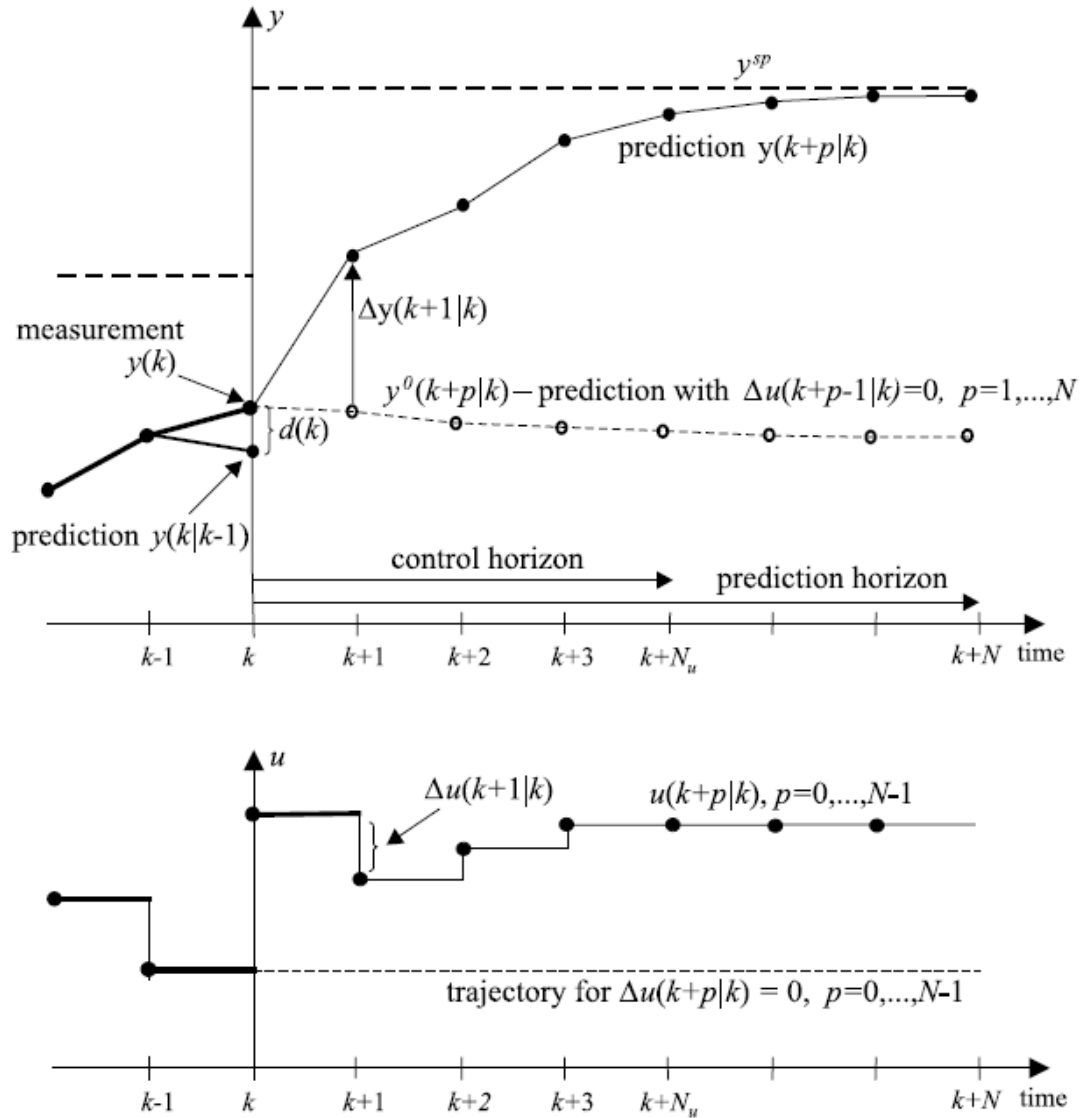


Figure 4 Model Predictive Control Strategy (Tatjewski, 2007)

The control input calculated at the given instant time k is represented by $u(k) = u(k|k), u(k+1|k), \dots, u(k+N_u-1|k)$ and is the decision variable which needs to be calculated. The control horizon is denoted by N_u and the prediction horizon by N ($p = 1, 2, 3 \dots N$) for which $N \geq N_u$. The notation $u(k+p|k)$ represents the prediction of the future input value at the sampling moment $k+p$, calculated at the current time k . The predicted future system responses is denoted by $y(k) = y(k+p|k), p = 1, 2, 3, \dots, N$, and the term denotes the prediction of the future output value for the sampling period $k+p$, calculated at the current time k .

The predicted future response $y^0(k+p|k), p=1, 2, \dots, N$, given by the dashed line, represents the situation where the process input is maintained constant over the prediction horizon with the values of $u(k-1)$ determined at the preceding sampling interval i.e. $u(k+p-1) = u(k-1)$ at all $p = 1, 2, \dots, N$. The trajectory defined this way shows that the future outputs depend on the previous inputs however there is no real influence on this trajectory at the current time k . Hence, this is often referred to as the *free component* of the predicted future response.

The predicted future response trajectory $y(k+p|k)$ is dependent on both the past input up to $u(k-1)$ and also the future predicted input $u(k+p-1|k) p = 1, 2, \dots, N-1$ which are calculated at the current sampling time instant k . The trajectory developed as a result is given in the figure as continuous solid curves. These values depend on the model of the process, which is used to calculate or predict the future input responses. The models differ in types however all models will only represent a good approximation of the real system responses. In reality, there are disturbances, which can either be

measured partly or not measured at all, and these disturbances cause a difference in the predicted output from the actual system output. The unmeasured disturbance can be estimated as $d(k)$, where $d(k)$ is given by $d(k) = y(k) - y(k|k - 1)$ which is calculated at the sampling time k .

The set point or the target reference is given by $y^{sp} = y^{sp}(k+p|k)$, $p = 1, 2, \dots, N$ and it is represented in the figure as thick dashed lines. The set point trajectory in figure 4 changes at sample time k , however this varies according to the operator and can remain constant over the entire prediction horizon or it can vary considerably.

Decision on the control inputs for the current sampling time k and for the future time in the control horizon is accomplished by minimizing some performance index so that the predicted response moves to the setpoint in an optimal manner. An important element describing this performance index or also known as cost function is the cost of variations of the predicted future outputs from the set points. The other significant element in the cost function is the inclusion of penalties for control input changes. Taking into account these elements, a generally utilized performance index is a quadratic cost function (objective function) which is formulated and optimally solved to determine the trajectory of the inputs over the control horizon. The cost function can be described as follows:

$$J(k) = \sum_{p=1}^N ||y^{sp}(k+p|k) - y(k+p|k)||^2 + \sum_{p=0}^{Nu-1} ||\Delta u(k+p|k)||^2 \quad (2.1)$$

The differences between the predicted response and the setpoints $y^{sp}(k + p|k) - y(k + p|k)$ are calculated starting from time instant $k + N_1$ where $1 \leq N_1 \leq N$ i.e. till the end of the prediction horizon. A value with $N_1 > 1$ can be considered reasonable if there is delay expected in the process causing a lack of reaction of response at time interval $N_1 - 1$ sampling instants, $k + 1, \dots, k + N_1 - 1$ to the shift of the control input at instant k . However, assuming $N_1 = 1$ is not an error and in many literatures, sources assume $N_1 = 1$ for the purpose of uniformity.

The basic idea is to use time-domain step-response model of the process to calculate the future changes (control moves) in the manipulated variable that will minimize some performance index so that the predicted response moves to the setpoint in an optimal manner.

2.2. Main components of MPC

MPC uses a prediction/anticipation strategy to optimally calculate control inputs over a fixed future control horizon. The following constitute the main components of MPC:

1. Receding Horizon

The length of the prediction horizon remains constant. However, it is relative to the current time and hence “recedes” away so that at each instant the horizon moves towards the future. Figure 5 illustrates the concept of receding horizon. The receding horizon time length is t_r . At time t , the controller solves an optimization problem defined for the time interval $t + t_r$. For the next time interval at time $t + 1$, the controller

solves an optimization problem for the next prediction horizon domain i.e. $t + 1 + t_r$. Hence, the concept of receding horizon moving constantly into the future however fixed in relation to the current time.

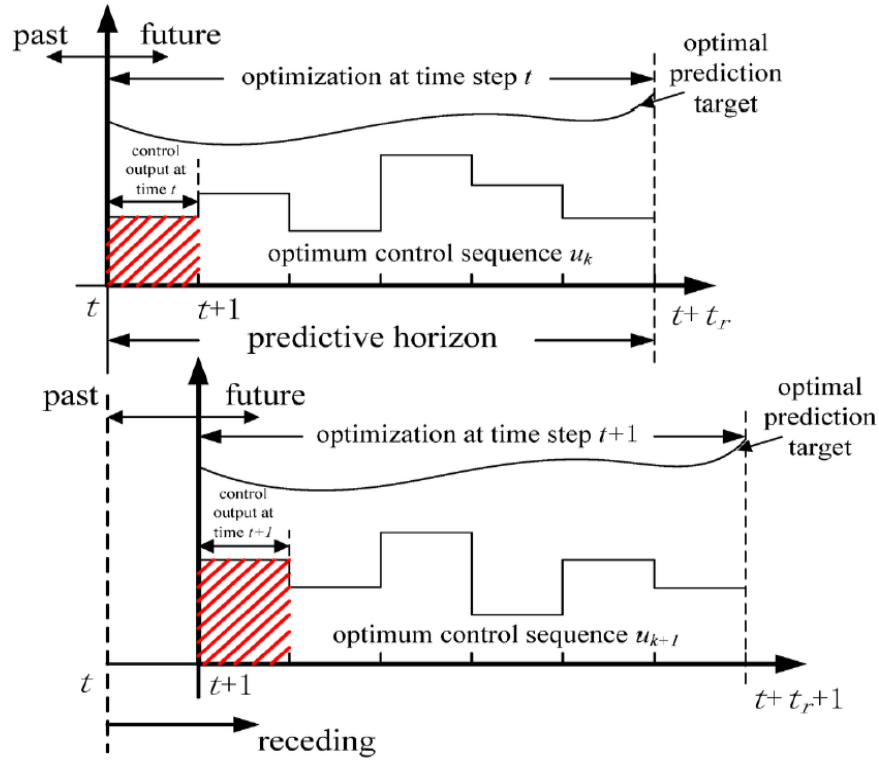


Figure 5 Receding horizon principle (Hu et al., 2015)

2. Prediction

To solve any MPC formulation, there must be a method of computing the predicted future outputs/responses of the desired controlled variables $y(k + p|k)$ from the best estimates of the current states, the assumed input $u(k|k)$ and assumed output $u(k + p - 1|k)$. The selection of this prediction methodology is considered another “tuning parameter and will have a significant impact on the performance of the closed loop system (Orukpe, 2012).

3. Modelling

A mathematical model that will allow the representation of the process is required to automate the predictions. Usually, simple models that can give accurate enough predictions while capturing key dynamics are normally preferred (Rossiter, 2005). The correction of such errors can usually be done through feedbacks (Al Naumani, 2017).

4. Constraint handling

The feature is arguably one of the most important features of MPC and that is the ability to explicitly incorporate constraints into the controller optimization problem. It is well known that the economic operating point of a typical process unit often lies in the intersection of constraints (Prett and Gillette, 1980). Therefore, a controller optimization system that have a provision for incorporating constraints will be of enormously beneficial and will provide a way operate unit in a more economical and safe way.

5. Performance Index

The performance index encapsulates control objectives which is to be minimized in the MPC calculations. The objective function constitutes elements pertaining to the difference between the predicted trajectory and the set point or desired state. It also incorporates an opportunity to penalize control inputs shifts thereby minimizing controller extra control efforts. It is also possible to incorporate different weights into

the performance index thereby translating the relative importance of each parameter into the objective function.

Quadratic performance objectives functions are commonly used since they provide for a simple, robust optimization and a unique minimum with generally smooth behaviors. Complex performance objective functions can affect system robustness and ability to deal with uncertainty and hence not desirable (Al Naumani, 2017).

2.3. Role of MPC in the plant optimization hierarchy

The MPC control layer is part of a multi-level order of control layers. Figure 6 shows a typical hierarchical order of control structures. At the top of the structure sits the layer responsible for the global economic optimization also known as scheduling. This layer deals with overall operational targets such as production targets, production timings, required production yields or qualities etc. Decisions from this layer is later communicated to the local economic optimization layer. It provides the set points required in order to achieve the goals of the global economic optimization layer. The unit optimizer computes an optimal economic steady state value, typical time scale of once an hour and then communicates this to the next layer below which is the dynamic constraint control that incorporates the MPC

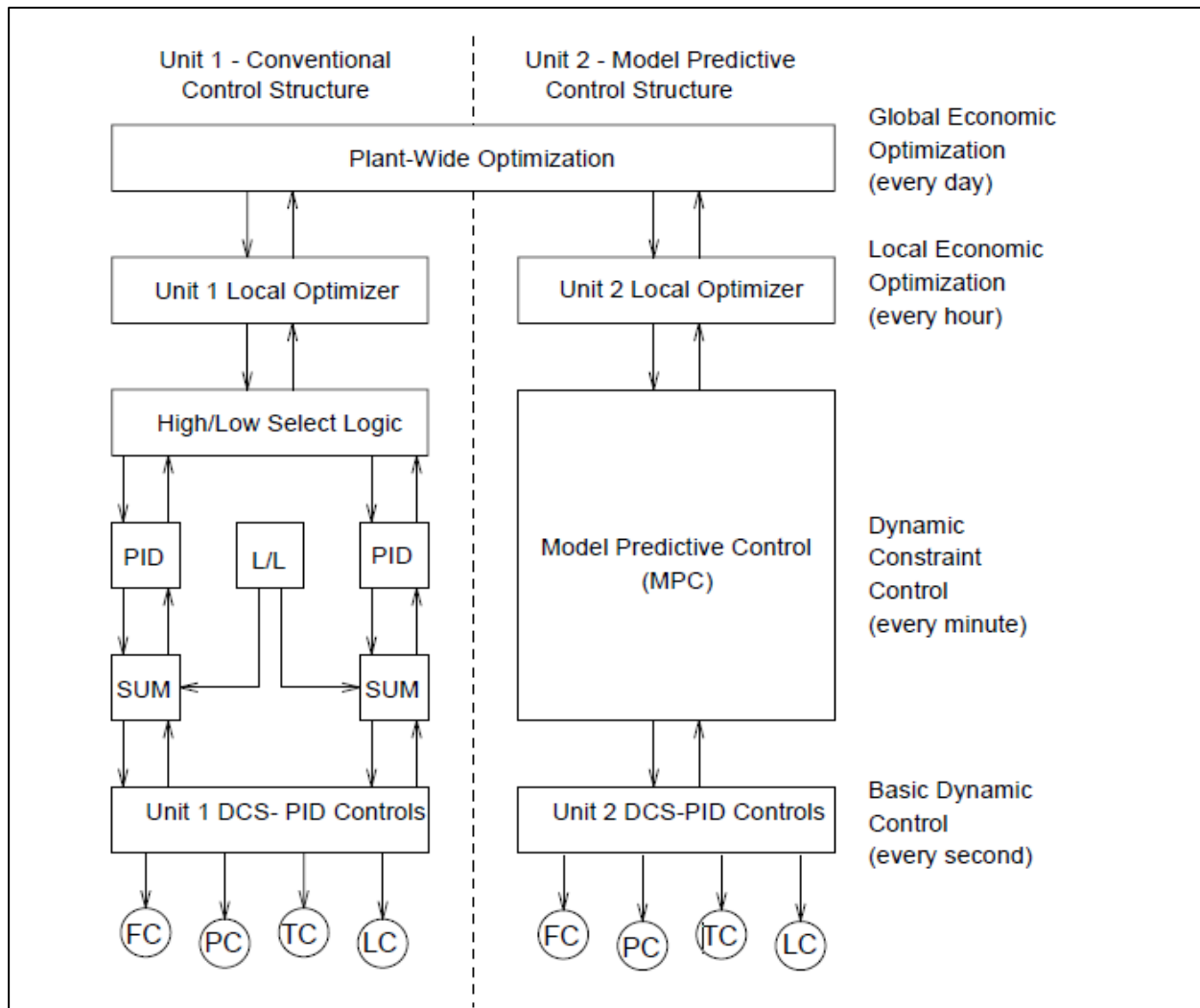


Figure 6 Hierarchy of control system - Planning and Operation, (Qin and Badgwell, 2003)

The MPC block is responsible for calculating the optimal control actions that need to be sent into the layer below (direct control) in order to bring the plant (process parameters) as close as possible to the identified setpoints from the RTO. This block is where the dynamic model of the plant is incorporated considering various operational and physical constraints thereby ensuring stability. Typical time scale is in the order of minutes.

However, a problem with this hierarchy is that the optimal setpoints calculated from the RTO layer can be inconsistent, unreachable or does not agree with

the dynamic model located in the MPC – mainly due to the differences between the RTO steady state model and the MPC dynamic model. Adding to this is the fact that both layers utilize very different time scales and hence the optimal setpoints calculated by the RTO may not coincide with dynamic model of the plant (Maestre, 2014).

A way to solve this problem is by incorporating another layer in between the RTO and MPC called the Steady State Target Optimizer (SSTO) which serves to calculate the steady state to which the system needs to be stabilized, by solving a linear or quadratic programming problem (taking into consideration data from the RTO. The plant model used in this intermediate layer is similar to the MPC thereby reducing inconsistencies. (Maestre, 2014).

However, taking a look at the left side of figure 6 presents the traditional/conventional way of control structure hierarchy. For a very long time instead of forwarding the setpoints to the MPC layer which calculates the most optimal action in order to achieve objectives, a PID layer is used instead. This layer uses a combination of PID algorithms, lead-lag blocks (L/L) and high/low selector logic in order to accomplish its objective. This however presents a difficult method to translate control requirements at this level into a suitable conventional control structure (Qin and Badgwell, 2003).

The PID controllers are good in trying to achieve its individual setpoint objectives however it takes no account of the dynamic interaction it plays and hence an action of one controller counters the action of another. PID controllers are able to keep its individual setpoints at the desired setpoint for a short time and has the ability to reject disturbances. However, PID controllers are very difficult when it comes to multivariable (multiple input multiple output) configuration.

The aforementioned reasons have led to a practice in the industry and that is to use the MPC on top of the PID/DCS as an advanced controller which usually can act as a standalone centralized unit.

2.4. Identification of potential benefits

Assessing the economic benefits of any planned modification or improvement will first require identification of the potential benefits. The key objective of advanced process control system (APC) in any industrial plant is to stabilize the process and achieve better reduction in variation. However, there are other benefits resulting from such improvement that can be quantified. Bauer and Craig (2008) in a study to develop an economic assessment framework for advanced process control projects – have conducted a survey amongst key MPC users and providers and have identified several benefits which will be summarized as follows:

2.1.1. Increased throughput

Under the assumption that additional feed raw materials are available and that the increased production will not alter market dynamics hence increased production can be sold at the same price – MPC increases throughput by approximately 5-10%. In a different study, Canney (2004) reports the increase in throughput by 3-5%. Generally, many continuous processes are constrained physically, and a trade-off exists between throughput and quality (Bauer and Craig, 2008). Hence increasing throughput without compromising quality will contribute towards the profitability of the plant operations.

2.1.2. Process stability improvement

Process stability is linked to the concept of variation reduction. Increasing process stability ensures that the products meet the required specification consistently with the process under smooth operation. The survey conducted by Bauer and Craig (2008) have found out that approximately 68% of MPC users reported achieving economic benefits from improved process stability.

2.1.3. Reduction in energy consumption

Industrial processes are generally very energy intensive with utility costs forming a significant part of operating expense. Management of utility/energy networks is a complex issue and achieving better control will most definitely contribute to reduction in utility consumption. Processes such as power generation, metal refining and petrochemical production facilities form good candidates for potential utility savings arising from implementation of MPC (Bauer and Craig, 2008). Their survey showed that 46% of the MPC users reported reduction in energy consumption as one of the top three benefits realized from MPC.

Energy savings can range 3-15% depending on the process and the current operational methods. Especially in steam generation, MPC can help match steam supply to steam demand and help reduce excess steam venting and pressure let downs in order to maintain system header performance. Companies have reported 60-90% reduction in excess steam venting post implementation of MPC (Singh and Kulik, 2016).

2.1.4. Increased yield of more valuable product and reduction of product giveaway

Implementation of MPC can result in an increase in the yield of the more valuable product while reducing product giveaways. Product giveaway is the difference between the quality of the product and quality specification. Producing better than specification means increased operating expenses on the manufacturer which are not necessary.

Minimizing product giveaway quality in a low margin and highly competitive market such as clean petroleum fuel products market can mean the difference between continuing operation of a plant or shutting it down. At a typical refinery, optimized blending could represent the majority of savings achieved due to MPC (Seiver, 2015). A 100,000 barrel per day refinery with production of gasoline reaching 45% of feed, a giveaway in Reid Vapor Pressure RVP specification with a standard deviation of 0.5 Pounds Per Square Inch PSI can result in losses up to \$7.6 million annually (Agrawal and Agrawal, 2018). The survey conducted by Bauer and Craig (2008) reports that around 41% of MPC users reported increased yield of the more valuable product as one of the top three benefits attained from MPC while around 60% of MPC suppliers reported it as the same. A possible explanation given by the authors for this difference reporting between MPC users and suppliers is that increasing the yield of the more valuable product is a long-term market position achieved through consistently performing better and does not reflect in the day-to-day operations.

Many organizations have reported an improvement in production yield between 1-2% through MPC installations. For plants with reactors, such benefits are attained by optimizing the operation of the temperature control. For the physical separation section of the plant, the improvement in production yield is attained mainly by increasing the amount of the lower value product while maintain contracted specifications (Singh and Kulik, 2016).

2.5. Benefit Assessment Methodologies – MPC

With wider application and advancement in the process control industry more advanced techniques for evaluating the benefits, arising from better control needs to be developed. However, control literature provides little information on assessment techniques as no significant attention has been given to linking monetary benefits to better control systems (Craig and Henning, 2000). Nevertheless, there are notable exceptions in this regard and as such, a brief review of these shall be done in this section.

From literature, most cases of economic benefits (process related) can be directly attributed to reduction in process variance thereby allowing the process to run closer to the constraints/specification limits with lesser violations. The heart of benefit analysis is derived from the ability of control system to reduce the variability of the key process variables and drive them closer to their targets (Marlin, Perkins, Barton and Brisk, 1991). Deming (1986) has described two main causes for producing off specification

(exceeding constraints) production, which he attributes to common causes and assignable causes. Common causes are sources of disturbance hence variability, that are embedded in the process nature or method of work. It is this type of variations that better control systems can be employed to reduce variability. Assignable causes are specific reasons for variability (can be assigned) and hence removed.

Generally, methods of benefit analysis especially for control systems revolve around the concept of estimating the benefits prior to implementation. This is mostly done in order to justify financial investments required. Several methods have been proposed from industry and academia and will be highlighted in this section. It is to be noted that although process variability is not the only method used to link control benefits to monetary value, it is the most commonly used (Bauer and Craig, 2008). A very important factor however, that affects results of such estimation methods lies in the ability to predict the future performance of the control system (Marlin, Perkins, Barton and Brisk, 1991). Certain methods have been adopted that relies on experience, rule of thumbs, best operator technique and plant test simulating control upgrades can be adopted. However, each of these have their advantages and disadvantages.

Marlin et al. (1991) provides a structured approach to economically evaluate benefits arising from improved control strategies in which quantitative methods of assessing benefits from improved controls were proposed and applied to seven different cases.

Key aspects of their work concluded that economic benefits generally arise from operation of key process variables closer to the constraint limits. They have proposed a method for calculating the benefit from control system by comparing averages and variances before and after system upgrades and have given the following general equation:

$$\text{Benefits} = \frac{\text{Improvement} \times \text{Incremental Value} \times \text{Unit throughput} \times \text{Time}}{\text{Service Factor}} \quad (2.2)$$

Where improvement is the difference between the base case and the improved operation, incremental value is the incremental economic value due to the improvement while the service factor is the fraction of time the unit will be under the application of the improved control strategy.

Craig and Henning (2000) conducted a study with the objective of producing statistically significant data (from the existing setup) for analysis. Their work provides for a method to investigate how improvements resulting from advanced control can be measured to a required level of significance.

Shunta (1995) proposed a statistical approach to identify the extent of improvement gained in terms of reducing variability and proposed an index called Process Performance Index.

It is crucial to note that most of the methods cited in literature in measuring the extent of the improvement and linking it to monetary gains derive their

basis from the ability of the advanced control system to achieve reduced variability and drive the process closer to the constraint boundary.

2.5.1. General categorization of benefit analysis methods

A broad categorization of the techniques available for benefit analysis will be given below:

2.5.1.1. Best Operator technique

This method depends on previous historical operating data of key process parameters. It relates to the idea that an improvement in process control can be estimated by looking at the set of best previously attained performance. Hence, the difference between the average and the best operator data represents the potential improvement to be gained from advanced process control. However, this depends on the assumption that the control strategy will perform as good as

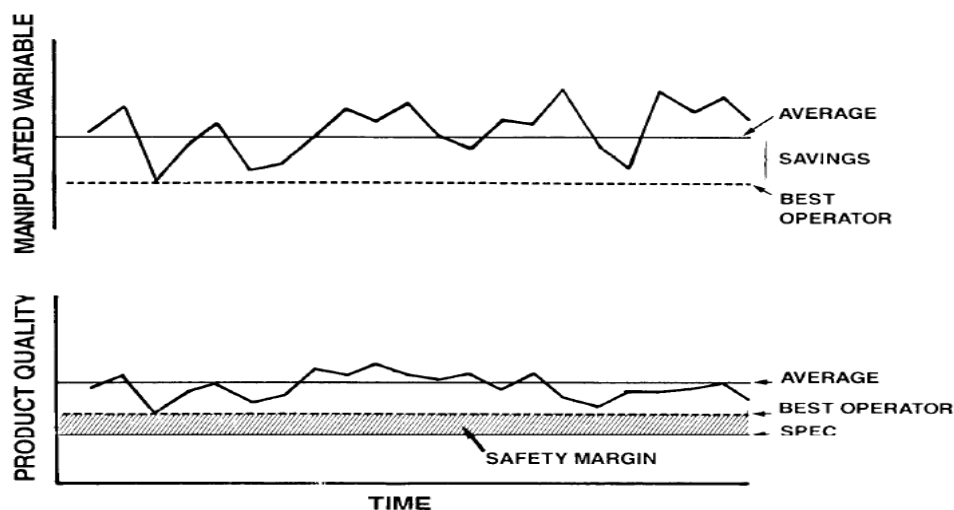


Figure 7 Illustration of the best operator technique (SETPOINT Inc., 1986)

the best operator and eliminates any possibility that the

control might result in a better performance than the best operator scenario. Accordingly, results from such analysis can be considered conservative.

The new average operating point can be calculated from the current average and the best operator scenario.

2.5.1.2. Experience based

Experience based methods or rule of thumb techniques are used in industry to have an order of magnitude estimates on the potential benefits. Possible choices of techniques that fall under this category include usage of previously attained reduction in process variation which can range from 50% to more than 90% (SETPOINT Inc., 1986). Since every plant is unique, such estimates are unlikely to produce accurate results.

There are various techniques under this category. All of the techniques assume that improvement arising from improved dynamic process control materialize in the form of reduced variance in the key process parameters and that the underlying data from the process is normally distributed. Variance reduction from advanced process control projects range from 50 to 90% (SETPOINT Inc., 1986). Once the estimated variance reduction has been made, an estimate of how much process improvement can be made.

Several techniques are classified under this technique.

1. Same Percent Limit Violation Technique

The basic concept behind this technique is the assumption that the advanced process control system will be allowed to violate the limit the same percent of the times as the current operator does however with an improved estimated standard deviation.

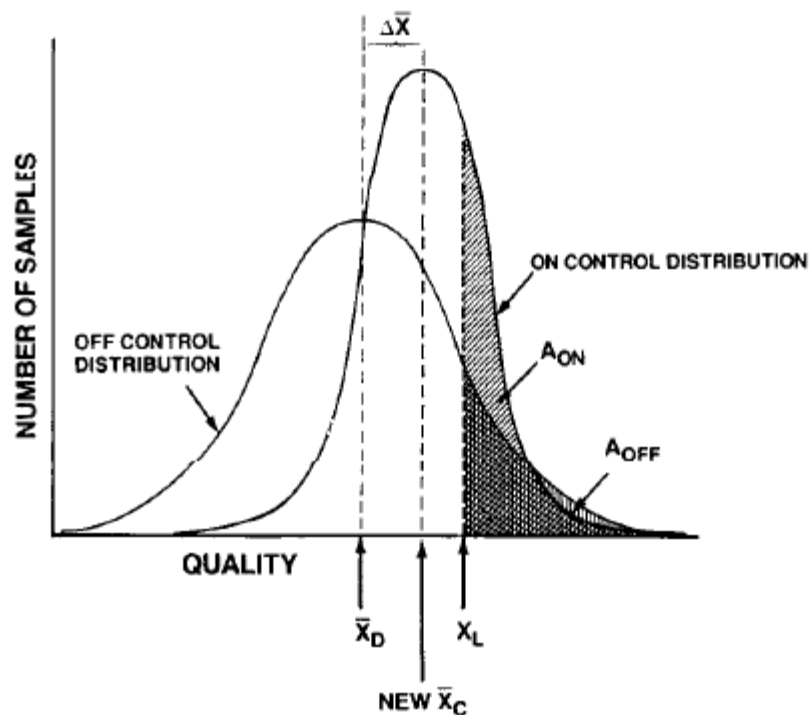


Figure 8 Illustration – Same Percent Limit Violation technique
(SETPOINT Inc., 1986)

2. Same Percent Violations of the 5% Limit

In cases where a specified limit cannot be obtained from plant data or when an actual limit is not clear, a more realistic method is to choose a value beyond which only 5% of the data falls. Figure 9 illustrates this concept.

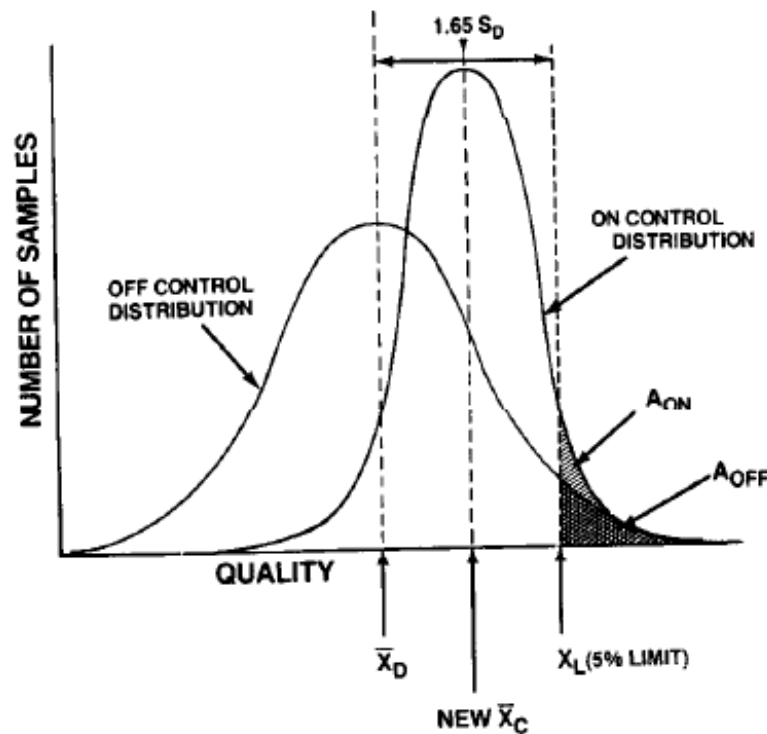


Figure 9 Illustration Same percent violation of the 5 % Limit
(SETPOINT Inc., 1986)

3. Limit Violation A% of Time

This technique is used to estimate performance improvement when the limit is rarely violated under the existing control system. This concept assumes that the

limit can be violated a certain percent A% of the time. This method will set the new average performance a certain number of standard deviations away from the limit. The selection of the violation percent A% determines how much conservative the system will be.

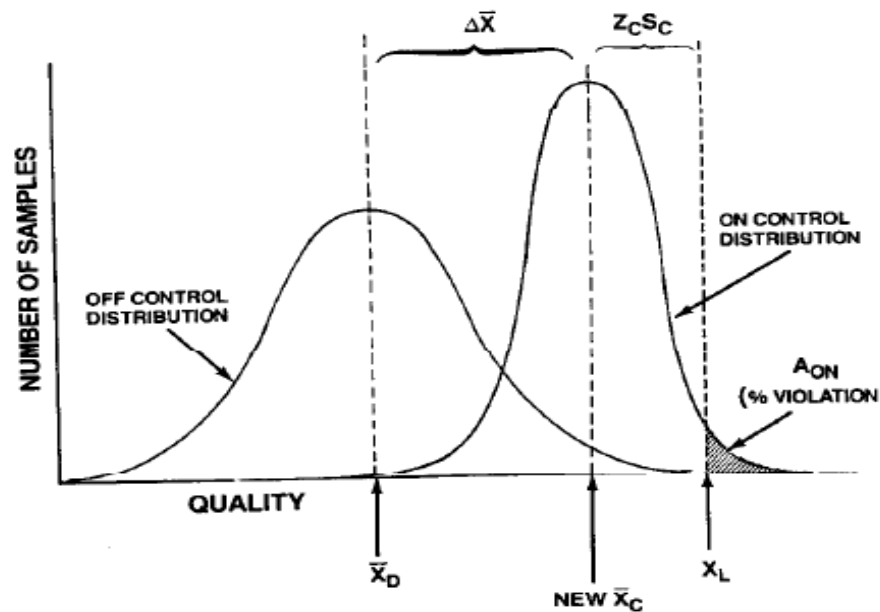


**Figure 10 Illustration Limit violation A% of time
(SETPOINT Inc., 1986)**

2.5.1.3. Statistical methods

This approach analyzes the means and standard deviation of sample data of the process parameter. It is assumed in this case that the process control will result in certain new operating point. The new operating point is chosen closer to the product specification (based on experience) and then a hypothesis testing is conducted to evaluate whether the new operating

capabilities is enough to justify the financial investment required.

- Craig and Henning (2000) presented a structured approach for analyzing benefit from improved control systems. Their work provided for a framework to investigate how improvements from advanced control projects can be measured to a required level of significance after the improved control system has been installed.

2.6. Base case identification

It is of paramount importance in giving due attention on establishing the requirements for identification of the base case. The base case provides the basis for comparing the extent of improvement of the improved control system. A wrong identification of the base case might lead to the conclusion of either overestimating or underestimating the benefits sustained which can be detrimental to future projects. Hence, it is quite important to identify key operating variables. SETPOINT Inc. (1986) recommends the analysis of the following –

- Critical product qualities
- Constraint variables
- Important manipulated variables

Determining the base case operation requires in depth knowledge and experience of the process under study including analysis of plant data, laboratory data sheets, operation targets as well as operator log sheets (Marlin, Perkins, Barton and Brisk, 1991).

It is also important to not consider data that does not represent other than the normal operation routine. Unusual operations such as equipment shutdowns or trips should not be included. Operational conditions and feed stock quality shall be typical of that prevailing after the upgrade of the control system. Equally important is that the improvements over the base case should preserve material and energy balance and avoid invalid savings. However, due to the cost of collection and analyzing of the required data, minimum amount of absolute necessary information should be identified (SETPOINT Inc., 1986)

2.7. Commercially available MPC technologies

A review of the commercially available MPC products shall be presented in this section. Most of the products here have been explicitly developed to be used for industrial purpose applications. A primary focus of this review shall be on the theoretical aspects of the products – the objective function, identification of constraints and their treatment and the tuning parameters. However, even though an emphasis on the theoretical elements of these products shall be done, it is to be noted that their due appreciation comes from the fact that these systems interact through interfaces with real time distributed control system of the plants, from the type of modelling facilities provided and

the support for commissioning and maintenance post set up. These products are also strongly influenced by the culture existing in industries especially amongst process control practitioners. This is a culture in which considerable authority is given to operators in running the plants mainly due to the entrustment bestowed on operators ever since the beginning of the industrial revolution. Hence, many of these products go to impressive lengths in trying to support the role of traditional operators rather than replacing them.

As mentioned earlier, MPC is usually implemented above the traditional control layer. In reality the MPC optimize and provide the set points to the lower control elements. The aim of these predictive controllers is to increase plant profitability. However, at times when the controllers appear to be malfunctioning or for some reason do not provide the most optimal outputs, the MPC can be taken offline and handed back to the operators.

2.7.1. DMC plus – AspenTech

DMC plus is product derivative from a company called AspenTech. The Dynamic Matrix Control DMC was first developed and patented by Shell Oil (Cutler and Ramaker, 1979). It was later further developed and marketed by a company called DMC Corporation that was subsequently acquired by AspenTech. The key features of the DMC included:

- A linear step response model of the plant.
- Optimal inputs as solution to a least square problem.
- A quadratic objective function defined over a finite prediction horizon.

- Future behavior of the plant output defined by trying to follow the set point as close as possible.

The linear step response model used by the DMC relates the changes in process output to weighted sum of the past changes in the input. The model can be given as (Qin and Badgwell, 2003):

$$y_{k+j} = \sum_{i=1}^{N-1} S_i \Delta u_{k+j-i} + S_N u_{k+j-N} \quad (2.3)$$

Where S_i are the step response coefficient. Mathematically the step response model can be obtained as the integral of the impulse response.

A modification of the of the DMC commonly known as the Quadratic Dynamic Matrix Control (QDMC) developed by Shell as well. There have been some references to it in literature however commercial rights have been retained by Shell (Maciejowski, 2000).

The DMCplus utilizes a multivariable finite step response model to represent the model of the plant. An extension permits the usage of the integrators to be modelled as well as asymptotically stable plants.

The DMCplus computes optimal solutions in two steps. The first step involves solving an LP problem to determine the set points or targets of the controlled variable and has the following elements:

- The LP problem is mainly a cost function involving factors affecting the economic performance of the plant. Operators are asked to input data on product quality, economic costs of

the input variables (generally the cost of the processed streams) etc.

- The cost function is a linear function with input and output constraints imposed and respected in this step. The constraints are expressed as linear in equalities.
- The target values are obtained by solving an LP problem. It must be noted that this step involves obtaining values from the steady state gain of the model.
- A steady state equal concern error is defined by the operator which is how much a deviation of a certain controlled variable will cause significant impact. If the LP problem results in an infeasible solution, the controller will use to steady state equal concern error to define priorities for the constraints and will allow the constraint with the least priority to be violated until a feasible solution is reestablished once again. Alternatively, a minimum distance approach is taken which will seek to minimize the sum of the weighted distances of targets from their constraint boundary with the weights reflecting the steady state equal concern error.

The second step involves solving to obtain the future trajectory of the input variables over the control horizon to achieve the values obtained from the LP problem in step one and has the following traits:

- The problem is formulated as a solution a least square problem which includes a quadratic function.

- A quadratic penalization of input variables is imposed (move suppression) and as such constraints are not explicitly addressed rather penalty weights are increased as the predicted variables approach the boundary constraints. Hence, constraints are treated quintessentially as soft constraint.
- A dynamic state equal concern error is defined by the operator in a manner similar to step one. Here, the equal concern error is used to implement soft constraints on the variables. Hence, if an output reaches close to the constraint boundary, the associated weights in the target of the least square problem increased to reflect this.

2.7.2. RMPCT – Honeywell

Another product providing MPC in the market is the Robust Multivariable Predictive Control Technology – by a company called

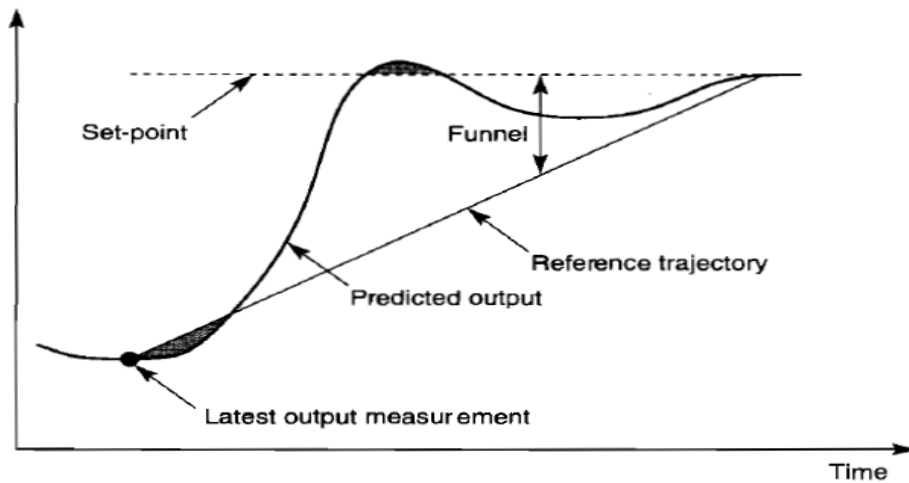


Figure 11 Concept of zone and funnel RMPCT - (Maciejowski, 2000)

Honeywell, which is an American company with its headquarters located in North Carolina.

The RMPCT uses a distinguishing concept of a zone or a *funnel* where the objective of the control is to maintain the output parameter within defined region rather trying to follow a specific target. The main points of the system include:

- The objective of the control is to maintain the output parameters within an allowed region defined by an upper and lower limit. As long as the output remains in this region, no penalty is given.
- If the output variable lies outside the allowed region, then a straight-line trajectory is constructed between the current value and the nearest edge of the allowed zone with the output variable is supposed to follow this trajectory to the allowed zone. In course if the predicted value happens to lie outside the *funnel* region formed by the target region and the straight-line trajectory, then an error is defined which is the distance of the predicted value to the nearest edge of funnel. A quadratic cost function is formed which is the weighted sum of these errors. The optimization problem hence formed involves a quadratic problem with the objective of choosing future inputs that would minimize this cost function subject to constraints on the input and output.
- It is evident that the *aggressiveness* with which an output is returned to its allowed range depends on the slope of the edge of the funnel formed by the allowed zone and the straight line trajectory.

- Similar to the DMCplus, the RMPCT is also arranged on two level – optimization and control. A steady state model is utilized in order to find the optimal steady state solution for the process with the assumption that the current steady state is the predicted steady state of the model. Here, a conventional linear economic function is used which captures economic benefits such as production quantities of certain products, cost of using control inputs together with a quadratic function that captures the economic cost of moving away from the current predicted steady state values. Here process variable can appear in this function but may not be allowed to appear in the dynamic optimization for the control inputs and vice versa.
- The handling of the new targets from the above steady state optimization to the dynamic optimization does not occur in a conventional way per se. Rather the new steady state values computed are incorporated in the dynamic optimization with a weighting scale that determines the aggressiveness with which the process is allowed to push to the new steady state values. As mentioned, the objective of the RMPCT is to maintain outputs in an allowed zone and as such placing a low weighing on the steady state would encourage the controllers to push for correcting the predicted excursions outside the allowed range but also to push the process for desirable steady state when no such excursions exist.
- The net outcome of the RMPCT combination of steady state and dynamic optimization is soft constraints on the controlled

variables with inexact penalty function hence in some circumstances constraint boundary violation is allowed to happen even if there is a feasible solution which is avoided.

- The RPMCT uses the Autoregressive with Extra Input ARX as a model to form predictions of the output.
- The RPMCT employs several strategies to increase its robustness in modelling for errors. The first strategy is to introduce scaling of the input and output variables to optimize the conditioning of the optimization problem. The idea here is to decrease the sensitivity of the output responses to the modelling errors. The second strategy involves adjusting the internal model by assessing the model using simulation to evaluate a certain acceptability criterion for a range of plant model errors.

2.7.3. Connoisseur – Simulation Sciences

SimSci is one of the leading process simulation providers. It was first established in 1976 and has provided the industry's first ever simulation software capable of running on any main frame and PC – SimSci PRO/II in 1987. In 2014, SimSci was acquired by AVEVA which is a subsidiary of Schneider Electric.

The Connoisseur has 3 models of predictive control which will be briefly explained below:

A. Long Range LR mode

- In this mode a quadratic cost function is optimized, ignoring the constraints at first.

- The model used is linear and hence can be considered as a linear quadratic LQ problem.
- The approach followed does not require open loop computations to be made but rather closed loop predictions with fixed gain feedback computed in order to evaluate whether there are any constraints being violated. If there are violations predicted then weights in the cost function are modified until no constraint violations are predicted.

B. QP – LR mode

- In this mode, a quadratic cost function is optimized considering hard constraints on the input while output constraints are ignored.
- Similar to the LR mode, weights are introduced in the cost function to remove any predicted violations on the predicted output.

C. QP mode

- In this mode a quadratic cost function is optimized similar to the QP-LR mode.
- However, all constraints are treated as soft constraints and included in the optimization function with weights used as penalties in order to define the constraint violation priorities.

The quadratic function can be formulated to include penalizing the control move and deviations of the control input from the target value

if desired. Optimization of set points is done by solving an LP problem at a greater interval than the control interval.

The main driver behind having these 3 modes of operation is to reduce the computational requirement when not needed. For example, in the QP-LR mode, the number of constraints is smaller than that in the QP model and hence easier computations will be required to obtain a solution. The prediction model used in this product is the ARX model, similar to that used by the RPMCT product from HoneyWell.

2.7.4. PFC and HIECON – Adersa

PFC is a MPC methodology and has the same name of a product offered by a French company called Adersa. The PFC differs from the other products in terms of original purpose application as it was not targeted for the petrochemical industry.

Predictive Functional Control PFC was originally pioneered by Richalet (Maciejowski, 2000) and has the following traits:

- This methodology of predictive control differs from the other in certain features – the use of relatively lesser number of coincidence points (matching of point between the ideal trajectory and the required parameter) and the use of reference trajectory distinct from the set point trajectory.
- However, the most significant difference in feature is that the future input is assumed to be a linear mixture of a few simple functions. Hence the notion of control horizon does not exist

here unlike the other products but rather replaced by a number of basis functions.

- The cost function takes the form of a quadratic problem QP but does not penalize the input movements mainly due to the reason that the required performance depends significantly on the reference trajectory and not on the weights imposed on the input variables. The other reason being that due to the polynomial structure of the basis function forming the future prediction of the input variables results in relatively fewer degree of freedom available for the optimizer and this generates smoother input trajectories.

Another variation of this product is called the HIECON which is based on the same methodology but aimed for multivariable applications with wider range of solution algorithms. Successful applications of both these products have been recorded in several different industrial fields such as batch reactors, milk dryers to fuel injection in car engine systems and altitude correction of launchers and missiles.

2.7.5. 3dMPC – ABB

The 3dMPC is the predictive control product offered by the company ABB with its head quarters in Zurich, Switzerland. The company is a multinational corporation operating mainly in the automation, electrical equipment, robotics and heavy machineries sectors. The main elements of this product are listed below:

- 3dMPC utilizes a discrete time space state model as its internal model for prediction. The parameters of the model are determined through multivariable identification methods.
- The product utilizes 3 degrees of freedom (and hence is the motivation behind the product name) that can be used to configure to have different dynamic responses to set point changes, disturbance changes (measurable and unmeasurable) and to model mismatch.
- A quadratic cost function is utilized for the optimization problem.
- Constraints for process variables, deviations and manipulated variables are ranked according to their priorities. Under certain operating conditions, the constraints with the least priority are allowed to be violated until feasible solutions are obtained again. However, there is a possibility for defining hard constraints on the manipulated variables which cannot be violated.
- Nonlinear model applications are also available as with the rest of the products.

2.8. Basis of Prediction – MPC Models

One of the main components of MPC is prediction. The concept of prediction in MPC quintessentially lies in predicting future output response. Prediction of future outputs $y(k + p|k)$ $p = 1, 2, 3, \dots, N$ is required at time k in order to solve of the cost function in equation (2.1).

Models allow for the automation of prediction and hence constitute a major component of any MPC form. Models predict the shift in the dependent variables for a shift in the independent variables. In process industries, the dependent variables are the parameters that represent process-desired parameter. These models require the usage of current plant measurements, the MPC model predictions, the current dynamic state of the system, process targets and constraints in order to calculate the most optimal trajectory for the independent or decision variables.

While most of real-life processes (and classical control law as well) operate in a continuous time domain, time is needed for various calculation and decision making which cannot be instantaneous especially considering interacting input/output constraints and performance. Hence generally, MPC models operate in a time discrete manner.

Modelling in MPC can take both linear and nonlinear models. However, a nonlinear model would imply that the subsequent optimization technique is of a tedious nature and may not even converge. Linear models can be taken as a fair representation of the system by assuming model uncertainty and allowing gain scheduling of control laws (Rossiter, 2005).

Real-life processes are mostly nonlinear. However, they can often be considered as an approximation to linear systems over a small range of

operation. Linear MPC approaches are commonly used in industrial applications with the provision of feedback mechanism to compensate for prediction errors due to inherent structural mismatch between plant and model (MEHTA and REDDY, 2017). It is generally preferred to implement MPC using linear models if they yield accurate enough predictions since their implementation is relatively easy.

In linear MPC models, the principle of superposition enables effects of changes in multiple independent variables to be combined together to predict the future response of the dependent variables. This simplifies the optimization problem to a series of direct matrix algebra calculations thereby resulting in a fast and robust system (García, Prett and Morari, 1989).

Commonly used linear models include the Space-State, Controlled Auto-Aggressive Integrated Moving Average CARIMA and Finite Impulse and step response models.

2.8.1. Basic format for prediction modelling

The general prediction format of future behavior of systems can be represented by equation (2.4) as presented by Rossiter, 2005.

$$\mathbf{y}_{\rightarrow k} = \mathbf{H} \Delta \mathbf{u}_{\rightarrow k-1} + \mathbf{P} \mathbf{x}_{\leftarrow k-1} \quad (2.4)$$

Where the following represent:

- $\mathbf{y}_{\rightarrow k}$ the vector of future predicted output responses.
- $\mathbf{u}_{\rightarrow k-1}$ the vector of current and future control inputs.
- $\mathbf{x}_{\leftarrow k-1}$ the vector of the past process states.
- $\mathbf{H}$ is the Toeplitz matrix of the system step response.
- $\mathbf{P}$ is a matrix where the coefficients depend on model parameters.

2.8.2. Prediction with State Space Model

A state space model is a mathematical representation of a physical process using a set of variables – input, state variables, and output – related in a first order differential equation. State variables are variables whose value change with time and can describe enough about the system to determine future behavior in the absence of external forces.

The term state space in control engineering first appeared in the 1960s (Kalman, 1960). The space state model provides for a framework for

analyzing deterministic and stochastic dynamic system which are measured through a stochastic process.

A common state space model with one step ahead prediction is given in equation (2.5).

$$x_{k+1} = Ax_k + Bu_k; \quad y_{k+1} = Cx_{k+1} + d_k \quad (2.5)$$

Where

- u represent the input variable.
- y represents the predicted output variable.
- x represents the state variable.
- d represents the unmeasured disturbance, assumed to be a slow varying disturbance

At $t = k + 2$,

$$x_{k+2} = Ax_{k+1} + Bu_{k+1}; \quad y_{k+2} = Cx_{k+2} + d_k \quad (2.6)$$

Substituting (2.6) in (2.7)

$$x_{k+2} = A^2x_k + ABu_k + Bu_{k+1}; \quad y_{k+2} = Cx_{k+2} + d_k \quad (2.7)$$

Similarly, for at $t = k + 3$ and substituting by equation (2.13)

$$x_{k+3} = A^3x_k + A^2Bu_k + ABu_{k+1} + Bu_{k+2}; \quad y_{k+3} = Cx_{k+3} + d_k \quad (2.8)$$

Hence a general recursion expression for n-step ahead predictions is shown below

$$\begin{aligned}
x_{k+n} &= A^n x_k + A^{n-1} B u_k + A^{n-2} B u_{k+1} + \dots + B u_{k+n-1} \\
y_{k+n} &= C[A^n x_k + A^{n-1} B u_k + A^{n-2} B u_{k+1} + \dots + B u_{k+n-1}] + d_k
\end{aligned}
\tag{2.9}$$

A vector of future predictions up to a horizon n can be formed as follows:

$$\begin{aligned}
\underbrace{\begin{bmatrix} x_{k+1} \\ x_{k+2} \\ x_{k+3} \\ \vdots \\ x_{k+n} \end{bmatrix}}_{\mathbf{x}_{\rightarrow k}} &= \underbrace{\begin{bmatrix} A \\ A^2 \\ A^3 \\ \vdots \\ A^n \end{bmatrix}}_{\mathbf{P}_{xx}} x_k + \underbrace{\begin{bmatrix} B & 0 & 0 & \dots \\ AB & B & 0 & \dots \\ A^2 B & AB & B & \dots \\ \vdots & \vdots & \vdots & \vdots \\ A^{n-1} B & A^{n-2} B & A^{n-3} B & \dots \end{bmatrix}}_{\mathbf{H}_x} \underbrace{\begin{bmatrix} u_k \\ u_{k+1} \\ u_{k+2} \\ \vdots \\ u_{k+n-1} \end{bmatrix}}_{\mathbf{u}_{\rightarrow k-1}}
\end{aligned}
\tag{2.10}$$

And

$$\begin{aligned}
\underbrace{\begin{bmatrix} y_{k+1} \\ y_{k+2} \\ y_{k+3} \\ \vdots \\ y_{k+n} \end{bmatrix}}_{\mathbf{y}_{\rightarrow k}} &= \underbrace{\begin{bmatrix} CA \\ CA^2 \\ CA^3 \\ \vdots \\ CA^n \end{bmatrix}}_{\mathbf{P}} x_k + \underbrace{\begin{bmatrix} CB & 0 & 0 & \dots \\ CAB & CB & 0 & \dots \\ AC^2 B & CAB & CB & \dots \\ \vdots & \vdots & \vdots & \vdots \\ CA^{n-1} B & CA^{n-2} B & CA^{n-3} B & \dots \end{bmatrix}}_{\mathbf{H}_x} \underbrace{\begin{bmatrix} u_k \\ u_{k+1} \\ u_{k+2} \\ \vdots \\ u_{k+n-1} \end{bmatrix}}_{\mathbf{u}_{\rightarrow k-1}} + \underbrace{\begin{bmatrix} d_k \\ d_k \\ d_k \\ \vdots \\ d_k \end{bmatrix}}_{\mathbf{L} \mathbf{d}_k}
\end{aligned}
\tag{2.11}$$

These general compact form of equation (2.10) and (2.11) can be given as follows -

$$\mathbf{x}_{\rightarrow k} = \mathbf{P}_{xx} \mathbf{x}_k + \mathbf{H}_x \mathbf{u}_{\rightarrow k-1}$$

$$\mathbf{y}_{\rightarrow k} = \mathbf{P} \mathbf{x}_k + \mathbf{L} \mathbf{d}_k + \mathbf{H} \mathbf{u}_{\rightarrow k-1}$$

(2.12)

Equation (2.12) represents the compact vector form with separate terms $\mathbf{P}_{xx} \mathbf{x}_k$ and $\mathbf{P} \mathbf{x}_k + \mathbf{L} \mathbf{d}_k +$ representing variables that are known while $\mathbf{u}_{\rightarrow k-1}$ represents the decision variables which are the future inputs.

2.8.3. Prediction with step response models

The concept behind using a step response model is that one can apply a step change into an input or manipulated variable and record the open loop response of each output variable until each variable has settled into a constant value. In essence it is possible to build a model of the plant by using utilizing data obtained through plant step test response models, hence is much more convenient, and consumes much lesser time than developing tedious state space models.

However, this concept does have drawbacks:

- It can only be used with plants considered asymptotically stable.
- It can be impractical to conduct step steps to all desired manipulated variables since this would cause disruption to normal operations.

- Step responses will only be of significance and benefits if all the controlled variables are measured. Otherwise, alternative methods will be required to estimate the outputs.
- Storing of multivariable step response models can be inefficient form of model representational especially in terms of storage requirements.

Given that a model of the plant behavior step response sequence matrix is available $\{S(0), S(1), \dots, S(N)\}$ or the impulse response sequence matrix $\{H(0), H(1), \dots, H(N)\}$ and no disturbance assumed, predictions of future outputs can be given by:

$$y(k+j|k) = \sum_{i=1}^{\infty} H(i)\tilde{u}(k+j-i) \quad (2.13)$$

With the input sequence $\tilde{u}$ consisting of known past inputs and predicted future input.

$$\tilde{u} = \begin{cases} u(i) & \text{if } i < k \\ u(i|k) & \text{if } i > k \end{cases} \quad (2.14)$$

With the following assumptions:

1. $H(i) \approx 0$ for $i > n$
2. $S(j) = \sum_{i=1}^j H(i)$ and $H(0) = 0$
3. $\sum_{i=1}^j H(i)u(j-i) = \sum_{i=1}^j S(i)\Delta u(j-i)$

It can be shown that the step response prediction model is obtained by:

$$\sum_{i=j+1}^N S(i)\Delta u(k+j-i) + S(j)u(k-1) + \sum_{i=1}^j S(i)\Delta u(k+j-i|k) \quad (2.15)$$

Chapter 3 – Process description

- 3.1 Condensate inlet receiving unit
- 3.2 Gas dehydration unit
- 3.3 Absorption unit
- 3.4 Deethanizer unit
- 3.5 Rich oil still unit
- 3.6 Fractionation unit

This chapter is intended to present a process description of the plant under study. The plant is located in the southern governorate of the Kingdom of Bahrain. The plant was first commissioned in 1979 with a design capacity of 110 MMSCFD. Subsequent debottlenecking projects undertaken with the most recent being in 2012 eventually raised the capacity of the plant to 170 MMSCFD. The primary source of feed gas to the plant is received from upstream associated gas gathering network which supplies all 3 plants of the facility. The network consists of 31 well manifolds together with compression and dehydration stations located in different areas. Another source of gas which is considered quite rich in NGL products is the Refinery Off Gas (ROG) coming from a nearby refinery. Considered a waste initially, the ROG was subjected to disposal by the refinery where the gas was routinely flared. However, with more favorable economics hence justifying the investments required to extract NGL products from the gas, the ROG was later introduced as a partial feed to the subject facility. Due to the declining quality of the associated gas – the NGL rich ROG feed stock is considered to be vital and has a crucial impact on the performance of the subject plant.

The plant can be divided into the following:

- A. Condensate Stabilization Unit
- B. Dehydration Unit
- C. Absorption Unit
- D. Deethanizer Unit
- E. Rich Oil Still Unit
- F. Fractionation Unit

A description on each of these unit outlining the intent, operational philosophy and main equipment shall be done in next sections.

3.1. Condensate Inlet Receiving Unit

This unit is the section that first receives the gas condensate feed from the gas in preparation for further processing. Condensate is obtained due to the compression of the feed gas in the compressor stations. Heavier products are condensed out and sent in a separate line to the plant specifically to the condensate inlet-receiving unit where initial processing is necessary in order to process further.

The condensate stabilizer column is a trayed column in which ethane and lighter components are stripped off. Heat is supplied to the stabilizer column to boil off these components. The condensate bottom liquid from the reboiler is later directed to the Depropanizer column in the fractionation unit for further processing

The overhead vapor draw from the stabilizer unit is combined with the inlet gas in order to recover the traces of propane and the products.

The unit is considered small in terms of capacity and hence provides no incentive to include this unit in the study.

3.2. Gas Dehydration Unit

The gas naturally will contain some amount of water. This water will need to be removed since it can lead to hydrate formation in pipelines and different equipment when the gas reaches low temperature especially in the absorption system. The dehydration unit is a typical Triethylen Glycol (TEG) system.

The glycol regeneration unit is considered a small subunit responsible for regenerating the spent glycol. From the operator's experience this unit is considered minor with no significant problem in terms of operation and hence no significant incentive to include the unit in the study.

3.3. Absorption unit

This is the unit responsible for recovering the main products of the plant, namely, Propane, Butane and Naphtha from the feed gas. The performance of this unit directly affects the main KPI in the operating facility, which is the propane recovery percent. The target design value for the propane recovery is 79%. However, this value depends significantly on the performance of the operation personnel, ambient weather conditions and changes in the feed gas composition. Changes in feed gas composition primarily when the ROG from the nearby refinery is reduced or completely stopped affects the production negatively (due to the leaner feed) especially in the Deethanizer section and on-specification propane production is not achievable without operator intervention on the process conditions.

A specific type of absorption oil with molecular weight 126 kg/kmole (Aviation fuel JP4 or equivalent) which exhibits preferential absorption (through absorption and condensation) towards heavier hydrocarbons Propane, Butane and Naphtha is utilized. The absorption unit mainly consists of an Absorber tower and a pre-saturator accumulator.

The feed gas is first cooled down to a temperature of -32 °F in a series of two steps before entering the Absorber column from the bottom. The feed gas is counter currently contacted with a partially pre-saturated

lean oil in the column. Here, most of the products are absorbed by the oil and leaves the absorber from the bottom to enter the next section of the plant which is the Deethanizer. The gas now mainly consisting of methane, ethane and traces of propane leaves the absorber from the top where it is allowed to mix with fresh lean oil at 15 °F in order to absorb the traces of propane. This mix is then cooled down to -28 °F before entering the pre saturator accumulator where the mix is allowed to accumulate. The gas disengages from the liquid mix and leaves the pre saturator accumulator. This is the source of the residue gas from a certain flow is taken to provide energy to the plant while the rest is sent to the nearby refinery which eventually gets distributed between the refinery, a power station and an aluminum smelting facility. This liquid mixture is the partially pre-saturated oil, which is sent to absorber tower to recover the bulk of the products. The feed gas prior to entering the absorber column, specifically before the second stage of cooling is mixed with pre-absorbed oil taken from the chimney tray of the absorber column. It is estimated that around 70% of the absorption takes place in this stage. This mixture of preabsorbed oil and fresh feed gas enters the column below the chimney tray where the gas goes through the chimney tray while the liquid now called rich oil (rich with products) leaves the column below the chimney tray.

It is important to note that this process is very much sensitive to the temperature profile of the process. Any increase in temperature will negatively impact the absorption process and results in lesser recovery of products. This also results in higher slippage of propane into the residue gas thereby raising the calorific value of the gas beyond the

allowed specification of 850 btu/scf. Maintaining calorific value is crucial for the end users as the gas turbines and fired heaters in the downstream (refinery, power station and aluminum smelting facility) rely on this specification of calorific value for the design of their equipment and for safe operation. A process flow diagram for the unit is given in figure 12.

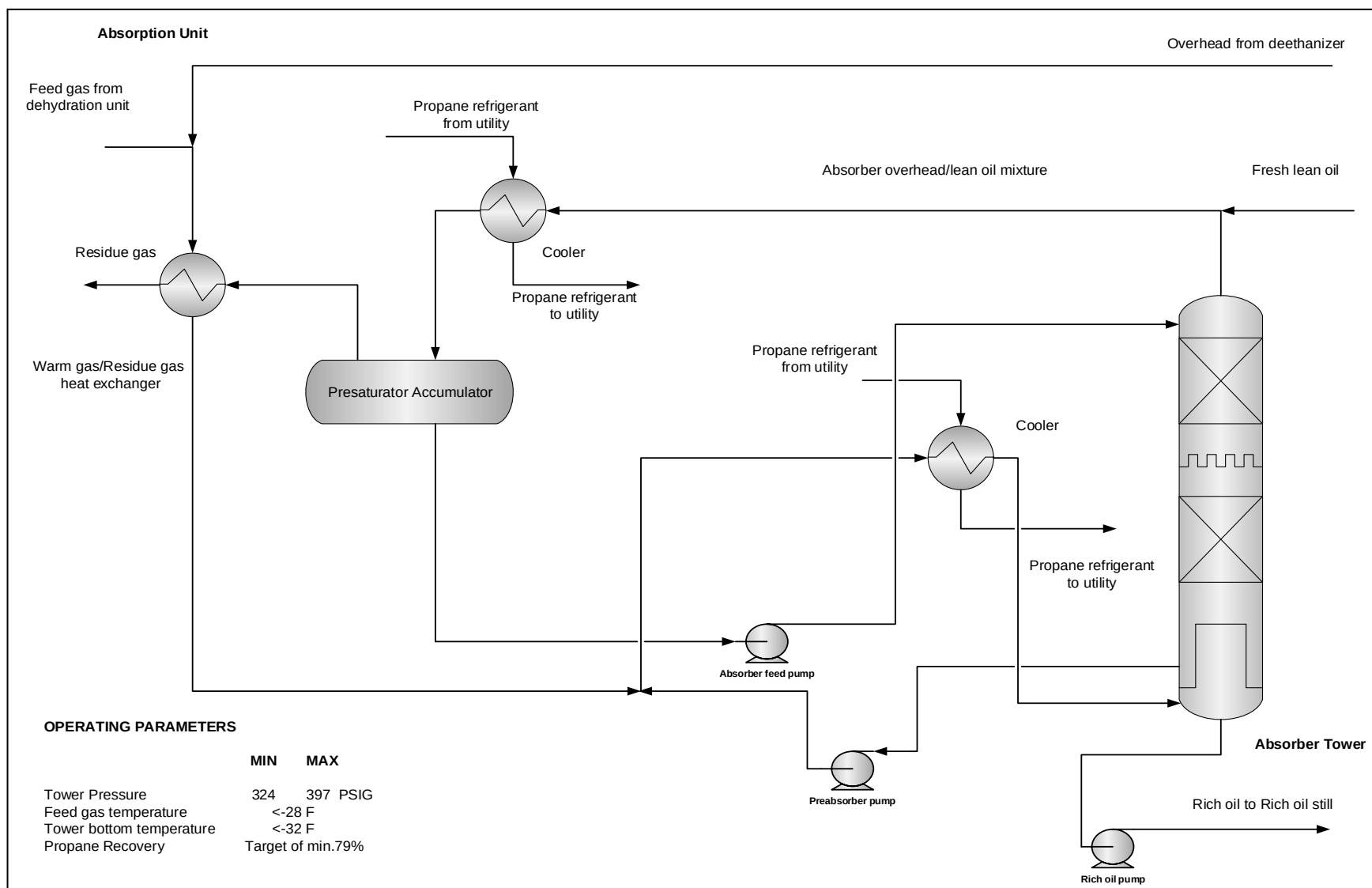


Figure 12 Absorption unit

3.4. Deethanizer Unit

After having absorbed the NGL products, there will be some traces of ethane, around 7%, slipping with the rich oil. The purpose of the Deethanizer unit is to remove the ethane hydrocarbon absorbed with the rich oil. Ethane is considered a contaminant in the final propane product with a maximum allowable content of 2 vol%. The performance of this unit is of crucial nature, as excess slippage of ethane will result in off specification of the propane product. The unit mainly consists of a one trayed and one packing column on standby. The trayed column is primarily placed in service while the packed column is commissioned during routine maintenance of the other column. The rich oil feed coming from the bottom of the Absorber column in the previous unit splits into two feed namely hot and cold feed. The hot feed enters a series of two heat exchangers in order to increase its temperature before entering the column. This is mainly done to ease and unload the bottom section of the Deethanizer column along with reboiler duty capacities. The Deethanizer unit is considered one of the most energy intensive units as stripping ethane from the product requires considerable amount of energy. From the experience of the plant operators, it has been observed that the performance of this unit is extremely sensitive to the composition of the feed gas. This is especially true when the ROG gas quality from the refinery changes or if the quantity reduces. The latter often happen when there is operational disturbance in the refinery or during turnarounds in some of the refineries crucial units supplying the ROG. A process flow diagram of the unit is given in figure 13.

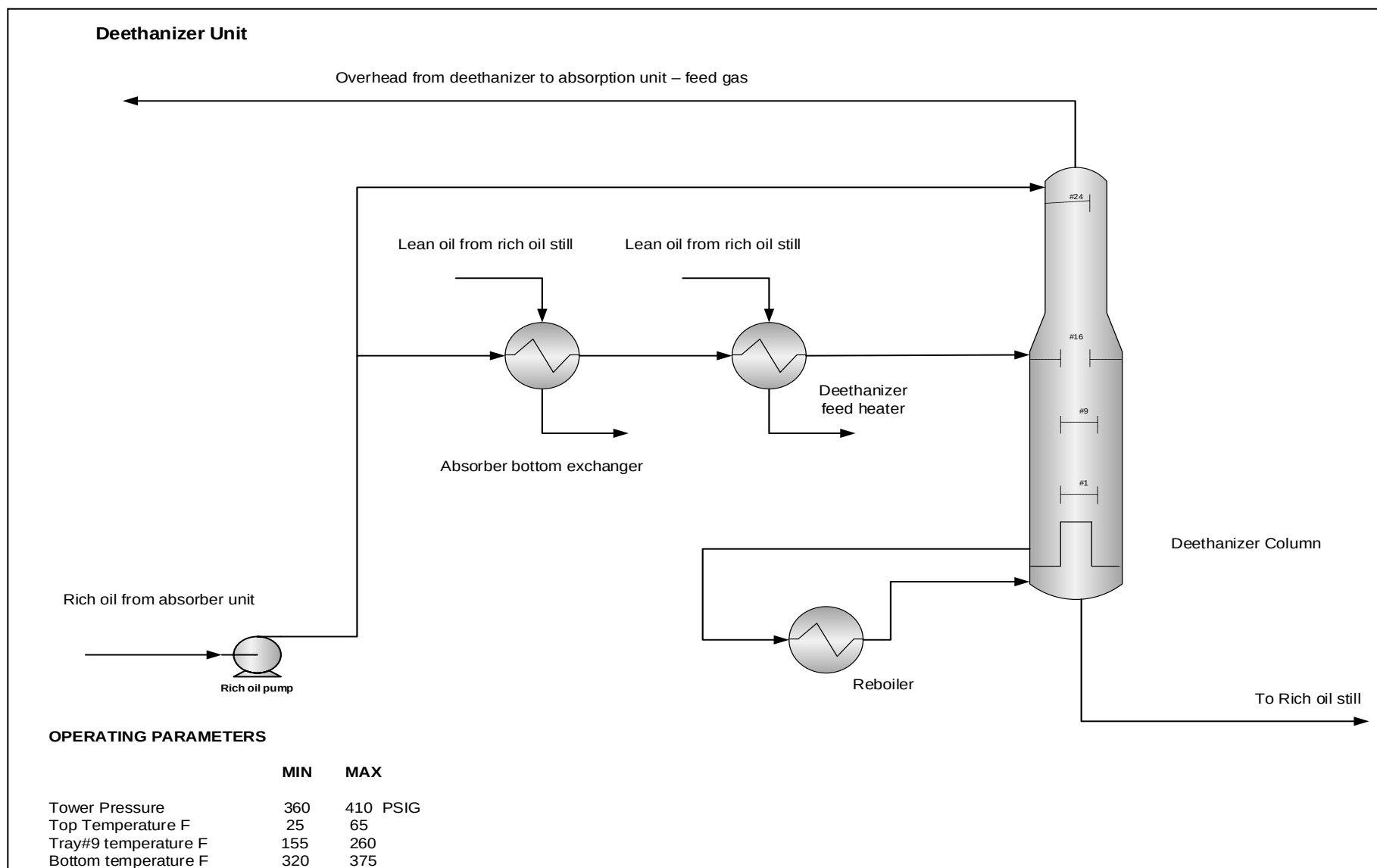


Figure 13 Deethanizer unit

3.5. Rich oil still unit

The rich oil after leaving the Deethanizer unit with only traces of ethane enters the next unit, which is the rich oil still. The purpose of this unit is to separate the absorbed products: propane, butane and naphtha from the Deethanizer rich oil (the absorption oil). The rich-oil still unit mainly consist of the rich oil still tower.

The deethanized oil at around 320 to 375 °F and 360 to 410 psig is first flashed to about 185 to 210 psig in the middle section of the tower. This process ensures that propane and butane move upward in the rectifying section while naphtha is gradually separated by contact with vapors coming from the lower stages. The bottom temperature is maintained between 510 to 540 °F to ensure all naphtha components are vaporized from the absorption oil. The overhead stream now consists of mainly propane, butane and naphtha while stream leaving the tower splits into two streams: part of the stream goes to supply hot oil to the utility network while the other stream goes to the furnace where more heat is applied before returning to the tower as liquid boil up.

The operation of the rich oil still is considered stable with no significant scenario of process upsets. A process flow diagram of the unit is given in figure 14.

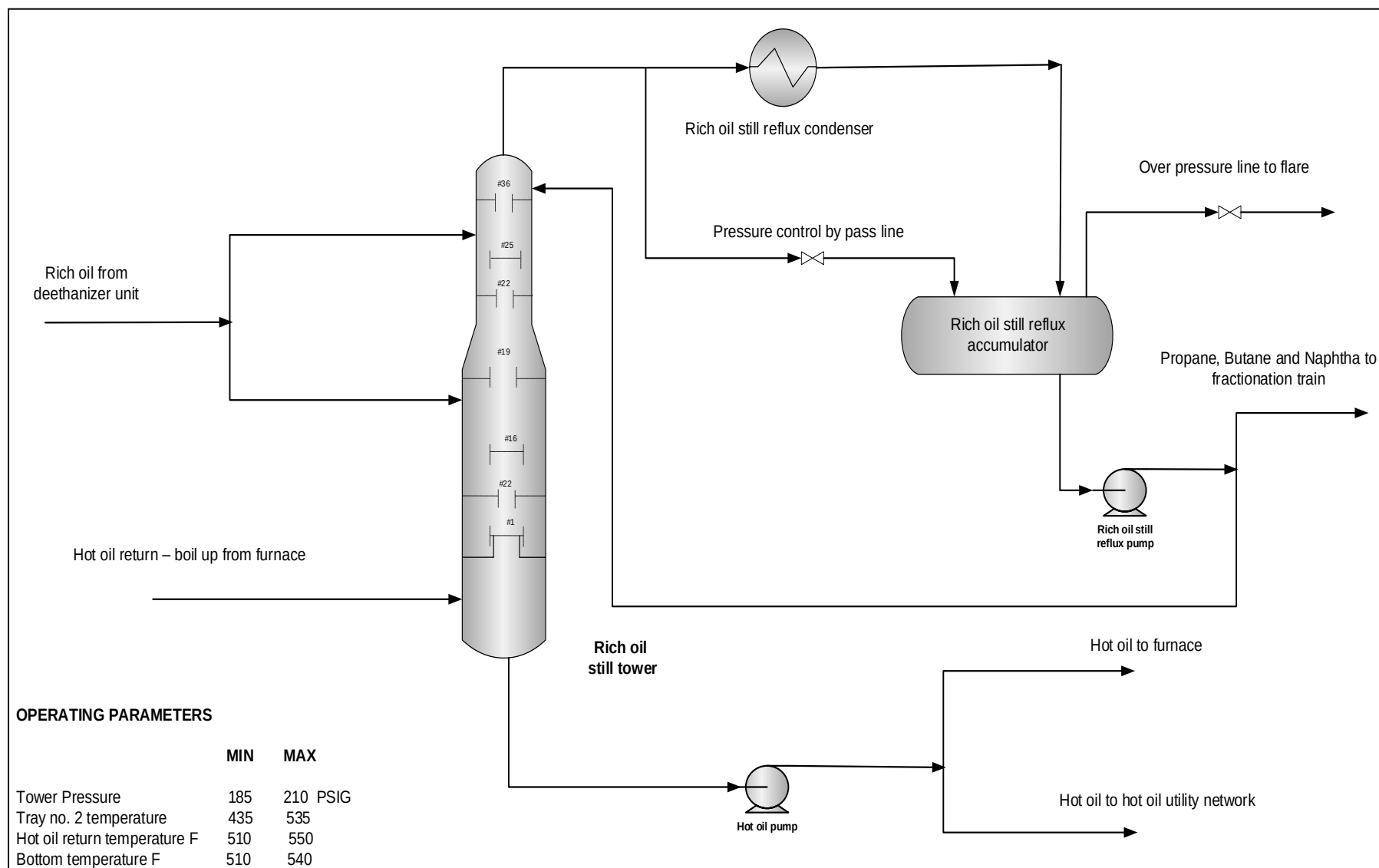


Figure 14 Rich oil still unit

3.6. Fractionation train

The liquid from the overhead of the Rich Oil Still enters the fractionation train. The purpose of the fractionation train is to separate the 3 different liquid products and to meet market specification in terms of composition and purity. The fractionation train consist of two columns namely the Depropanizer and the Debutanizer, in respective order. The Depropanizer is primarily responsible to separate the propane from the rest of the liquid products while the Debutanizer separates the butane product from the naphtha.

There are primarily two feeds for the fractionation unit, both of which first enters the fractionation unit through the Depropanizer column. The first is the liquid stream coming from the Rich Oil Still unit and it is called the cold stream while the second feed comes from the bottom of the condensate stabilizer unit and it is called the hot feed. The hot feed primarily consists of heavier hydrocarbons butane, naphtha, and minor amount of propane.

The cold feed at a temperature of around 140 to 170 °F enters the middle section of the Depropanizer column while the hot feed at round 380 to 415 °F enters the bottom section of the Depropanizer. The Depropanizer column is a tower consisting of 42 stages with a normal operating pressure of 235 to 300 psig. The overhead vapor from the top leaves the tower where it is condensed in an air cooler and sent into a reflux accumulator, which serves as a surge drum supplying the reflux and where the final liquid propane is drawn. The target purity of propane is

a minimum of 95 vol% with a maximum ethane and butane content of 2 vol% and 4 vol% respectively. The controlled variable that is mostly correlated and used to regulate the propane purity level is the temperature of tray no. 35 which is maintained between 130 and 145 °F by regulating reflux flow which is cascade controlled. The on-spec propane leaves the fractionation unit and enters the amine treatment unit to remove sulphur impurities such as mercaptans and hydrogen sulphide. The amine treatment unit is a minor unit where no significant impact in terms of the scope and objective of this thesis and hence will not be discussed in detail.

The bottom of Depropanizer column leaves the tower via the reboiler. This liquid stream containing mostly butane and naphtha now enters the second column in the unit, the Debutanizer. The purpose of the column is to separate the butane product from the heavier naphtha. The feed at 235 to 300 °F coming from the bottoms of the Depropanizer column enters the debutanizer column via the middle trays where it is flashed to around 80 psig. The flashing helps in stripping off the lighter butane product where its vapors pass through the rectifying section of the tower. The Debutanizer column has 42 trays with a normal operating pressure of 80 to 110 psig. The tower bottom temperature is maintained at around 225 to 255 °F with the help of the reboiler, which is also partially used to maintain tray no. 37 temperature between 140 to 160 °F, that is, the controlled variable mostly correlated with the butane product specification. The tray temperature however is primarily maintained by the regulation of the reflux flow, which is on a cascade control system. The minimum purity specification for butane is 97 vol%

with a maximum of propane and heavier pentanes content of 2 vol% and 1 vol% respectively.

The bottom product of the debutanizer column is the naphtha product. The main product specification of naphtha is the Reid Vapor Pressure RVP, which should not exceed 12 psia. The RVP of naphtha is lowered or increased by regulating the bottom temperature of the column. Increasing the bottom temperature of the column will strip lighter butanes from the heavier naphtha and thus lowers the RVP property and vice versa. The RVP is one of the missed opportunities in terms of continuous optimization. Naphtha product generally has a higher margin profit than butane. Lowering the RVP will result in more production of butane compared to naphtha. The operators of the unit generally leave a safety margin below the 12 psia RVP specification in order to avoid any off-specification production. However, this practice of operation results in lost opportunity since maintaining higher RVP (without violating the 12 psia limit) is economically favored. However, due to issues pertaining of better control without violating specification limits, operators generally prefer to operate away from the optimum zone.

A process flow diagram of the unit is given in figure 15.

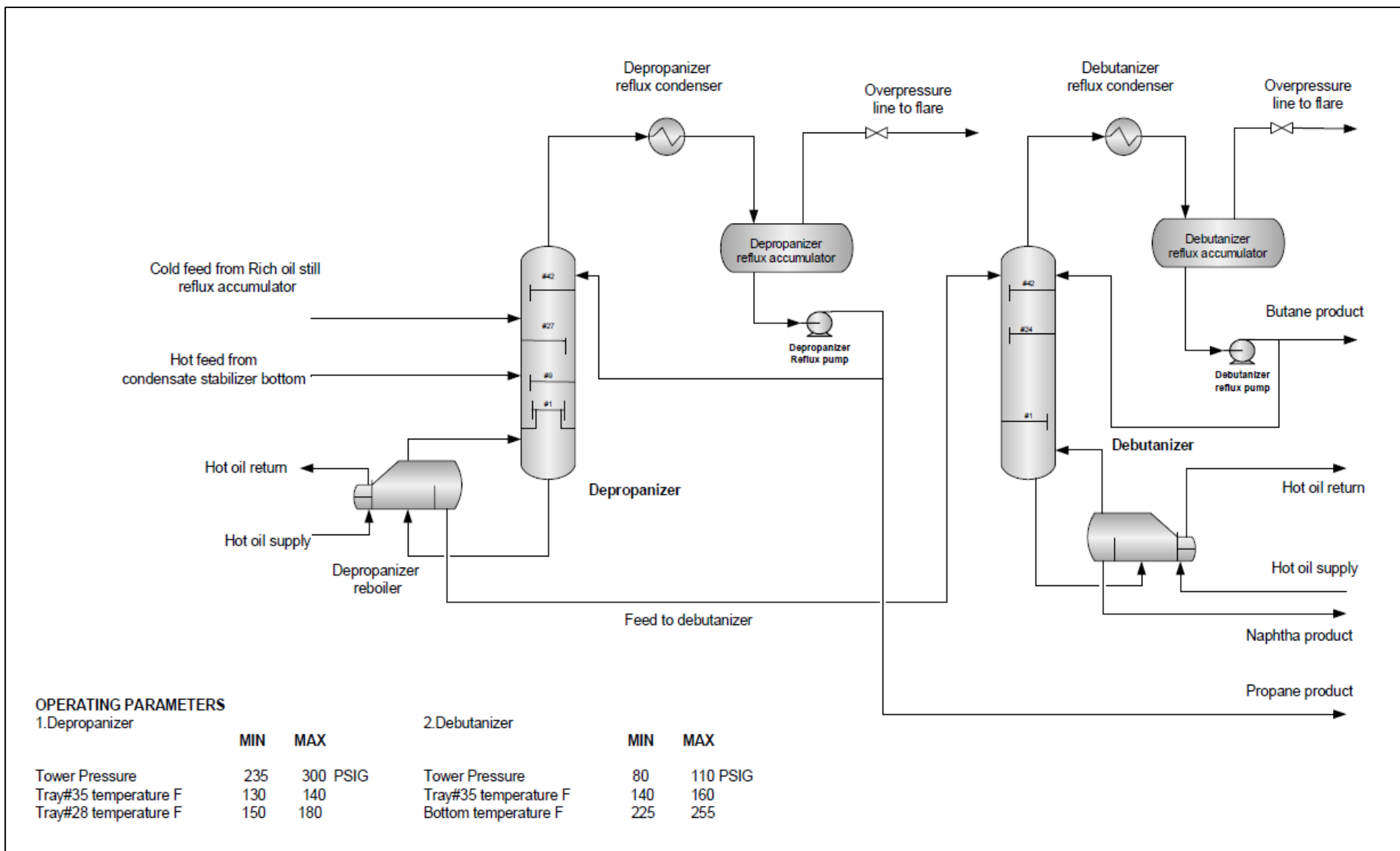


Figure 15 Fractionation train - Depropanizer and Debutanizer train

Chapter 4 – Process modelling

- 4.1. Methodology
- 4.2. Base case – Steady state simulation
- 4.3. Dynamic simulation
- 4.4. Functional design of MPC

4.1. Methodology

A primary objective of this thesis is to formulate the NGL control problem within the MPC framework in order to maximize benefits for the existing plant. As such, the results obtained would only be of value if the models used in the work were representative of the actual plant taking into account actual plant dynamics. While a dynamic model is required to generate the data needed to design the MPC controller, the workflow in Aspen HYSYS® for building dynamic simulation requires that steady state simulation of the plant be built first. All design parameters used for the steady-state simulation are available in Appendix A-1. The results henceforth obtained from the steady simulation was validated using actual official plant references. After obtaining satisfactory agreement, the next level of was to build the dynamic simulation that is necessary in order to realize and observe the transient plant responses taking into account actual equipment residence times and configuration. The simulation will be built using the Aspen HYSYS Dynamics. The dynamic simulation naturally incorporates the conventional multi-loop PID controllers. All design details for this dynamic simulation are available in Appendix A-2. The output from the second level of simulation will be validated with actual transient response records acquired from the plant DCS and compared. Finally, the dynamic simulation will incorporate the designed MPC on top of the conventional multi-loop controllers. The latter requires the identification of MVs (independent variables) and CVs (dependent variables). A proper understanding for the functional design and implementation of MPC system is crucial in order to achieve the seeking operational and economical objective. The functional design of the MPC controllers shall be done and the subsequent system architecture shall be deployed. The results arising from this shall

comprise the objective of this thesis. Results obtained shall be discussed and analyzed.

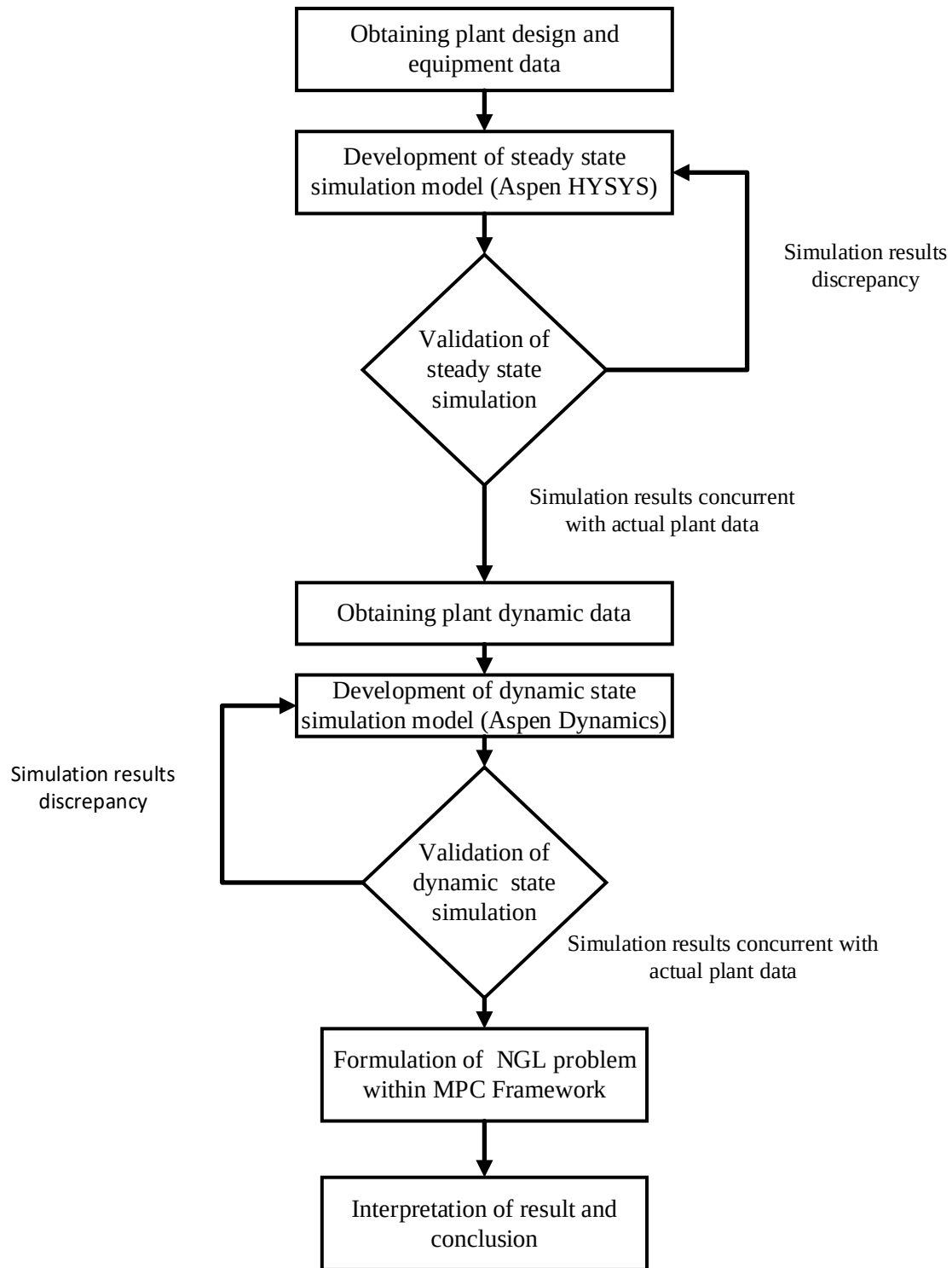


Figure 16 Methodology adopted in this thesis

4.2. Base Case – Steady state simulation

A steady state simulation model for the subject NGL plant has been built as per actual design data. The process flow diagram (PFD) of the model is given in figure 17 and figure 18. Complete Heat and Mass Balance (HMB) of the simulation case is given in Appendix A-3. Briefly, the workflow in building and validating the steady-state simulation involves the followings: (1) Define specification of product and intermediate product around columns, vessels, heat exchangers and rotating equipment from plant process design specifications – process material and heat balances. (2) Develop the steady simulation case using most recent plant data, (3) Obtain results and validate them using actual plant records and design conditions.

Results from the simulation will be compared for validation purposes on 3 main elements – column performance, heat exchanger performance and overall plant KPI. All design parameters for the steady case simulation model including feed basis are given in Appendix A-1.

4.2.1. Column performance

Column performance are primarily characterized by its temperature and pressure profiles, target separation quality and product flows. A validation of the simulation model built will be conducted by comparing values of parameters predicted by the simulations to that of actual designs values taken from the plant's official process data records.

Table 1 Column temperature profile - Simulation vs Plant design

Column temperature parameters	Simulation temperature (°F)	Column design temperature (°F)
Absorber bottom 204C-002		
- Bottom	-30	-30
Deethanizer 204C-003		
- Top	27	25 - 65
- Tray no. 9	215	155 - 260
- Bottom	349	320 - 375
Rich oil still 204C-005		
- Tray no. 2	484	435 - 535
- Tray no. 1	512	510 - 550
- Bottom	520	510 - 540
Depropanizer 204C-006		
- Top	145	115 - 130
- Tray no. 10	243	195 - 250
- Bottom	275	235 - 285
Debutanizer 204C-007		
- Top	144	135 - 150
- Tray no.37	165	140 - 160
- Bottom	255	225 - 255

Table 2 Column pressure profile – Simulation model vs Plant design

Column	Simulation pressure (psia)	Design pressure specification, (psia)
Absorber 204C-002	384	384
Deethanizer 204C-003	410	410
Rich oil still 204C-005	210	210
Depropanizer 204C-006	300	300
Debutanizer 204C-007	105	105

The below table summarizes the comparison between the simulation model and actual design in terms of columns' product rates and separation efficiency.

Table 3 Comparison column flow rates and performance – Simulation model vs Plant design

Column	Feed rate kmole/hr		Top product rate kmole/hr		Bottom product rate kmole/hr		Product Spec	
	Simulation	Design value	Simulation	Design value	Simulation	Design value	Simulation	Design value
Absorber 204C-002	8294.00	8291.10	7731.00	7717.20	1094.00	1175.70	Recovery - 67.7%	Recovery - 67.7%
Deethanizer 204C-003	1094.00	1175.70	339.20	384.20	755.00	664.90	C2 bottom content- 0.38%	C2 bottom content - 0.36 mol%
Rich oil still 204C-005	755.00	791.20	392.00	397.00	N/A	N/A	LPG content top - 82.35 mol% Lean oil molecular weight - 118 kg/kmole	LPG content top - 87.36 mol% Lean oil molecular weight - 124 kg/kmole
Depropanizer 204C-006	486.25	491.12	166.80	171.70	319.50	319.70	Propane Purity - 96.71 mol%	Propane Purity - 95.47 mol%
Debutanizer 204C-007	319.50	319.70	184.60	180.30	134.90	139.40	Butane Purity - 97.35 mol% Naphtha RVP - 9.321 psia	Butane Purity - 98.31 mol% Naphtha RVP - 9.515 psia (max 12 psia)

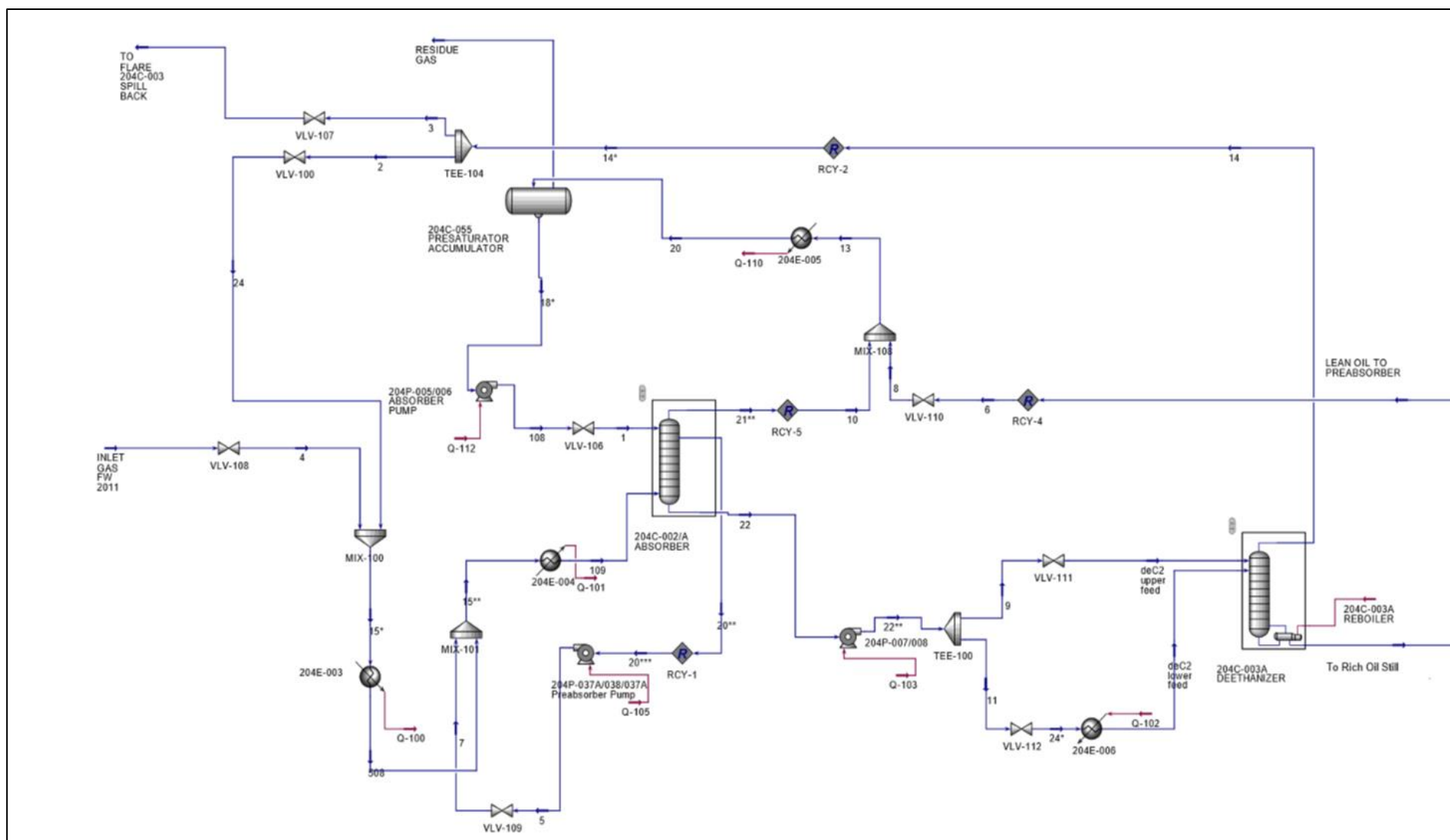


Figure 17 Process Flow Diagram PFD of the simulated model in Aspen HYSYS - Absorption and Deethanizer unit

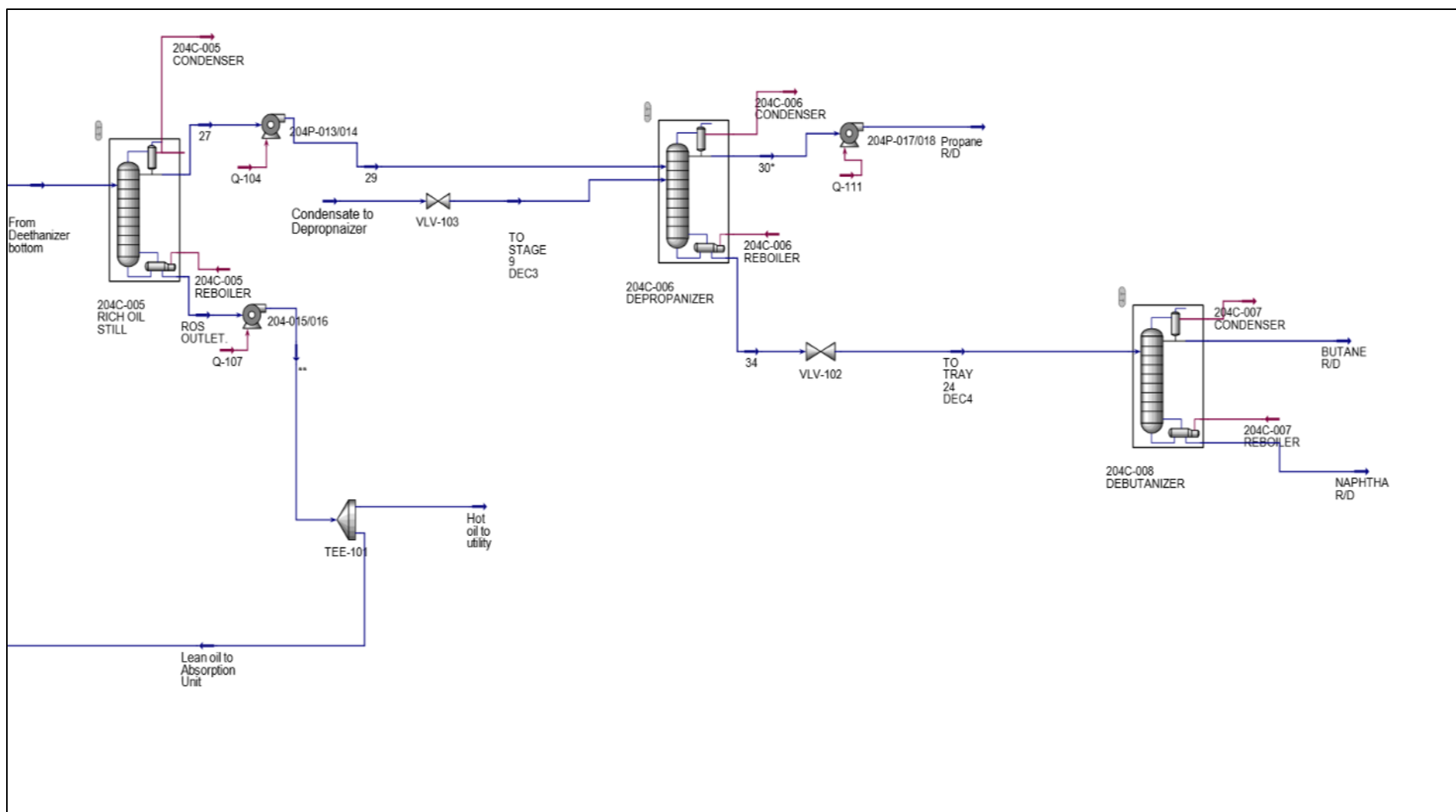


Figure 18 Process Flow Diagram PFD of the simulated model in Aspen HYSYS - Rich oil still, Depropanizer and Debutanizer

4.2.2. Heat exchanger performance validation

The following table provides a comparison overview of the heat exchanger duties between the simulated model and actual design.

Table 4 Heat exchanger duty comparison - Simulation model vs Plant design

No.	Name	Heat Duty, MMBtu/hr	
		Simulated Model	Design
1	Cold gas/gas heat exchanger 204E-003	9.44	9.99
2	Inlet chiller 204E-004	12.87	13.50
3	Pre-saturator chiller 204E-005	3.38	3.89
4	Absorber bottom heat exchanger 204E-006	7.56	8.15
5	Deethanizer feed heater 204E-500	8.32	8.98
6	Deethanizer reboiler 204E-008	18.60	16.96
7	Rich oil still condenser 204E-052	15.96	12.30
8	Depropanizer reboiler 204E-009	9.85	9.00
9	Depropanizer reflux condenser 204E-053	9.21	8.40
10	Debutanizer reboiler 204E-011	7.00	6.81
11	Debutanizer reflux condenser 204E-054	9.62	9.50

4.2.3. Overall plant KPI

From the above and having demonstrated equipment wise validation between the developed steady state model and the plant's actual design data, a further validation for the simulation model shall be illustrated by comparing plant wise KPIs which have been developed by the company to monitor their daily plant performance. An overview of these KPIs shall be given below:

1. Propane recovery: This is considered the most important KPI in the subject NGL plant as this indicates the percent of product recovered from the feed. The propane recovery is calculated by the following:

$$C_3 \text{ recovery}\% = \frac{(\text{Total } C_3 \text{ amount in feed} - C_3 \text{ amount in residue gas}) \times 100}{\text{Total } C_3 \text{ amount in feed}}$$

2. C3% in residue gas: This indicates the amount of slippage of propane product into the residue gas leaving the plant. The higher the slippage of propane, the higher is the economic loss.
3. Liquid production – This KPI represents the total liquid barrels of propane, butane and naphtha. This figure generally varies depending on the feed gas quality especially from refinery off gas (ROG).
4. C2% in Deethanizer bottom – This KPI represents the performance of ethane stripping from the product. The higher the slippage into the bottom stream, the higher are the chances of producing off specification production and hence increasing cost for reprocessing.
5. Final product purity – This KPI indicates the purity percentage of each product.

Table 5 Plant wide KPI comparison - Simulation model vs Plant design

KPI	Simulation	Design
Feed gas rate MMSCFD	161.4	161.4
Condensate rate std barrel/day	1884	1866
Production		
Propane std barrel/day	2183	2251
Butane std barrel/day	2818	2742
Naphtha std barrel/day	2672	2763
Total production std barrel/day	7673	7756
C3% recovery	81.03	83.22
C3 mol% in residue gas	0.50	0.68
C2 mol% in Deethanizer bottom	0.30	0.36
Product Purity		
Propane mol%	96.71	95.47
Ethane in propane mol%	1.36	1.64
Butane mol%	97.33	98.31
Butane in propane mol%	1.64	2.54

4.3. Dynamic simulation

After completing and validating the steady state simulation model, the next step is to build a representative dynamic simulation model of the NGL plant. The objective of the dynamic simulation is twofold. First, the dynamic model is necessary in order to capture plant dynamics in response to manipulated variables moves and disturbances and is a prerequisite to evaluate the effectiveness of MPC to the plant. Second, the validated dynamic simulation model will serve as a virtual plant from which dynamic data will be generated to identify the internal model of the MPC. It is of crucial importance that the dynamic model built is cross-validated and verified so that the simulated and real dynamic characteristic of the actual plant are matched. In essence, the dynamic model can be considered as the digital twin of the plant. In order to build the dynamic model, the following data are considered essential. These data have been obtained from actual plant data: (1) Equipment and Plantwide control structure, (2) Column geometry and residence time, (3) Vessel geometry and residence time.

Complete details concerning the above elements can be found in Appendix A-3.

The control structure employed will be discussed in the below section. It is quite important to properly understand the control structure in order to build a model that will be representative and reflect the real transient response of the actual plant.

4.3.1. Control structure configuration and Dynamic simulation model development

4.3.1.1. Absorber and Deethanizer unit

The development of the control structure employed in the dynamic model is based on actual control configurations. All liquid holdups and residence times are derived from actual equipment dimensions. The control configuration of this unit is given in figure 19. The corresponding dynamic simulation model developed is given in figure 20.

The multi-loop control structure for this unit are briefly explained below:

- Liquid feed to 204C-002 column is flow controlled by manipulating control valve FV-018 downstream of 204P-005/006 Absorber feed pump.
- Lean oil flow to absorber unit is flow controlled by manipulating control valve FV-014 downstream of 204P-009/010 lean oil pump.
- 204C-002 absorber below chimney tray liquid hold up is controlled by level valve LV-031 downstream of pump 204P-037/038 Pre-absorber feed pump.
- Cold liquid feed flow to 204C-003 Deethanizer column is controlled by level control valve LV-023 on 204C-002 Absorber.
- Hot liquid feed to 204C-003 Deethanizer column is flow controlled by control valve FV-034.

- Propane recovery is affected by manipulation of lean oil flow through FV-014 together with temperature of absorber, which is affected of cold utility flow to the exchangers 204E-004 and 204E-005.
- Pressure of 204C-003 Deethanizer tower is controlled by pressure control valve PIC-186A. Excess pressure can be relieved to flare through over pressure control valve PV-186B that is operated in split range control with PCV-186A.
- Bottom flow from 204C-003 Deethanizer is regulated by tower bottom level control via control valve LV-036.
- Target composition of ethane vol% in final product propane should not exceed 2 vol%. This is achieved by maintaining tray no. 9 temperature of the Deethanizer 204C-003 between 155-260 °F. The higher the temperature of tray no.9, the higher is the ethane stripping from the product while more propane product will be lost in the overhead and vice versa. Maintaining the temperature of this tray is done by controlling hot utility flow to the Deethanizer reboiler. Bottom temperature of the Deethanizer is maintained between 320-375 °F and is achieved by temperature control valve TV-066.

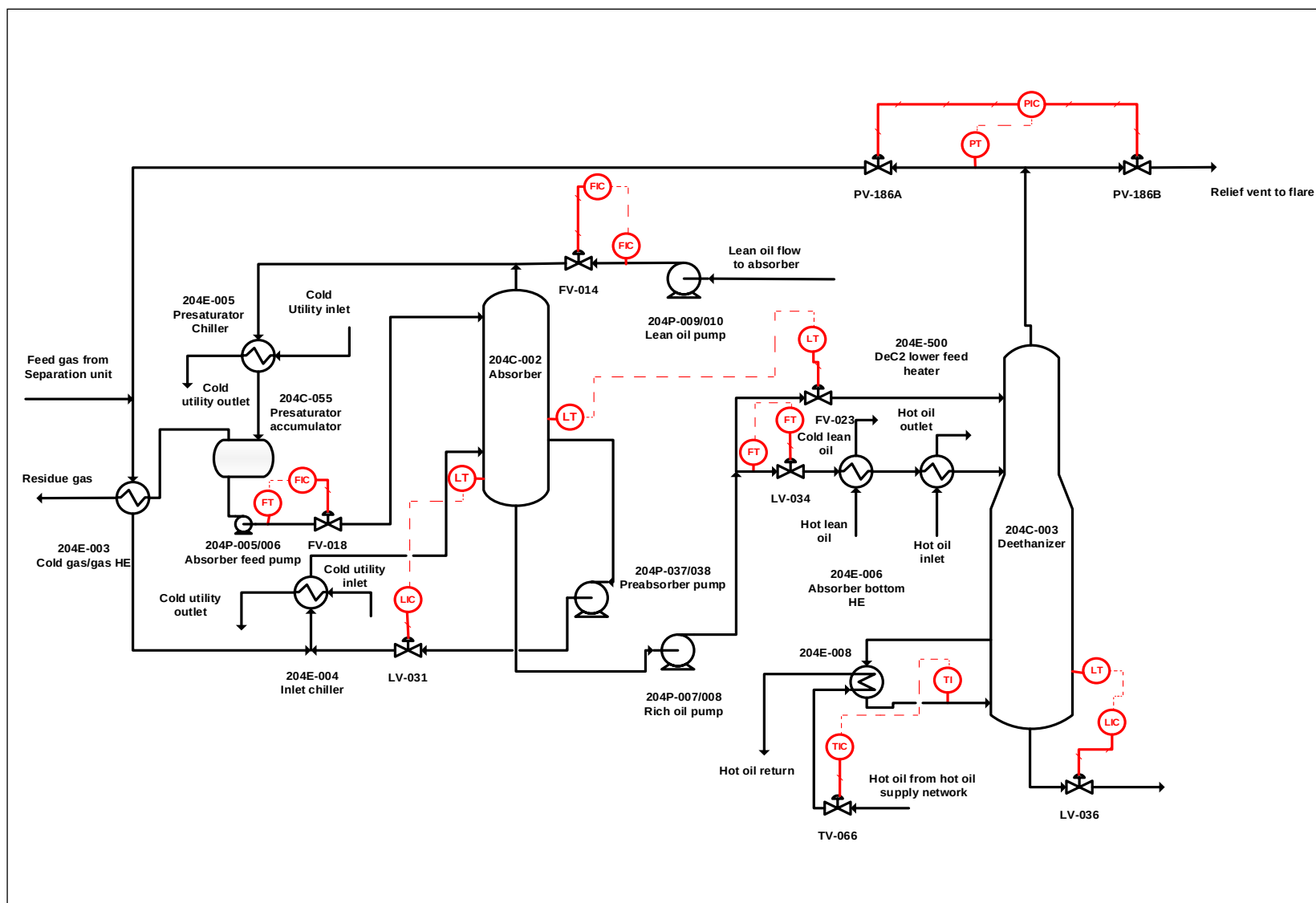


Figure 19 Schematic of the Absorber and Deethanizer unit control structure

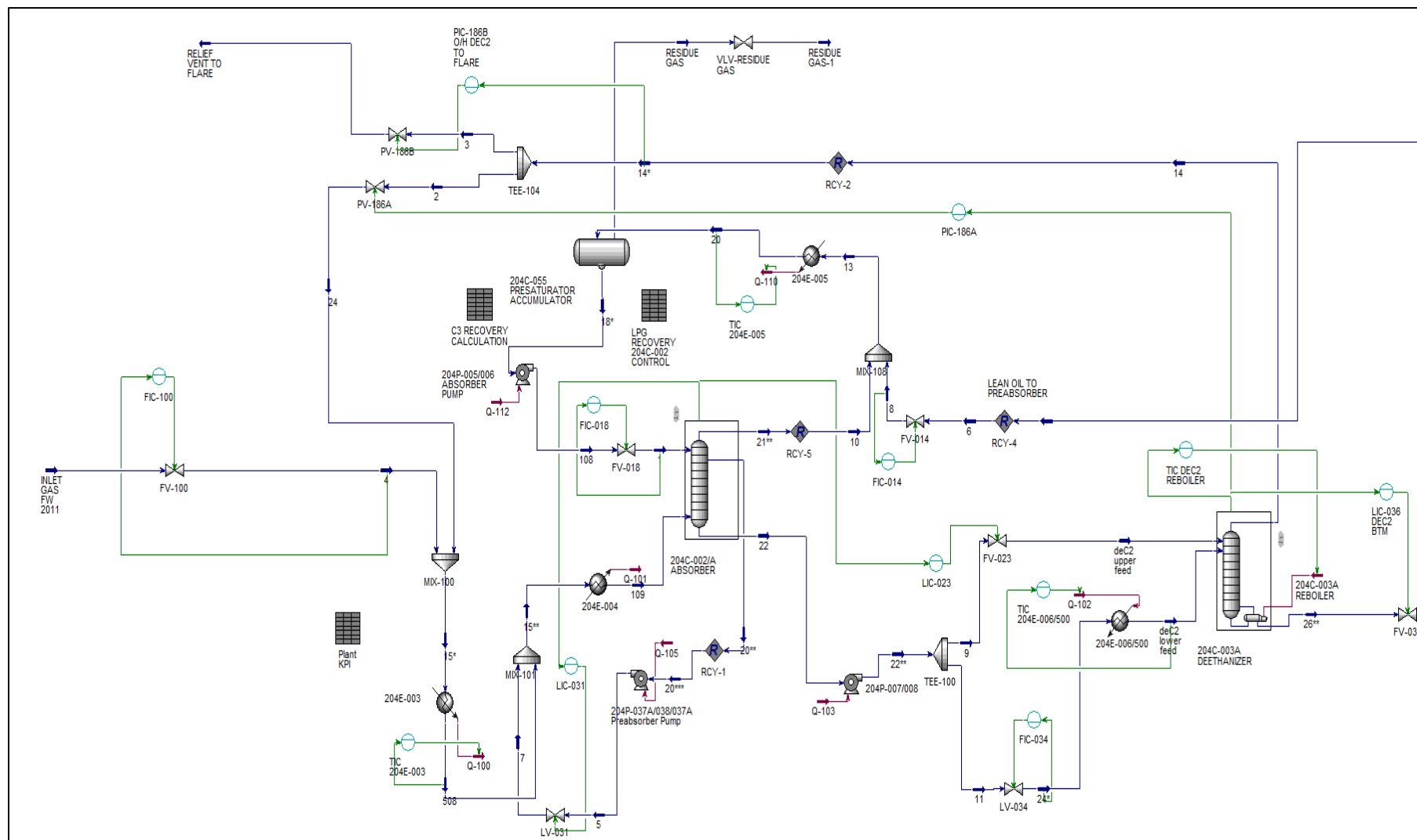


Figure 20 Absorber and Deethanizer unit - Aspen HYSYS Dynamic simulation model

4.3.1.2. Rich oil still unit

Figure 21 shows the control configuration employed for the dynamic simulation of the Rich oil still unit, which is derived from actual configuration, equipment data and control structure. The corresponding dynamic simulation model developed is given in figure 22. The control loops are briefly explained below:

- Feed flow to 204C-005 Rich oil still unit is level controlled via LV-036 which regulates level of 204C-003 Deethanizer bottom level.
- Reflux flow is flow controlled via FV-031. The reflux rate is not varied for the control of the top temperature. Therefore, the top temperature varies in temperature between 190 and 230F depending on the feed rate and bottom temperature being maintained. Very high reflux tends to flood the upper section and inadequate ratios will result in poor fractionation.
- Control of bottom temperature is done TIC-086 which maintains the temperature of tray no. 2 between 435 – 535 F by maintaining the temperature of the hot oil outlet from the furnace between 510-550. It is cascaded to fuel gas pressure regulator supplying the furnace thereby maintaining amount of energy supplying the hot oil system. The normal operating temperature of at the bottom of the column is a few degrees lower than the

temperature of the hot oil from the furnace. Lower bottom temperatures may result in inefficiency in separating some of the components of the lean oil and will result in higher demand for hot oil flow through the reboilers. High bottom temperatures on the other hand may result in vaporization of certain components from the lean oil and will result in lower liquid hold up.

- Pressure of 204C-005 Rich oil still is controlled by PIC-049 via PV-049 regulating back pressure buildup towards the column. Normal operating pressure of the tower is between 185 and 210 psig. Lower tower pressures will result in an increase in vaporization rate which will demand a reduction in bottom temperature in order to compensate. This reduction in bottom temperature is achieved by drawing more oil from the furnace while keeping fuel rate constant which is an undesirable operation. Higher tower pressures will result in a suppression of vaporization rate which will lead to lower/lighter lean oil and hence affecting the absorption unit.
- Top product flow is level controlled by LIC-032 via control valve LV-032 regulating the liquid level hold up in the 204C-056 reflux accumulator.
- Pressure of 204C-056 reflux accumulator is maintained between 170 to 190 psig via PV-056 in case of excess pressure.

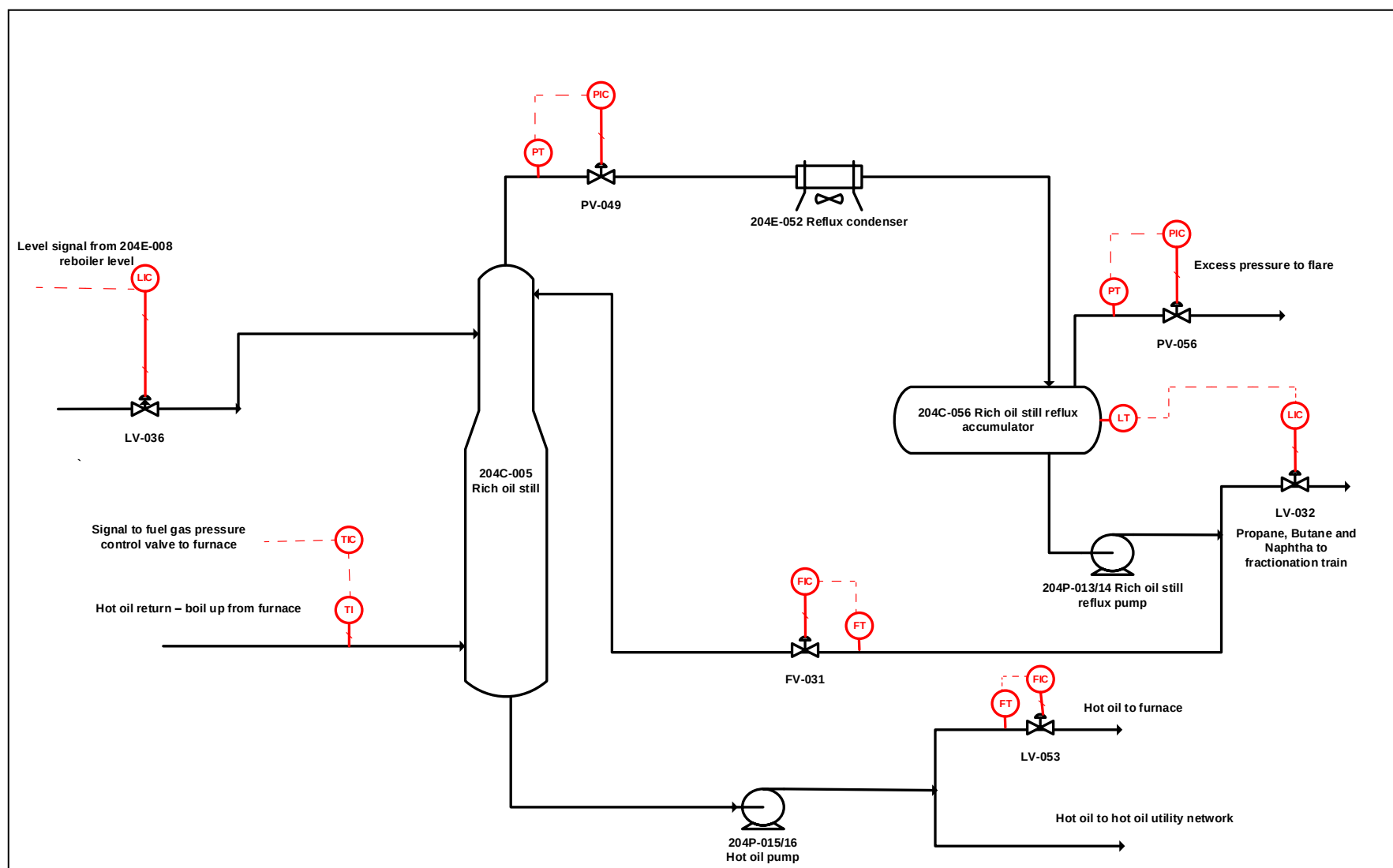


Figure 21 Schematic of the Rich oil still unit control structure

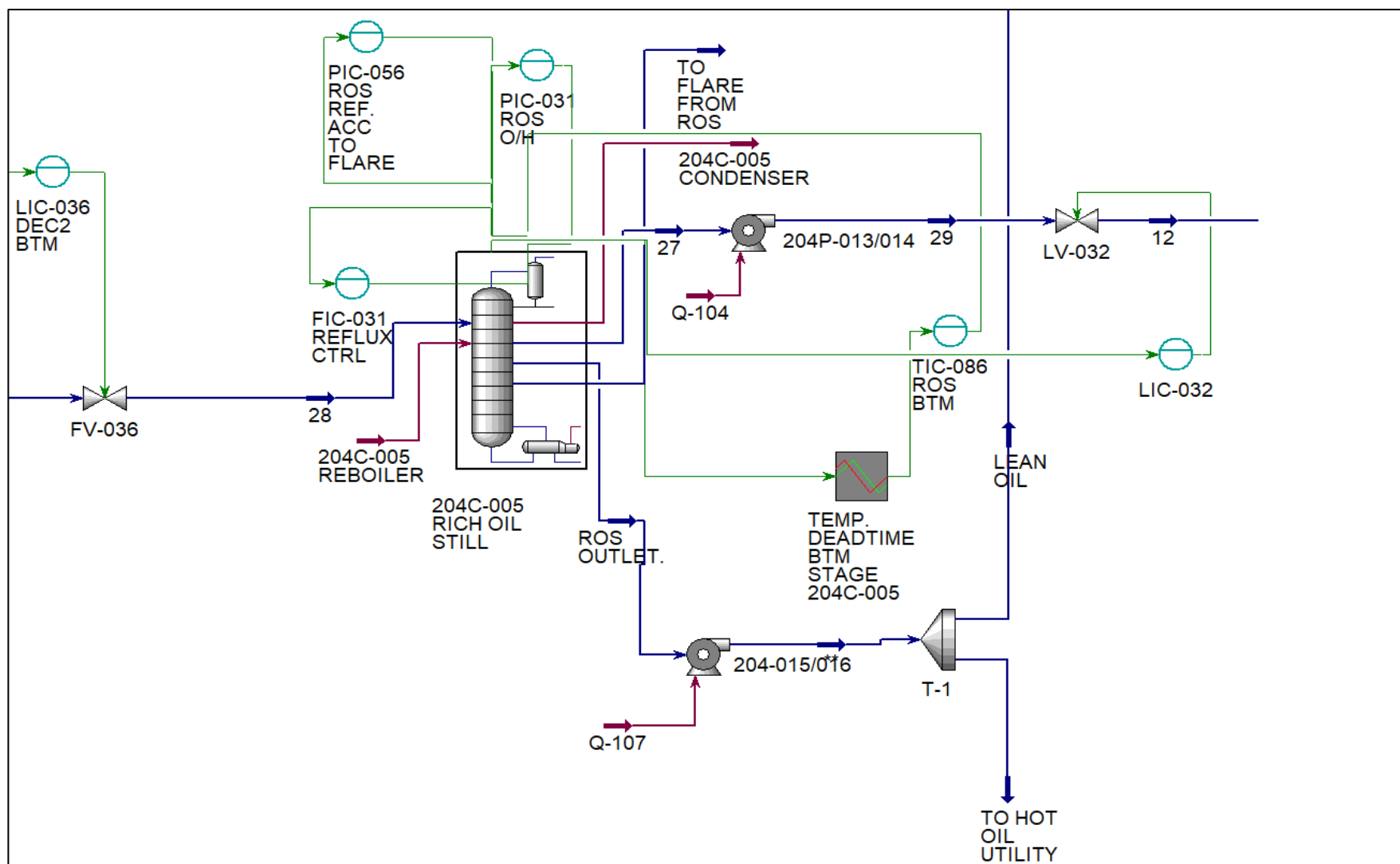


Figure 22 Rich oil still unit - Aspen HYSYS Dynamic simulation model

4.3.1.3. Depropanizer unit

Figure 23 shows the control configuration employed for the dynamic simulation of the Depropanizer unit, which is derived from actual configuration, equipment data and control structure. The corresponding dynamic simulation model developed is given in figure 24.

The specific control loops in this unit are explained below:

- Cold feed to the column is level controlled via LIC-032 by manipulating control valve LV-032 which regulates the liquid level in 204C-056 Rich oil still reflux accumulator. However normally this is kept on manual control and varied accordingly when fluctuations are observed. Sudden decrease in cold feed flow rate would result in the middle section temperature of the tower going up and adjustment to the reboiler must be made. Sudden increase in cold feed will result in drop in middle section temperature. More propane components will escape with the heavier bottom components. Adjustment to the reboiler heat duty will be needed in order to rectify temperature profile of the column. Sustaining of high cold feed rate for a long period of time will cause flooding of the tower which is indicated in the sharp rise in tray no. 35 temperature.
- Hot feed to the column is level controlled via LIC-008 by manipulating control valve LV-008 which regulates the 204C-001 stabilizer tower bottom level control.

Since the flow is under level control, certain fluctuations are expected. Small fluctuations in flow will not affect tower operations. However, large variations in flow will upset the temperature profile and appropriate adjustment to the reboiler and reflux must be made in order to rectify the temperature profile of the tower. Sudden decrease in the hot feed flow rate will cause bottom temperature to decrease and will affect the vaporization of cold feed as the vapor traffic in the tower will reduce considerably. Adjustment must be made to reboiler duty level and rectify the temperature profile. Sudden increase in hot feed flow rate will result in increase in the tower bottom temperature and reboiler duty level will need to be reduced appropriately.

- Tower pressure is controlled via PIC-086, which manipulates control valve PV-086 in order to maintain tower pressure between 240 and 300 psig. Lower tower pressure will cause increased tendencies for heavier components such as butane to vaporize and will result in off spec propane product spec. Higher tower pressure will reduce propane stripping and may cause excess slippage of propane along with the bottom product.
- An important control element from this tower from control point of view is the temperature of tray no. 35 and 28. Optimal control of this tray temperature will result in on spec production. Tray no. 35 temperature is regulated via TIC-052 which is cascaded with FIC-052

manipulating FV-052 which controls reflux flow to the column. Tray no. 28 temperature varies with the control of tray no.35. Optimum temperature range of tray no. 35 is between 130 and 145 F while for tray no. 28 it is between 150 and 180 F. Lower temperature profiles will cause reduction in propane evaporation and will cause slippage of propane product along with the bottom product and possibly causing off spec butane production. Higher temperature profiles will have the effect of increased vaporization of heavier components such as butane and can cause off spec production of propane product.

- Control of tower bottom section temperature is done by varying hot oil flow to 204E-009 Depropanizer reboiler via FIC-045 by manipulating FV-045 to maintain tower bottom and tray no. 10 temperature between 235 – 285 F and 195 – 250 F respectively. Lower temperature profile of the tower will not be sufficient to vaporize all propane product from the bottom and hence causing excessive slippage of propane along with the heavier bottom products. Higher temperature profiles can cause excessive boil off of heavier components from the bottom resulting in tower flooding and off spec production.
- Over pressure control for 204C-057 Depropanizer reflux accumulator is provided via PIC-089 which

manipulates PV-089 in case of pressure exceeding 270 psig thereby relieving to flare.

- Level controller is also provided for 204C-057 which regulates liquid flowrate through LIC-051 by manipulating LV-051 physically located downstream of the sulphur treatment unit. Since the treatment units have been excluded from the simulation work scope, the control loop has been relocated to control flow of top propane product with appropriate tuning parameters in order to have a representative simulation.
- Bottom flow to 204C-007 debutanizer is level controlled via LIC-048 by manipulating control valve LV-048 to maintain liquid level in the reboiler.

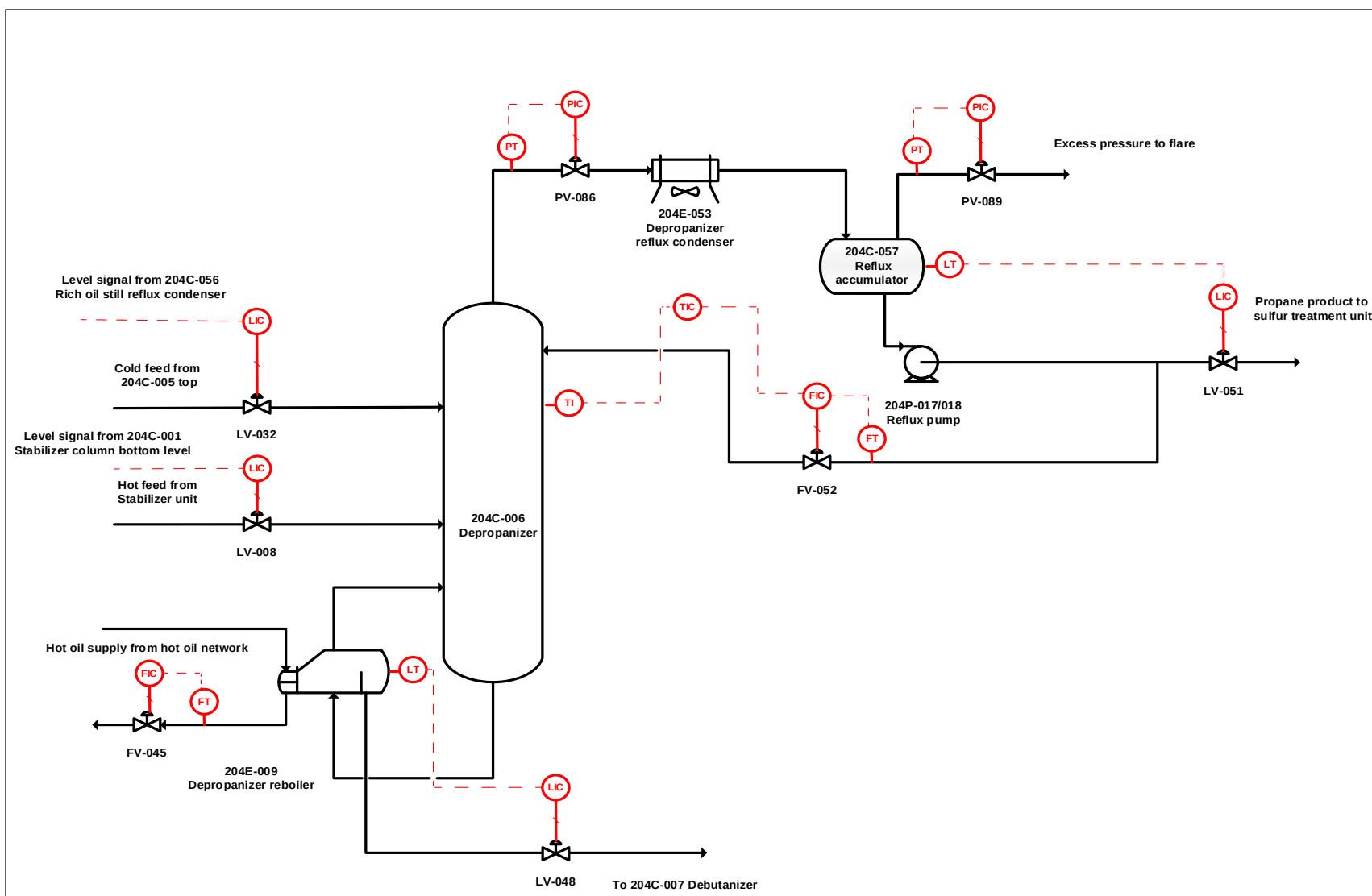


Figure 23 Schematic of the Depropanizer unit control structure

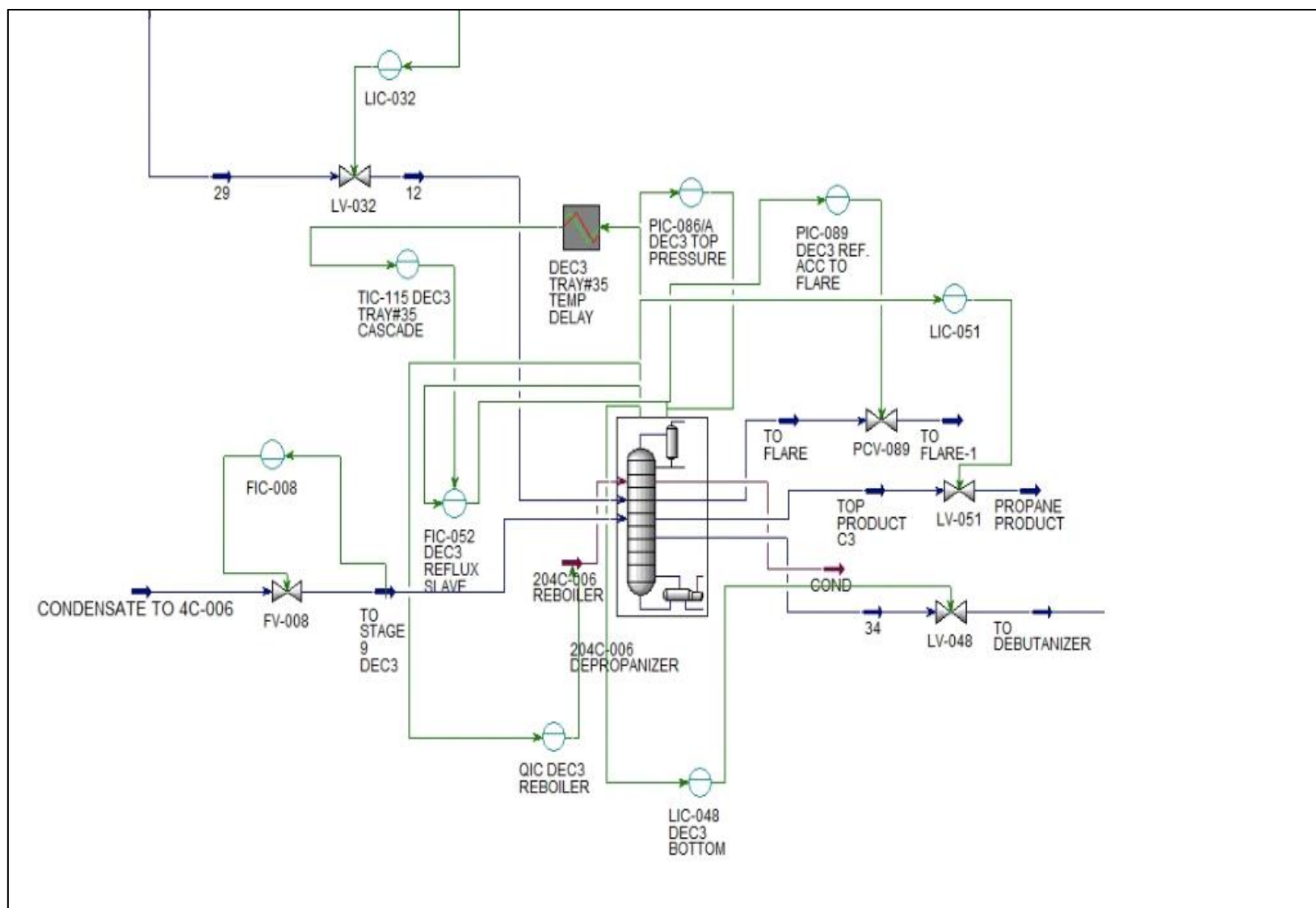


Figure 24 Depropanizer unit – Aspen HYSYS Dynamic simulation model

4.3.1.4. Debutanizer Unit

Figure 25 shows the control configuration employed for the dynamic simulation of the Debutanizer unit which is derived from actual configuration, equipment data and control philosophy. The corresponding dynamic model developed is given in figure 26. The specific control loops in this unit are given below:

- Feed to the column is level controlled via LIC-048 which manipulates control valve LV-048 to regulate liquid level in 204E-009 Depropanizer reboiler. High or low feed to the column affects the temperature profile of the tower and may result in off spec of both products – Butane and Naphtha.
- Top tower temperature is maintained between 135 to 150 F. However, the point of temperature control is tray no. 37 which has the heaviest effect on product quality. This tray temperature is maintained between 140 and 160 F by temperature controller TIC-128 which is cascaded with reflux flow controller FIC-058 regulating FV-058 reflux flow to column. Higher temperatures can cause the pentanes to boil off overhead causing off specification of Butane product while lower temperatures will result in insufficient boiling thereby increasing RVP specification of Naphtha beyond allowed specification.

- Column pressure is maintained between 80 to 110 psig via PIC-093 by regulating pressure control valve PV-093 which maintains back pressure on the column. High column pressure will tend to reduce vaporization and affect separation of vapor overhead. This may result in off specification production of Naphtha due to high RVP. Lower column pressure has the reverse effect and can accelerate the vaporization of heavy fractions and hence may result in off spec production of Butane product.
- Tower bottom temperature is maintained between 225 to 255 F to regulate the Naphtha RVP of 12 psia (max). This is done by manually regulating FV-053 – hot oil utility supply to 204E-011 Debutanizer reboiler. High bottom temperature will allow cause excessive vaporization of heavier components and will cause off specification production of butane. Lower temperature will result in slippage of butane into the bottom and will increase naphtha RVP thereby off specification production.
- Over pressure control for 204C-058 Debutanizer reflux accumulator is provided via PIC-096 which manipulates pressure control valve PV-096 in case of pressure exceeding 85 psig thereby relieving to flare.
- Level controller is also provided for 204C-058 debutanizer reflux drum which regulates liquid flowrate through LIC-055 by manipulating level control valve LV-055 physically located downstream of the sulphur treatment unit. Since the treatment units have been

excluded from the simulation work scope, the control loop have been relocated to control flow of top butane product with appropriate tuning parameters in order to have a representative simulation.

- Bottom flow from 204C-007 debutanizer is level controlled via LIC-053 by manipulating control valve LV-048 to maintain liquid level in the reboiler.

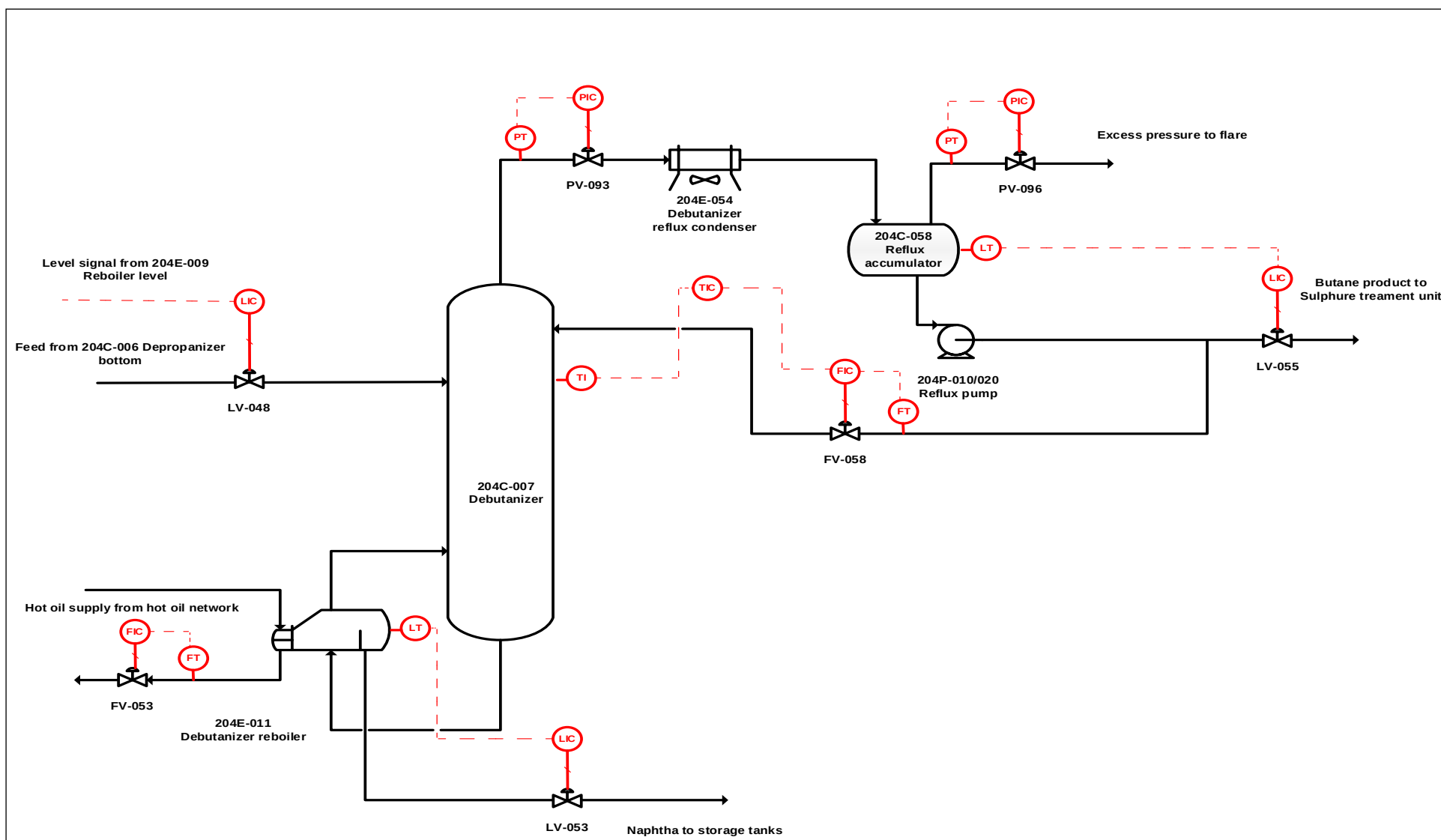


Figure 25 Schematic of the Debutanizer unit control structure

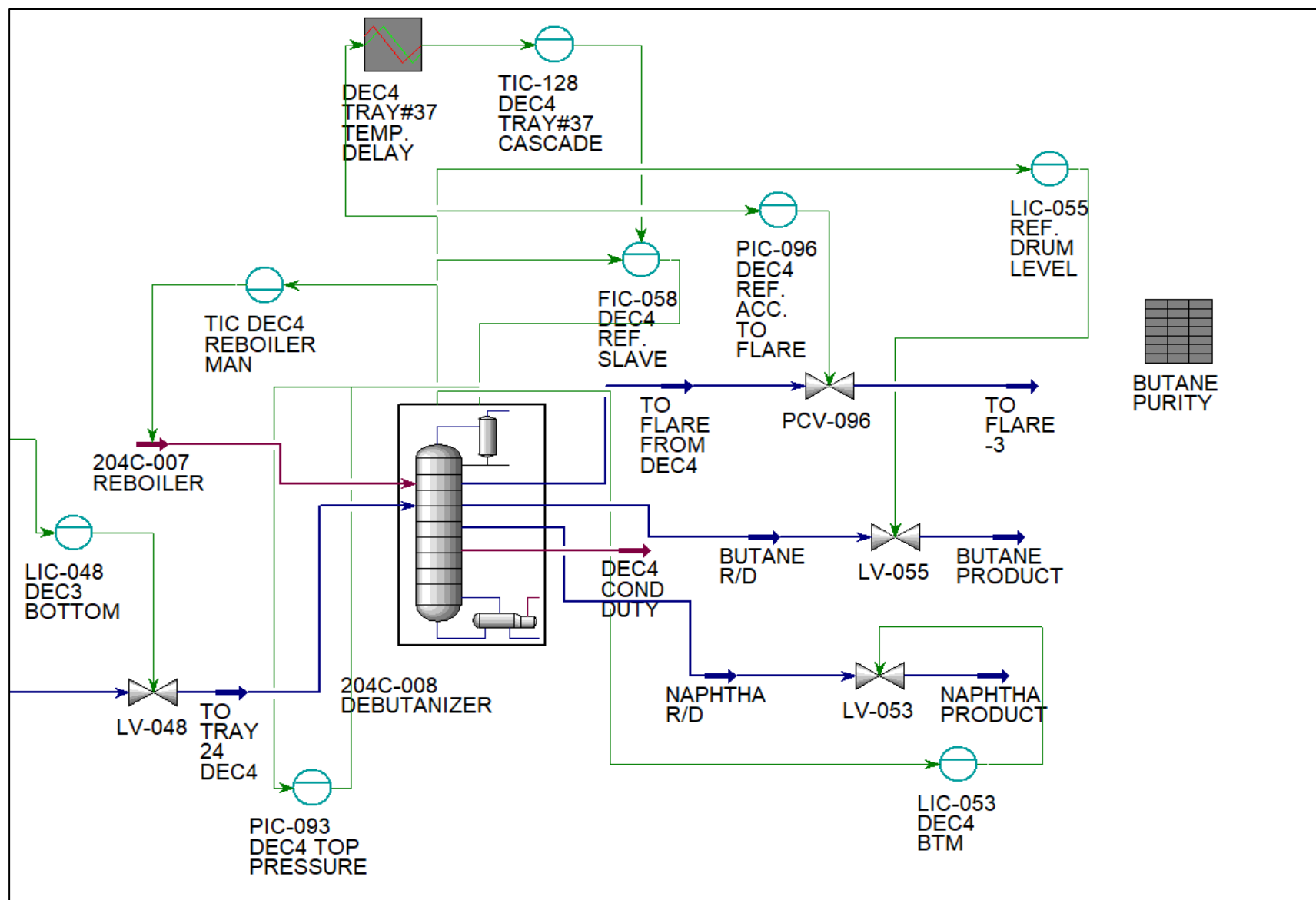


Figure 26 Debutanizer unit – Aspen HYSYS Dynamic simulation model

4.3.2. Dynamic simulation validation with real data from the NGL plant

This section will present result comparison between the dynamic simulation model and actual plant behavior. For each of the following units, key parameters that are crucial for operational control and monitoring have been selected for comparison and to provide basis for validation of the dynamic model. The dynamic simulation is the backbone upon which the MPC shall be built on. Hence, it is of utmost importance that the validation shall be done in order to ensure the fidelity of the plant's model.

4.3.2.1. Deethanizer column

Key crucial parameters for the modelling and operation of the 204C-003 Deethanizer column are given below:

- Column overhead temperature
- Tray no. 2 temperature
- Tray no. 9 temperature
- Tray no. 16 temperature

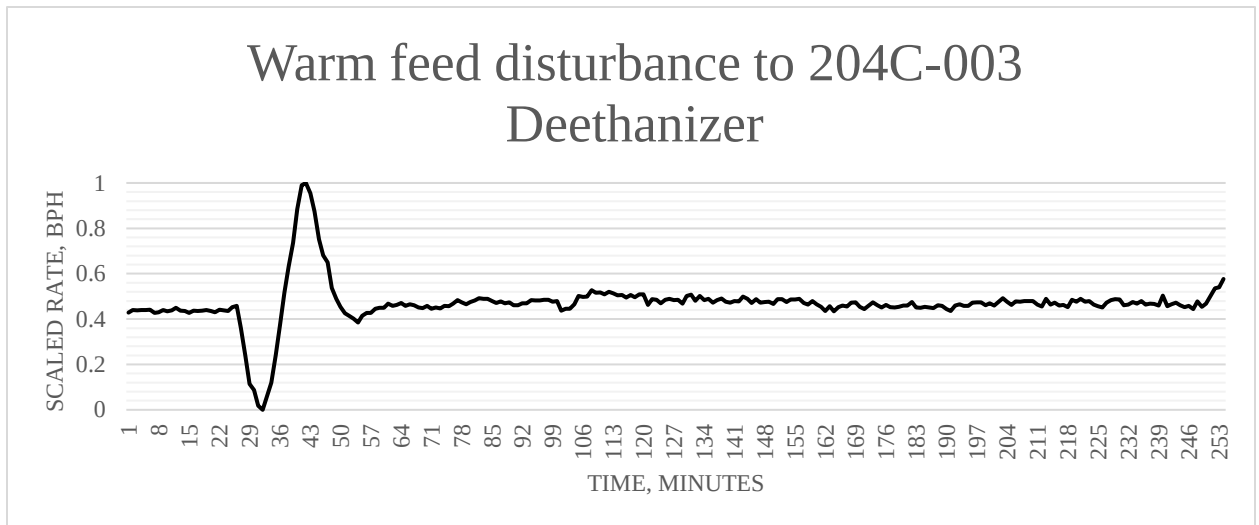


Figure 27 Hot feed disturbance - Data taken from plant historical trends

A feed flow rate disturbance from the plant operational historical trend of the hot feed to the column was taken and a similar disturbance was dynamically simulated in order to test and compare the dynamic response.

Below are the corresponding dynamic behavior trends of the column – both simulation model and actual plant behavior.

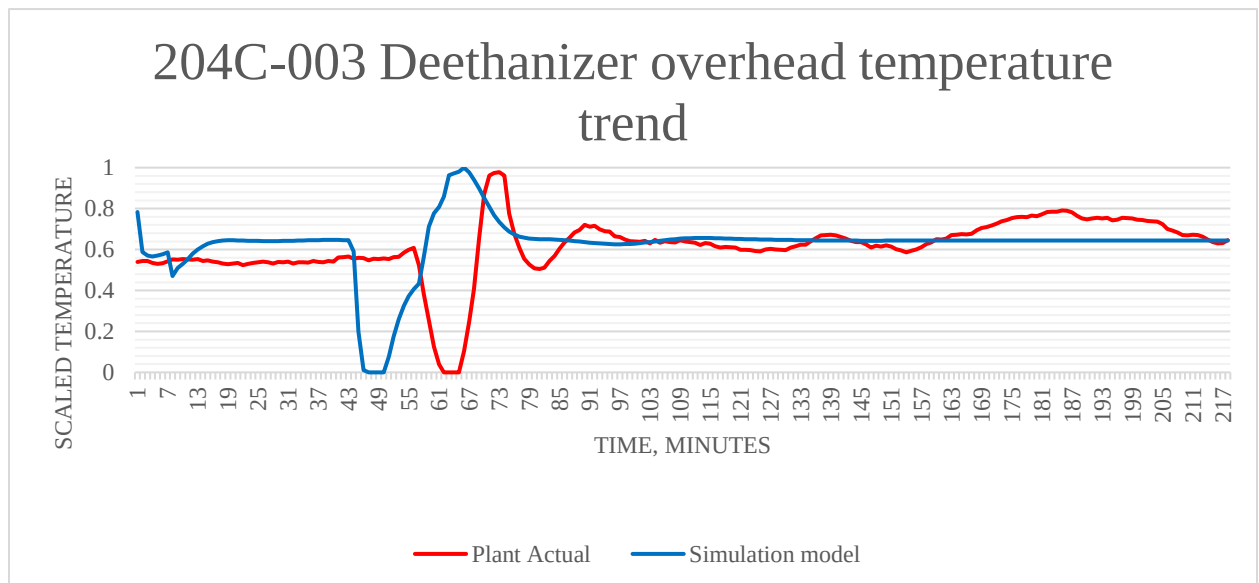


Figure 28 Deethanizer overhead temperature trend – Aspen HYSYS Dynamic model vs Actual

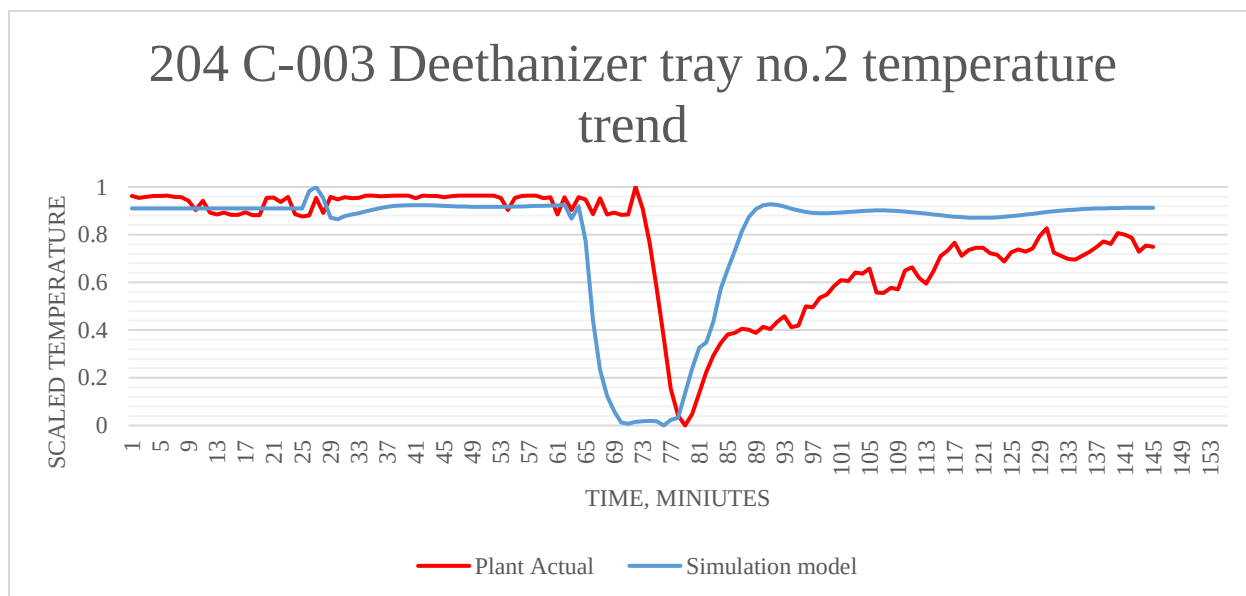


Figure 30 Deethanizer tray no.2 temperature trend – Aspen HYSYS Dynamic model vs Actual

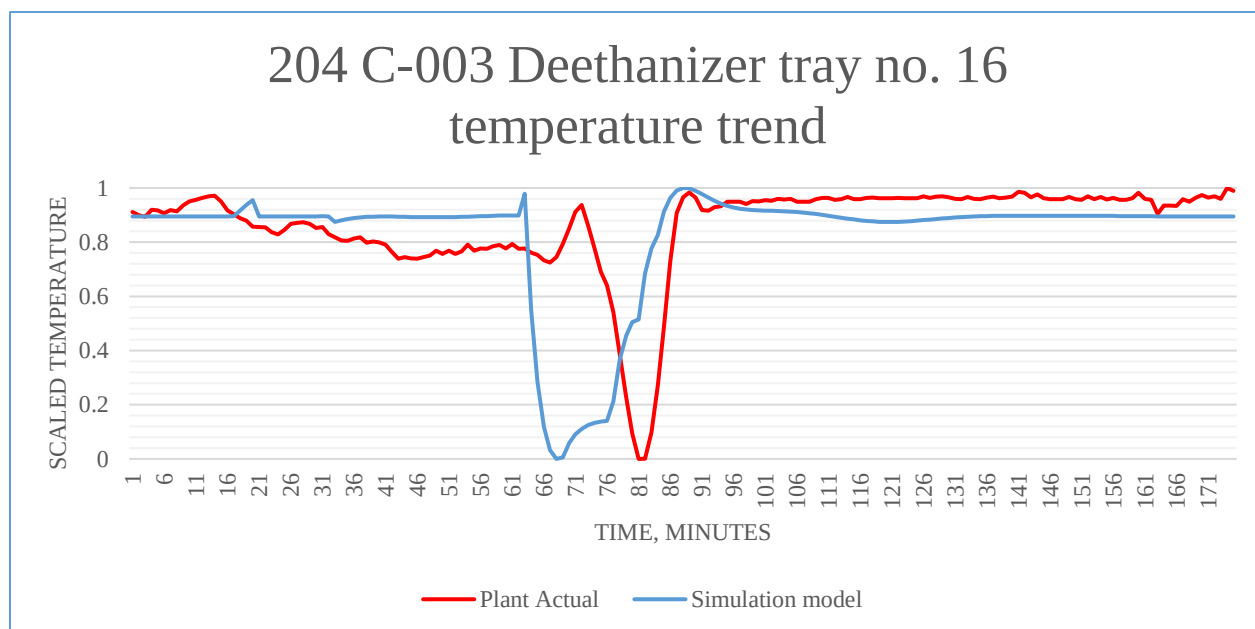


Figure 29 Deethanizer tray no. 16 temperature trend – Aspen HYSYS Dynamic model vs Actual

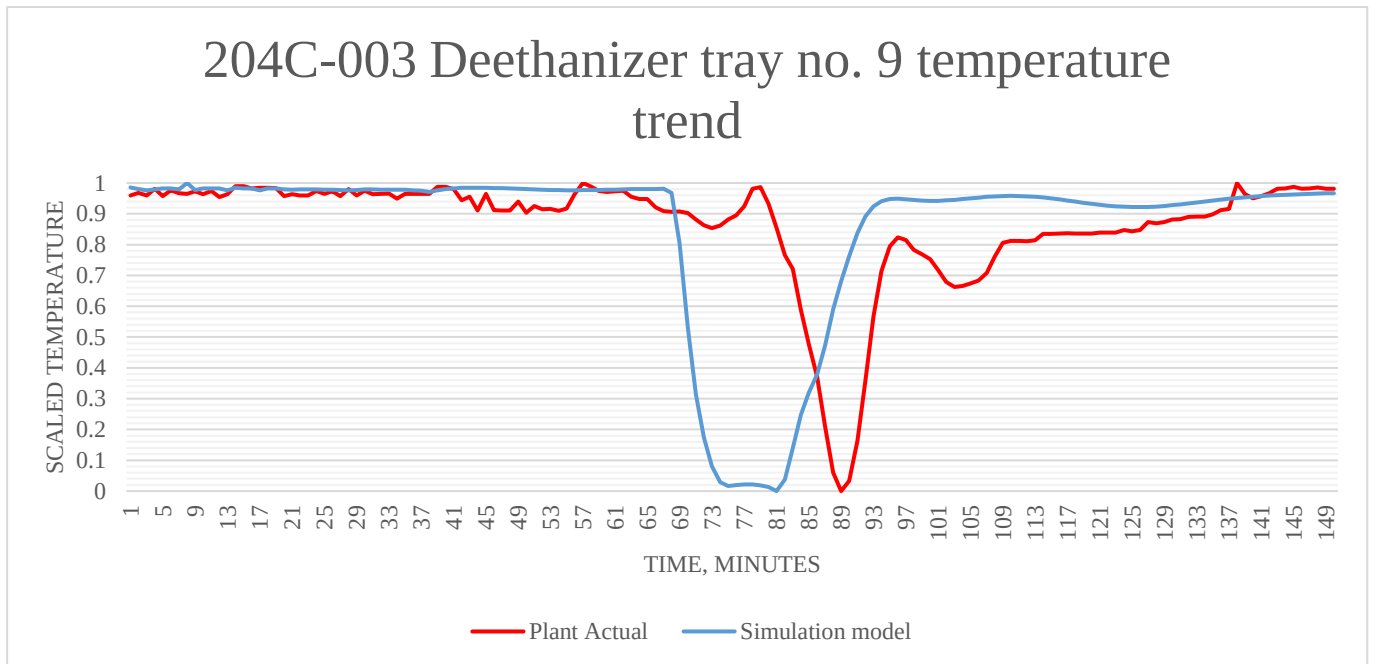


Figure 31 Deethanizer tray No.9 temperature trend – Aspen HYSYS Dynamic model vs Actual

4.3.2.2. Rich oil still column

Key crucial parameters for modelling and operation of 204C-005 Rich oil still column are given below:

- Column overhead temperature
- Tray no. 2 temperature
- Tray no. 22 temperature
- Tray no. 25 temperature

A feed flow rate disturbance from the plant operational historical trend of the feed to the 204C-005 Rich oil still column was identified, and a similar disturbance was simulated in the dynamic model in order to compare dynamic behavior.

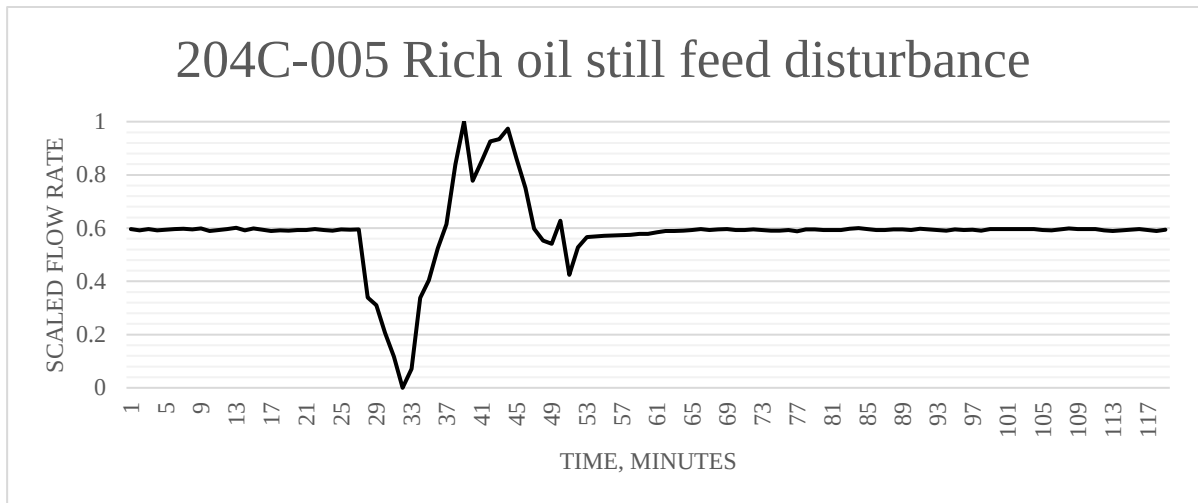


Figure 33 Rich oil still feed disturbance - Data taken from plant historical trends

Below are the corresponding dynamic behavior of the column – both simulation model and actual plant behavior.

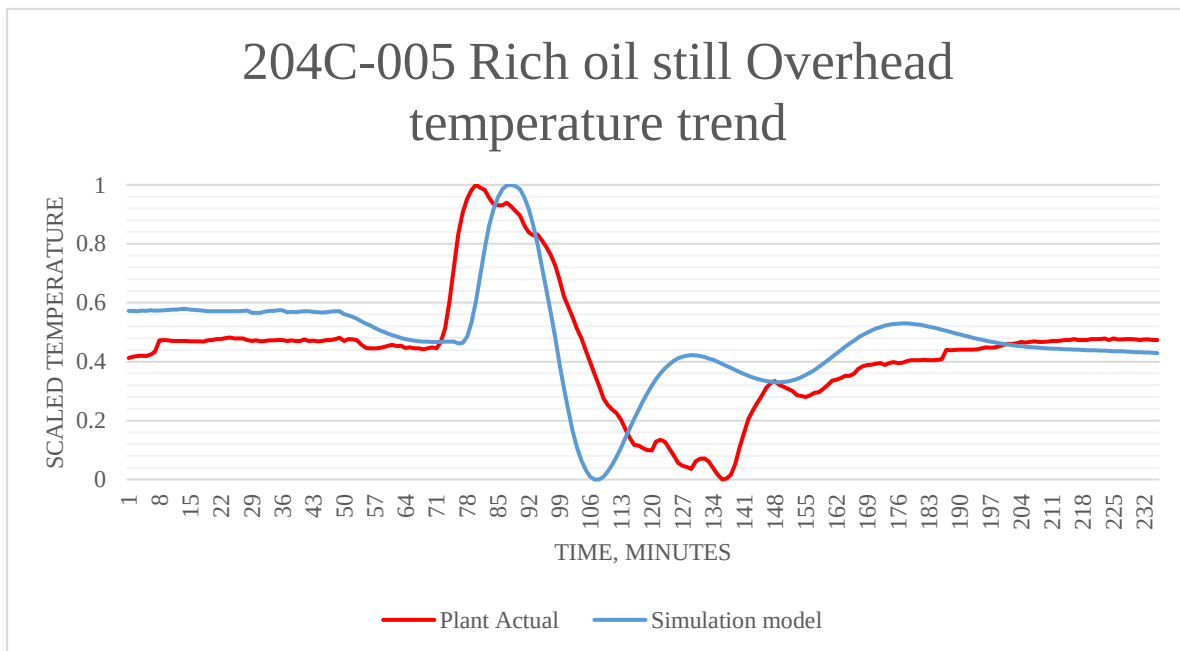


Figure 32 Rich oil still overhead temperature trend – Aspen HYSYS Dynamic model vs Actual

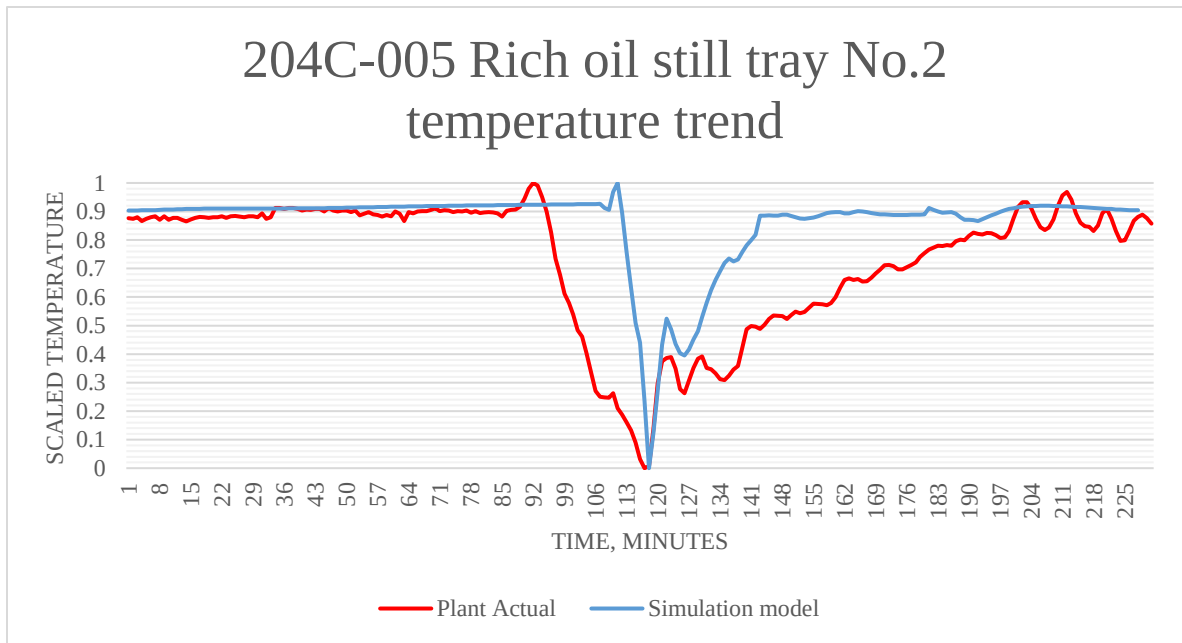


**Figure 34 Rich oil still tray No. 2 temperature trend – Aspen HYSYS
Dynamic model vs Actual**

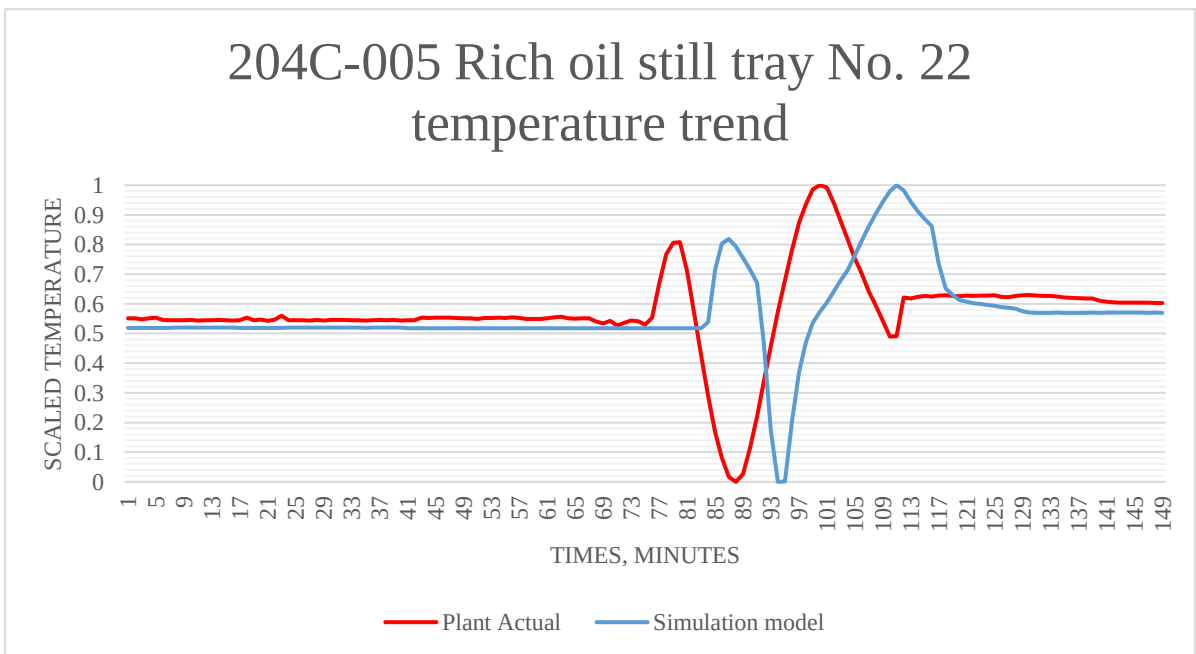


**Figure 35 Rich oil still tray No. 22 temperature trend – Aspen HYSYS
Dynamic model vs Actual**

4.3.2.3. Depropanizer column

Key crucial parameters for modelling and operation of 204C-006 Depropanizer column are given below:

- Tray no. 28 temperature
- Tray no. 35 temperature

A feed flowrate disturbance from the plant operational historical trend of the cold feed to the 204C-006 Depropanizer column was identified, and a similar disturbance was simulated in the dynamic model in order to compare dynamic the behaviors.

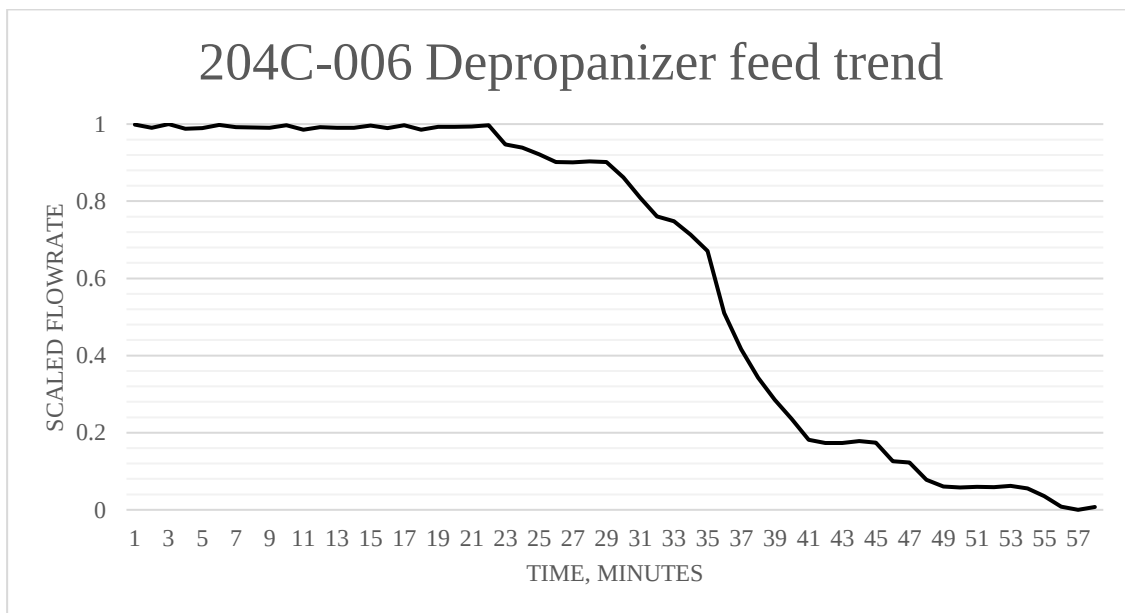


Figure 36 Depropanizer feed disturbance - Data taken from plant historical trends

It was observed from the historical trend of the plant and confirmed by plant operation personnel that many of the crucial control loops

of the unit are usually kept on manual control. This inevitably means that the performance of the unit reflected by the historical dynamic trends depends primarily on the performance of the specific plant operator (technical ability, personal judgement and decision-making). Accordingly, to validate the performance of the unit, the open loop response is better to be compared to that predicted by the simulation model.

Figure 37 and figure 38 depict the corresponding dynamic behavior of the column for different tray temperatures.

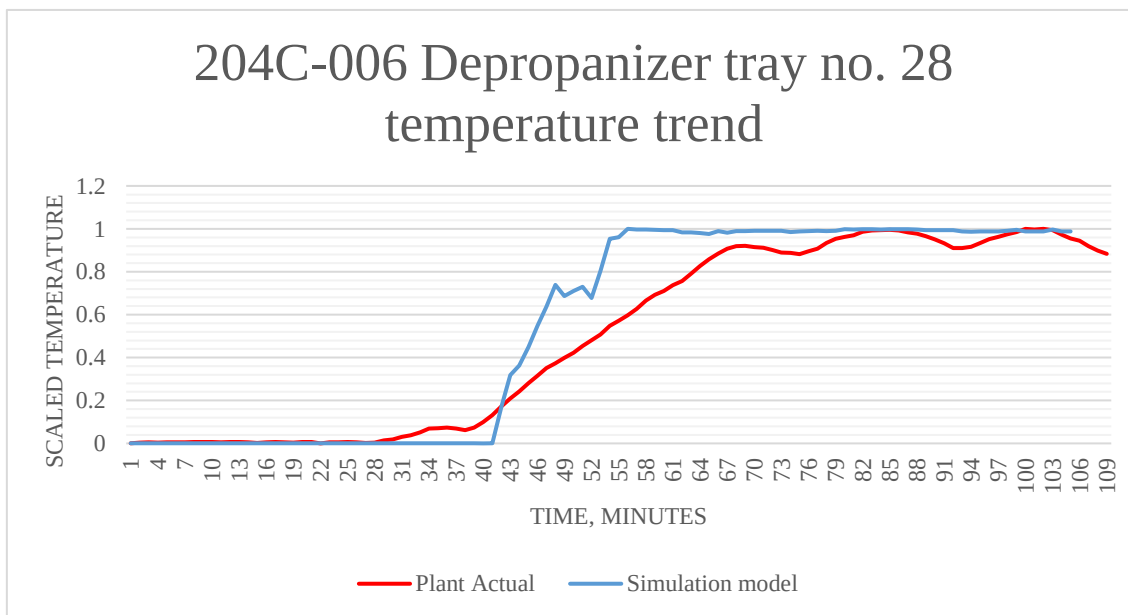


Figure 37 Depropanizer tray No. 28 temperature trend – Aspen HYSYS Dynamic model vs Actual

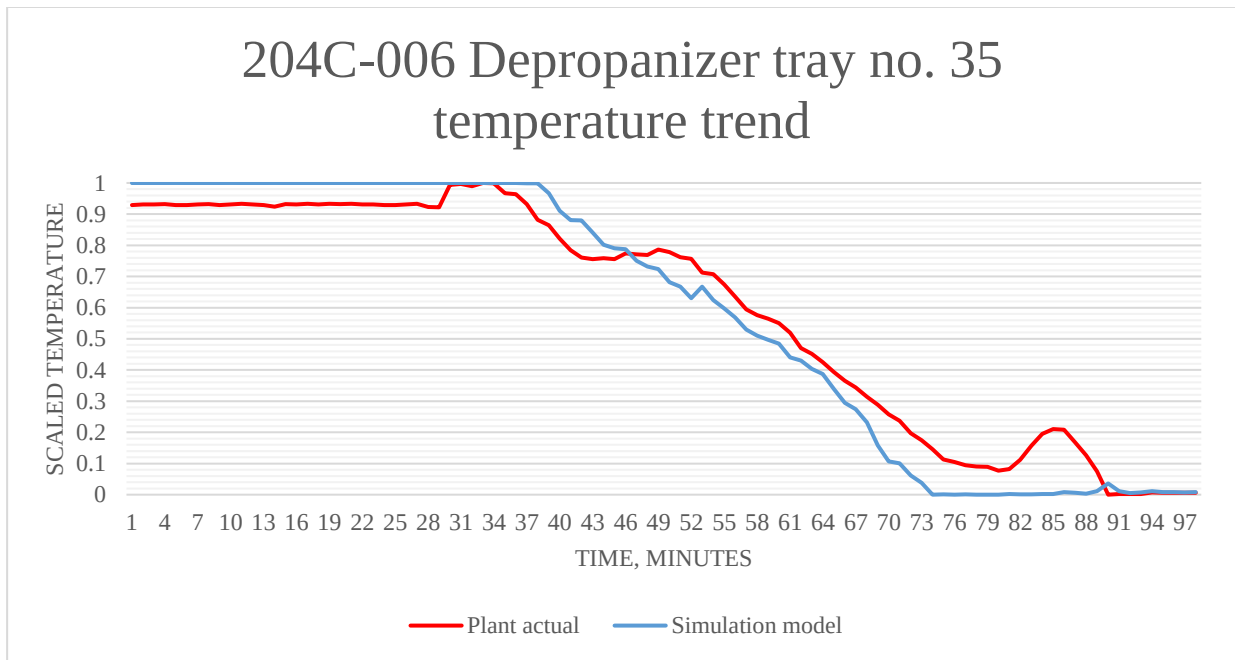


Figure 38 Depropanizer tray no. 35 temperature trend – Aspen HYSYS Dynamic model vs Actual

4.3.2.4. Debutanizer column

Key crucial parameters for modelling and operation of 204C-007 Debutanizer column are given below:

- Tray no. 10 temperature
- Tray no. 37 temperature

A feed flowrate disturbance from the plant operational historical trend of the feed to the 204C-007 Debutanizer column was identified, and a similar disturbance was simulated in the dynamic model in order to compare dynamic the behaviors.

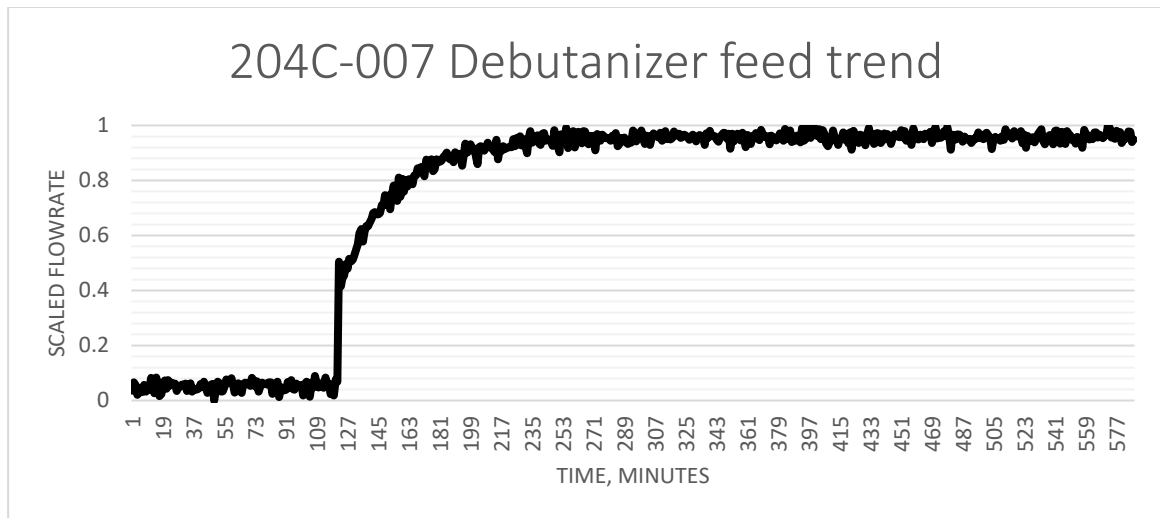


Figure 39 Debutanizer feed disturbance - Data taken from plant historical trends

Similar to the Depropanizer column, a validation of the open loop responses shall be done due to the same reasons cited above.

Figure 40 and figure 41 depict the corresponding dynamic behavior of the column for different tray temperatures

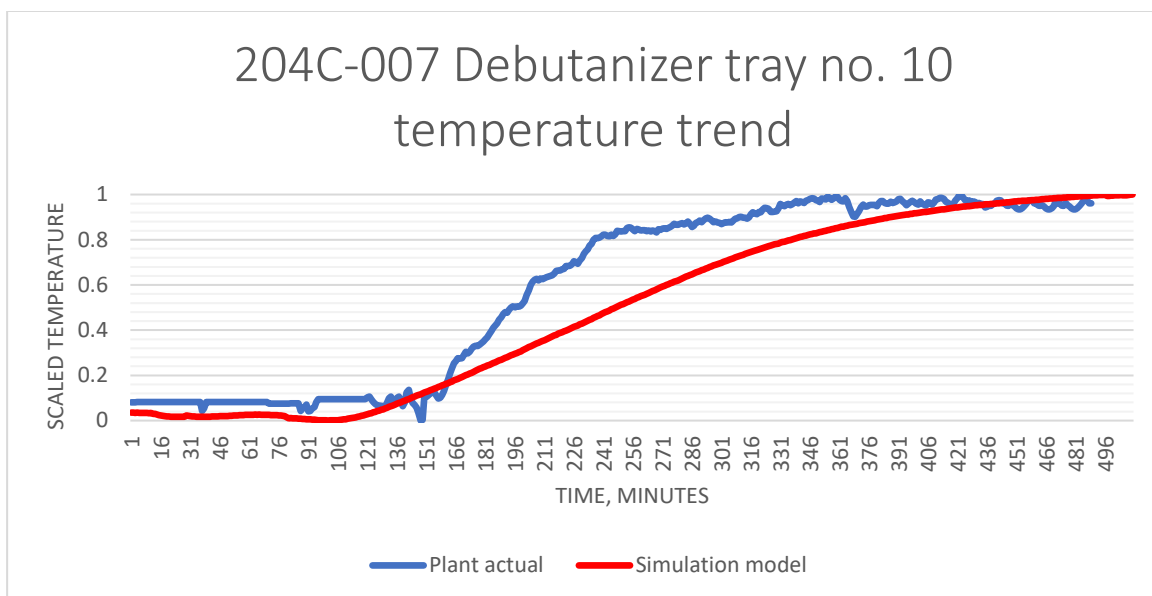


Figure 40 Debutanizer tray No. 10 temperature trend – Aspen HYSYS Dynamic model vs Actual

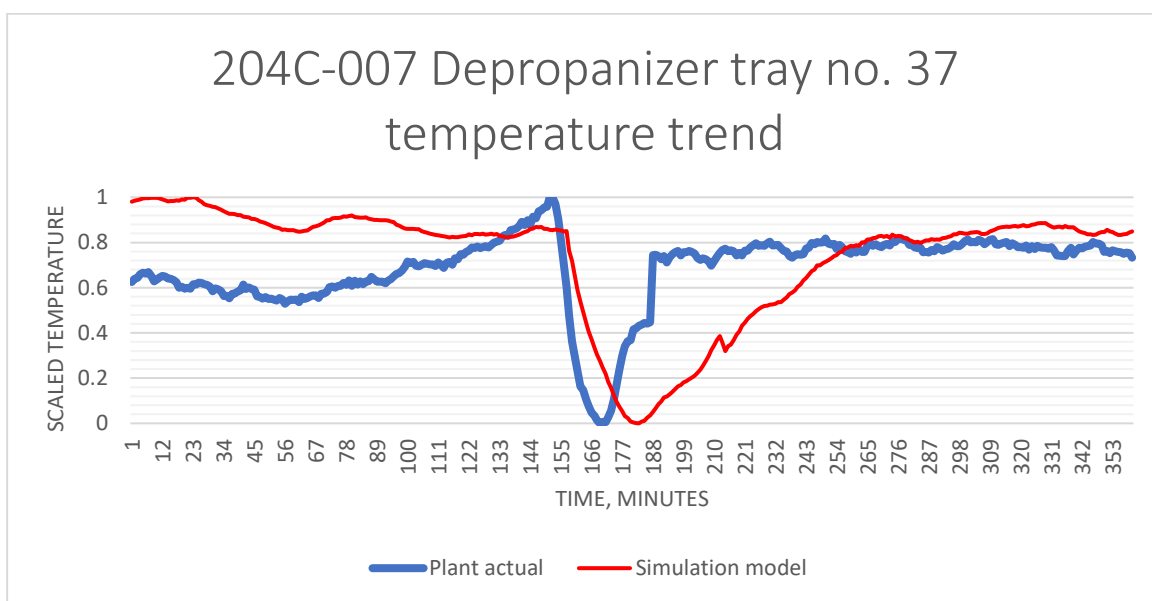


Figure 41 Debutanizer tray No. 10 temperature trend – Aspen HYSYS Dynamic model vs Actual

A cross verification with actual plant records for the disturbance scenario show that the dynamic model developed exhibits a good representation of the dynamic behavior of the plant.

4.4. Functional design of MPC

The main driving force behind the deployment of MPC in oil and gas industries is to maximize potential economic benefits. Functional design may be defined as the proper planning and design of MPC controller to meet the operational and economic objectives of the plant. It is worth noting that there is no universally accepted procedure to do a functional design. Much of the design comes from properly understanding the process under consideration and understanding the hidden economical critical points that can enable maximization of profitability. Rigorous process, control and engineering knowledge together with experience facilitate successful implementation of MPC systems in general.

However, generally the following serve as guidelines in the functional design stage

1. Defining of objectives – Economical and process control objectives.
2. Identifying constraints – Product quality specification, operational safety limitations and equipment limitations.
3. Defining the control system scope – How much the system should be included in the MPC scope.
4. Regulatory control issues – Identification and assessing of basic regulatory control performance.
5. Optimization opportunity – Areas of the process unit where opportunities for economical optimization exists.

Usually, all other steps in the MPC formulation are well defined except for the functional design, which primarily depends on the experience and the depth of knowledge of the plant's engineers. Different applications of

MPCs applied to the same process can yield different results even though the process is identical. This is due to the different methodologies employed in the functional design stage of the MPC.

(Lahiri, 2017)

Analyzing the process plant under consideration is vital to identify the main areas of the process where economic opportunities exist in order to maximize benefits. The followings serve as guidelines in deciding which areas of the process shall be –considered for MPC application.

1. Areas where there is scope to improve economics by controlling a specific process unit.
2. A particular process unit that affects economic objective of the plant and its associated constraints.
3. Include enough of the process in order to cover the significant constraints that are considered important for the economical operation of the process.
4. Generally, – end storage vessels, programmable logic controller operated vessels such as pressure and temperature swing adsorption units are excluded from the MPC scope as well as spots where little process variation is expected.

The identification of the objective functions is another important aspect in the MPC functional design that needs to be carefully examined. There is no standard approach that can be followed and understanding the unit makes profit within different ranges of constraints is essential.

Since the MPC is essentially a control optimization, the followings serve as guidelines in order to identify the objective function of the corresponding problem:

1. Economic objectives

- (a) Understanding of how the plant makes profit.
- (b) The main types of feeds and products involved in the unit.
- (c) Understanding how profit margin can be maximized which includes minimization of product giveaways and increasing of the more valuable product.

2. Operational objective

- (a) Achieve and exceed daily production target.
- (b) Minimize process variation.

Another important phase is the identification of the different constraints involved. However, it is imperative that the constraints considered are properly studied since generally, the most economical operating point lies close to the constraint envelope. The following can serve as guidelines in identifying the constraints –

1. Process limitation

- (a) Such as heating and cooling duties, flooding in distillation columns etc.
- (b) Process overloading in process equipment.

2. Product quality limitation

- (a) Minimum product quality specification.
- (b) Maximum impurity specification.

Selection of the MVs and the CVs should be done carefully. As with the constraints, such variables are usually known, however, a deep knowledge of the process is required in order properly understand the interaction between the different MVs and CVs. To further explore the interaction between identified MVs and CVs, a step test is employed. Generally, the following serve as guidelines in selecting the MVs and CVs

1. Manipulated variables MVs

- (a) They are usually the regulatory level (i.e., DCS) setpoint of the different multi-loop controllers.
- (b) Some MVs are obvious such as commonly moved or changed controller setpoints.
- (c) Other MVs are not so obvious such as setpoints that rarely change. However, such MVs should also be considered.

2. Controlled variables CVs

- (a) Parameters that improve plant unit stability and profitability.
- (b) Parameters that represent operational limit.
- (c) Ensure that each CV is dependent on a minimum of one MV.
- (d) The CV can be measured or inferred/estimated.

4.4.1. MPC functional design for the Deethanizer unit

The primary function of the Deethanizer is to strip off all ethane and lighter components from the feed in preparation for the fractionation train. The feed entering contains around 7 mol% of ethane that needs to be removed in order to meet the maximum impurity specification of 2 vol% in the final propane product. The performance of this column depends heavily on the quantity and quality of the feed gas entering the plant and, on the operator, handling the unit. As been mentioned earlier, there are two sources of gas entering the plant. A significant part of the feed gas comes as associated gas from the Bahrain oil fields. This gas is usually stable in terms of its composition but is considered lean in terms of heavy components. Another important source of gas is the ROG that comes from a nearby facility. The ROG is the waste / vent gases from different units in the refinery and its composition is affected heavily by the performances of these different units. In terms of gas quantity, the ROG is much lesser when compared to the associated gas; however, it is much richer in terms of the heavy components. The number of liquid products extracted per MMSCFD of ROG is around 10 times more than that extracted from the associated gas. It is worth noting that most of the NGL facilities that were originally designed and commissioned in 1979 did not consider ROG as a possible feed. As a result, the performance of Deethanizer is heavily influenced by the quality and quantity of the ROG.

The Deethanizer column is controlled primarily by the temperature profile. Tray no. 9 is an effective control point for the tower operation. However, there is no direct control over this tray. Control is mainly

achieved by controlling the bottom temperature via hot oil flow to the reboiler.

The following were chosen as MVs and CVs after carefully analyzing the process.

Table 6 Manipulated and controlled variables - Deethanizer unit

Manipulated variables	Controlled variables	Remarks
1. QIC DEC2 REBOILER (Deethanizer reboiler)	1. C2 Vol% in final propane product	• The column directly controls the amount of ethane slippage in the bottom product. An appropriate conversion factor has been used to calculate the C2 vol% in the final propane product • The TIC DEC2 REBOILER directly controls the amount of heat that goes into the column which affects the stripping of ethane from the bottom product
2. LIC-036		• From step test conducted, it was found that controlling the bottom flowrate affects ethane stripping from the bottom product as it alters the

		residence time of the liquid in the reboiler. • For safety purposes, the MPC directs control over the bottom flowrate through LIC-036 and not directly controlling the valve since this would pose a problem incase MPC fails.
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The control structure for the MPC on the Deethanizer unit is given in figure 42. The corresponding MPC model developed for the Deethanizer is given in figure 43.

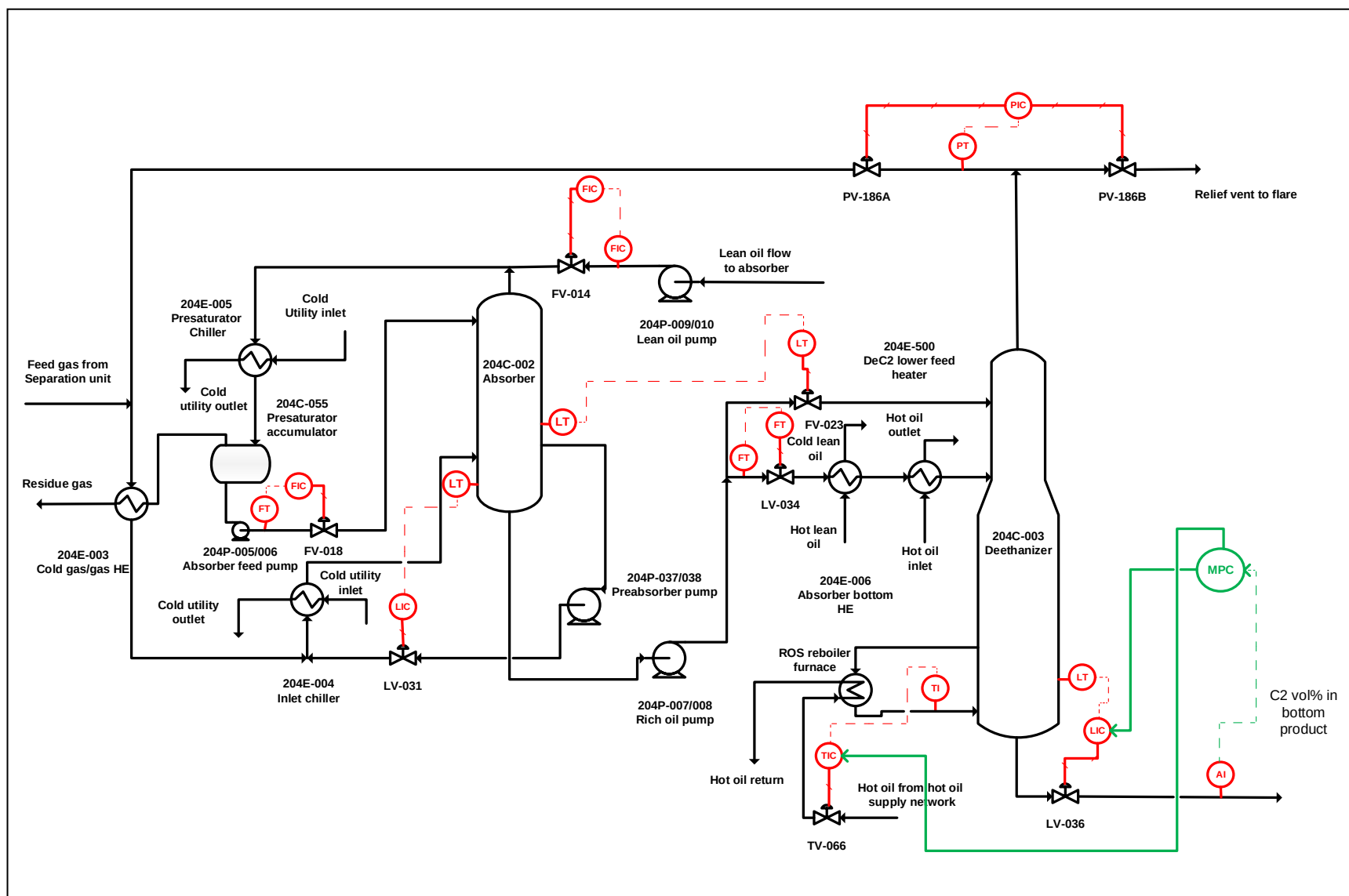


Figure 42 Deethanizer unit - MPC structure

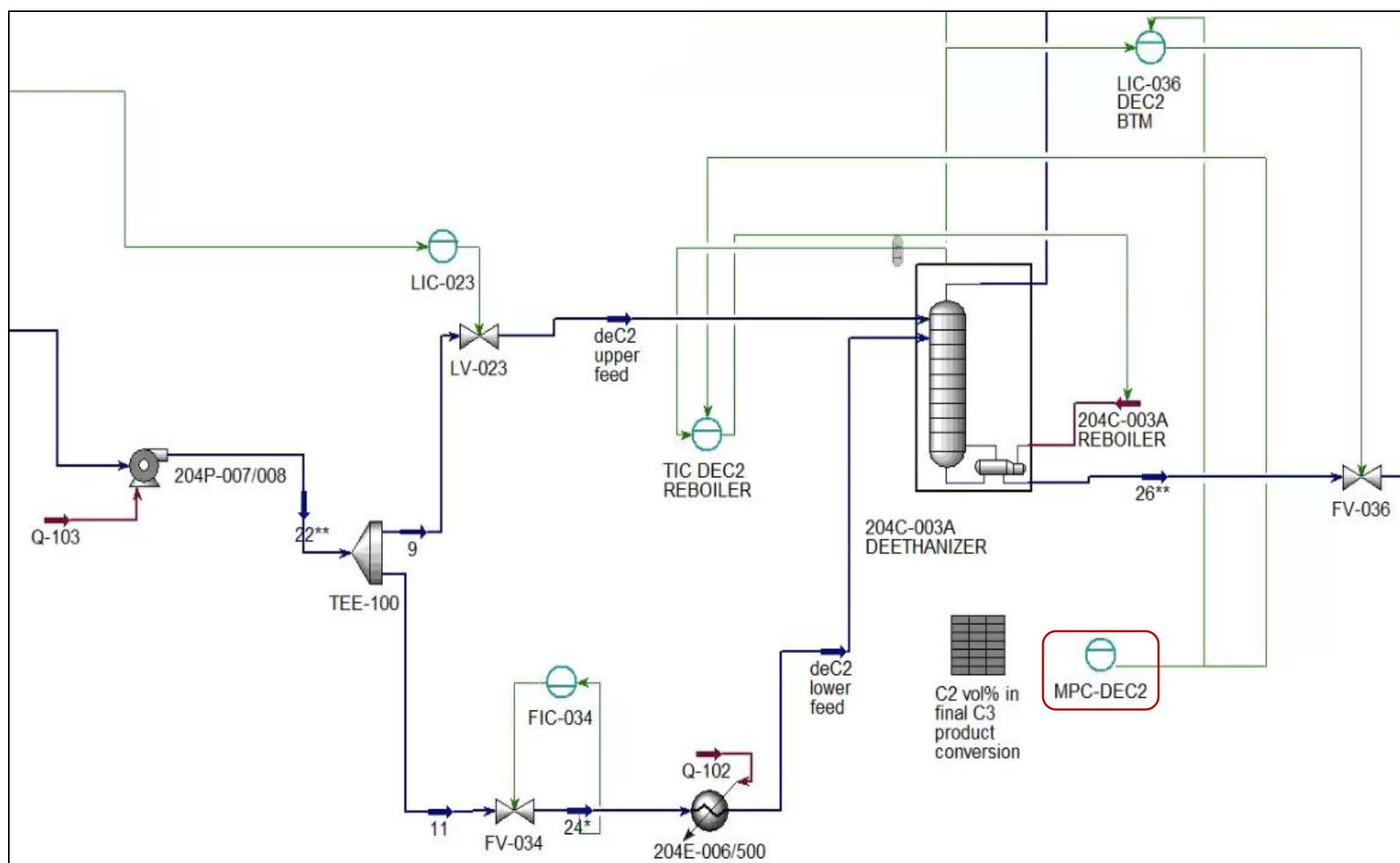


Figure 43 MPC model developed in Aspen HYSYS Dynamics - Deethanizer unit

The objective function for the Deethanizer aims to maximize the ethane slippage in the bottom product in favor of the Depropanizer economics. That is, to achieve a maximum product specification of 2 vol% of ethane in the final propane product as far as possible, since any reduction from this level of the ethane is deemed unnecessary. The manipulated variables (i.e., decision variables in the optimization problem) are the setpoint to the TIC DEC2 Reboiler controller, which will control the heat input to the Deethanizer column, and the setpoint LIC-036 which will control indirectly the amount of bottom product leaving the column. As mentioned in Chapter 2, the MPC control system requires a model in order to predict the future behavior of these different controlled variables. The model employed in this case has been derived through step testing and model identification from which a prediction model of the system is identified. The step test data for the system can be found in Appendix A-4.

An important aspect of the MPC development is the tuning parameters. However, it is to note that, even though the tuning parameters affect largely the performance of the system, it also depends heavily on the developed model. Without a good model, the MPC system will not be able to contribute positively. Without a good model, the MPC will not be able to contribute positively (Forbes et al., 2015).

The tuning parameters for the Deethanizer MPC system is given in Appendix – A5.

4.4.2. MPC functional design for the Depropanizer unit

The main function of the Depropanizer unit is to extract propane from the mixture of liquid coming from the Rich oil still overhead product. The control of this column is via TIC-115, which controls tray no. 35 temperature through a cascaded controller FIC-052 that adjusts the amount of reflux entering the column. Tray no. 35 is a very effective control point that determines the top product impurity. Another important controller is the 204C-006 Reboiler controller that adjusts the amount of heat entering the column. The specification of the products from the Depropanizer columns are propane (95 vol% min), Ethane (2 vol% max) and Butane (4 vol% max).

There are two feed streams entering this column. The first feed (cold feed) contains the liquid products extracted from the gas feed entering the plant and reach the Depropanizer after the Absorption, Deethanizer and Rich oil still unit. The second stream (hot feed) contains mostly gas condensates, that is, hydrocarbon liquids that are present as gaseous components in the natural gas entering the plant. The gas condensates are mostly obtained through liquid knock offs following the compression of the gas in the compressor stations located around the Bahrain oil fields. In fact, the hot feed does not differ significantly in quantity and quality. However, the cold feed is characterized by its high variations in both, quantity, and quality. The cold feed is very sensitive to changes especially in the ROG operation from the nearby refinery. Ethane is considered as an impurity and is not sold as one of the products from the NGL plant. Instead, it either leaves the plant as part of a large residue gas stream, which is sent back for gas reinjection in the reservoir to maintain the wells' pressure or sent as fuel gas to different utility users. Both propane and butane are products from the plants, where butane is considered more valuable

(higher market price). As such, there is incentive in order to maximize ethane slippage into propane product while maintaining product specification. Even though as per specification, propane has margin of 4 vol% to accommodate butane, there is no motive in maintaining such margins since butane is considered a more valuable product.

Hence, the objective function of the MPC should aim to minimize butane content (by controlling propane product purity) in the overhead propane product while maximization of propane slippage in the bottom product while maintaining all product specification. Again, the prediction model required for the MPC system has been identified by performing a step testing on relevant variables that are deemed significant for the operation of the Depropanizer (see Table 7). The details concerning the step testing data for the Depropanizer unit can be found in Appendix A-4. The MPC tuning parameters for this unit can be found in Appendix A-5.

After analyzing the unit, the following have been chosen as MVs and CVs:

Table 7 Manipulated and controlled variables: Depropanizer unit

Manipulated variables	Controlled variables	Remarks
1. QIC DEC3 Reboiler (Depropanizer reboiler)	1. C3 vol% in final propane product	1. From the step testing done to determine the extent of interactions between variables, we have found both MVs have

2. TIC-115	2. C3 vol% slippage in final butane product.	significant impact on the C3 vol% in final product. 2. QIC DEC3 reboiler has a significant influence on the amount of C3 slippage in the bottom stream. 3. TIC-115 mainly control the purity of the final propane product. 4. An appropriate conversion factor has been used to calculate the C3 vol% in the final butane product.
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The control structure for the MPC on the Depropanizer unit is given in figure 44 and the corresponding MPC model developed for the unit is given in figure 45.

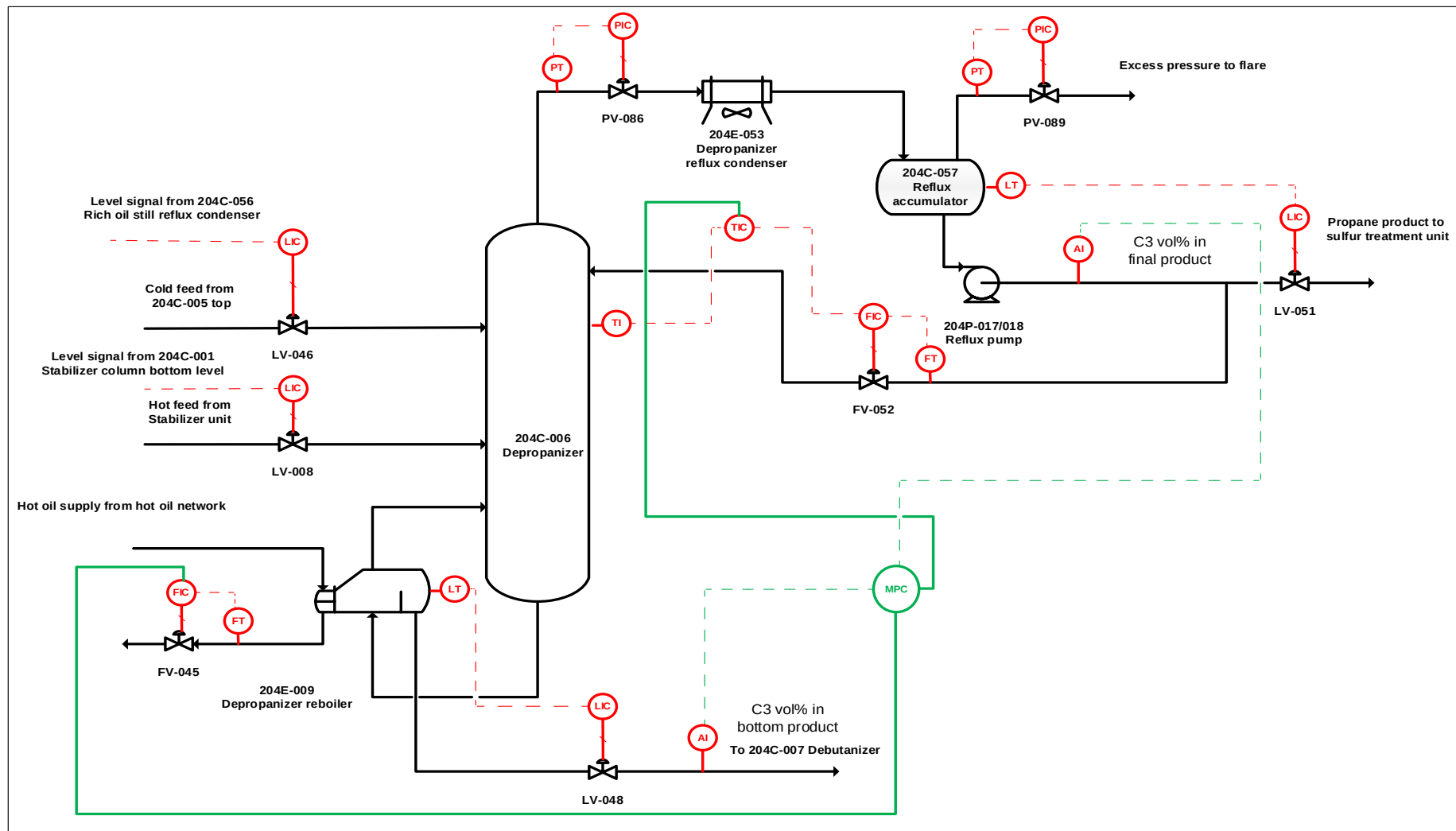


Figure 44 Depropanizer unit – MPC structure

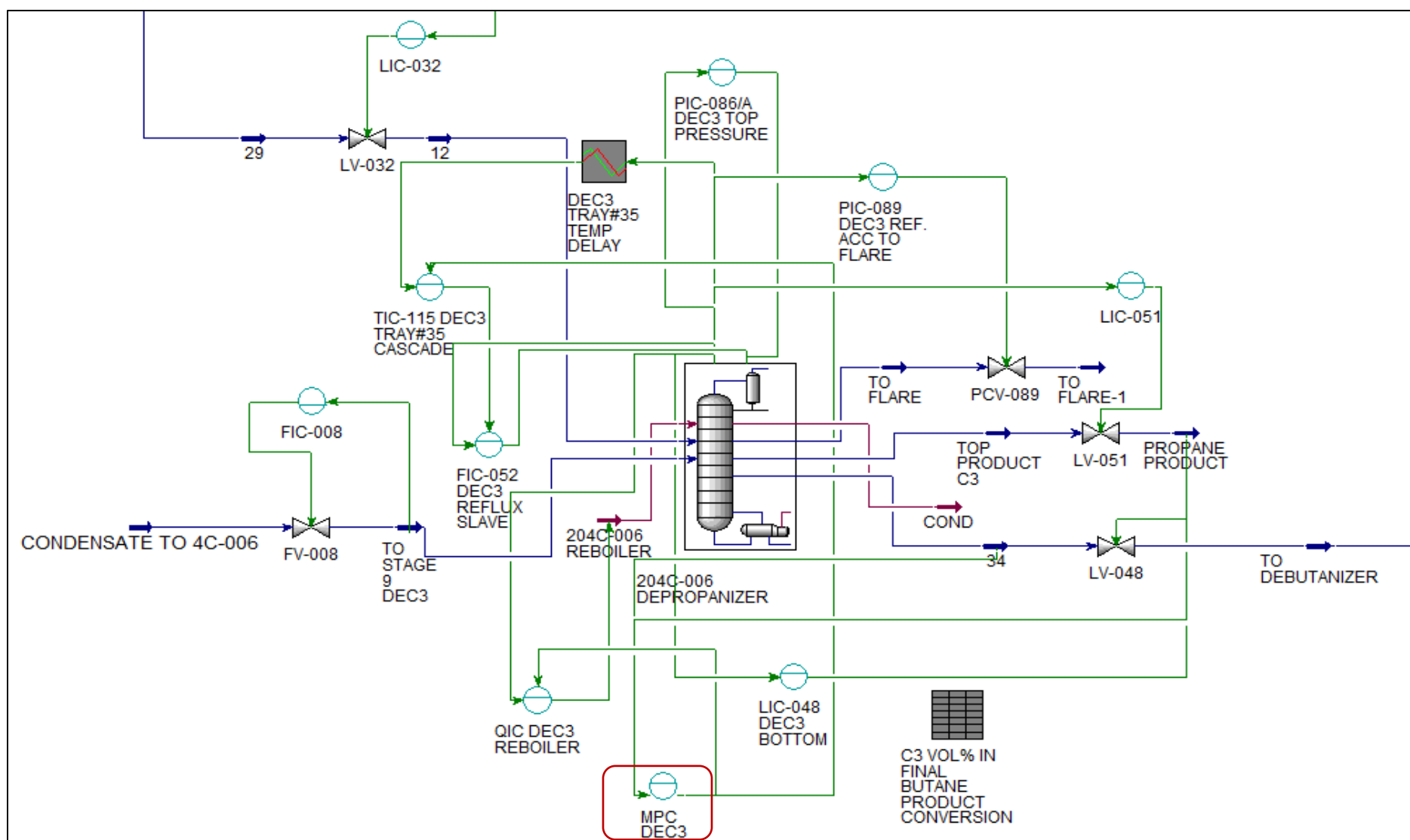


Figure 45 MPC model developed in Aspen HYSYS Dynamics – Depropanizer unit

4.4.3. MPC functional design for the Debutanizer unit

The Debutanizer unit is the final unit in the fractionation train and serves to extract butane from naphtha, and hence is the only unit in the plant that has quality specification for both the top and bottom product. The feed to this unit primarily comes from the bottom stream of the Depropanizer unit. Control is rather unusual for this column. The quality of the top butane product is mainly controlled by controlling temperature of tray no. 37 via TIC-128 that adjusts the amount of reflux returning to the column. The control of Naphtha RVP product specification is mainly done by controlling the hot oil flow to the reboiler and hence the amount of heat entering the column.

It has been observed that the plant's operators usually maintain the naphtha RVP away from the maximum specification. The plant's operator behavior is attributed to the relatively large variation resulted from the current implemented control system. Therefore, this way of operator's reaction results in a lost opportunity. That is, since naphtha is sold at a higher price compared to butane, there is incentive to increase the RVP of the bottom naphtha product (i.e., maximize butane slippage into the bottom naphtha product) as long as the naphtha's RVP specification is met.

Noticeably, the operation of the Debutanizer is heavily affected by the quality of the feed gas, in particular, the quality of the ROG. Heavier ROG inevitably means that more naphtha and butane content is present in the feed and hence attention should be given to maintain the operation of the unit in spite of this variation in order to achieve an economical

operation. However, as mentioned earlier, it has been found that operators will generally follow this conservative path to keep the column away from the quality limits. It has also been found that the Naphtha RVP specification is very sensitive to the reboiler controller, which is evident from the large fluctuations in the final RVP specification. The specifications for Butane and Naphtha for the Debutanizer are, 97 vol% min and 12 psia RVP max, respectively.

After carefully analyzing the unit, the following have been chosen as MVs and CVs:

Table 8 Manipulated and controlled variables - Debutanizer unit

Manipulated variables	Controlled variables	Remarks
1. QIC DEC4 Reboiler (Debutanizer reboiler)	1. C4 vol% in final butane product	1. TIC-128 heavily influences the butane purity while QIC DEC4 reboiler controls mainly the RVP in the Naphtha product.
2. TIC-128	2. RVP in final Naphtha product	2. Both TIC-128 and QIC DEC4 reboiler influence naphtha RVP and butane purity respectively and hence represents an interaction.

The objective function for the Debutanizer should maximize butane slippage into the Naphtha product while maintaining the aforementioned

butane and naphtha products quality constraints. The internal prediction model required for the MPC system was obtained by performing a step testing on the variables deemed significant for the operation of the unit. The details concerning the step test can be found in Appendix A-4. The MPC tuning parameters can be found in Appendix A-5. The control structure for the MPC on the Debutanizer unit is given in figure 46 while the MPC control strategy developed for the Debutanizer is given in figure 47.

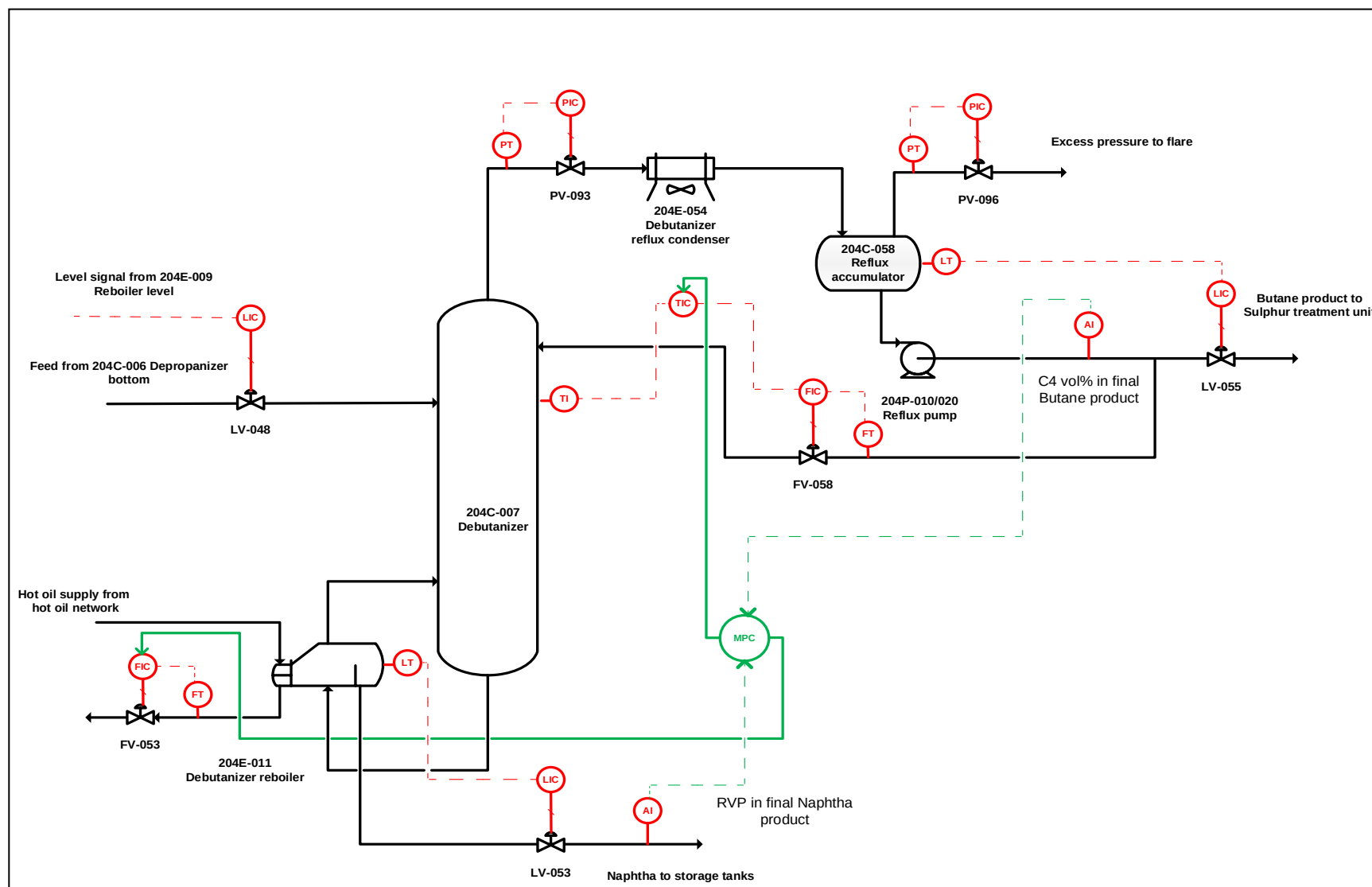


Figure 46 Debutanizer unit – MPC structure

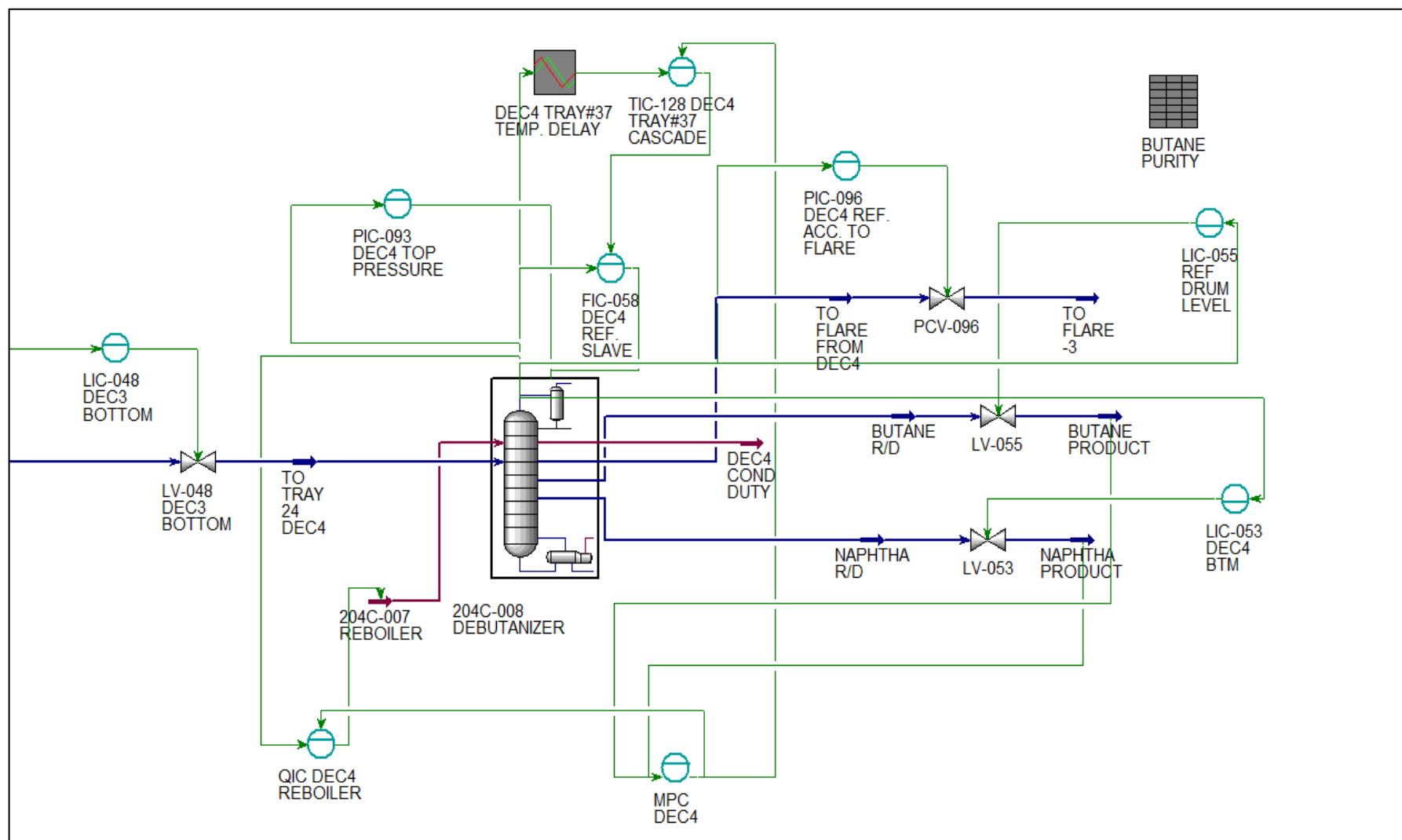


Figure 47 MPC model developed in Aspen HYSYS Dynamics - Debutanizer unit

Chapter 5 - MPC simulation results and analysis

5.1 Deethanizer unit

5.2 Depropanizer unit

5.3 Debutanizer unit

5.4 Conclusions

The results obtained from the simulation after having employed the MPC are presented and discussed in this chapter. For each of the unit presented in section 4.4, different scenarios have been simulated in order to measure how well the MPC can achieve its corresponding objective. The scenarios have been carefully selected to mimic real problems experienced in the NGL plant. The most challenging scenarios faced by the plant operation and where the proposed MPC is expected to provide advantages is the changes in feed gas composition especially in the ROG quality and disturbances in flow rates.

The results of the simulation shall be presented along with the details of the scenario basis. A comparison shall be done with the current existing control structure of the plant in order to demonstrate the benefit of proposed MPC over the current best practice.

5.1. Deethanizer unit

The results obtained from the simulation of the Deethanizer unit under MPC will be presented below. The results include the details obtained from simulating 4 different scenarios – the details and basis for the scenarios shall be stated along, and a comparison between operating the unit *with* and *without* MPC in terms of variance handling capabilities.

5.1.1. Variance handling capability

One of the strongest advantages that was observed in the MPC employment through the performed simulation studies is the efficacy to significantly reduce variations in the controlled variables and hence

shifts the average line of the controlled variable closer to the quality specification limits (constraints). A parameter that was identified as an opportunity for economical maximization is the ethane slippage into the bottom product (feed to the Depropanizer). Ethane is typically an impurity in the product with a maximum specification of 2 vol% in the propane product.

Figure 48 shows the superiority of the MPC in attempting to reduce variations in the ethane slippage to bottom product and hence naturally allowing the operating envelope to be pushed closer to the constraint limit of 2 vol% in the final propane product. The simulations include reasonable allowances in process noise to simulate real plant noise

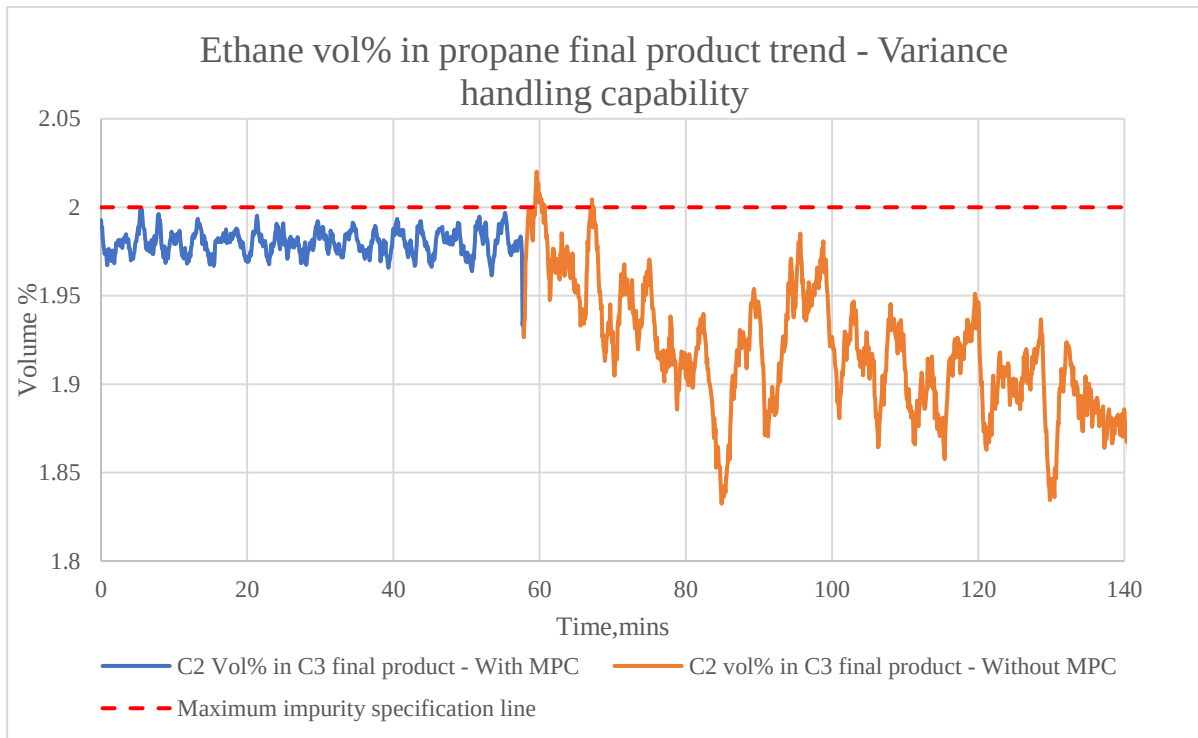


Figure 48 Variance handling capability, Operating closer to constraint limit - MPC vs Plant current control system

conditions. Without the MPC, the simulations show that the Deethanizer unit can achieve an average overall ethane content of 1.90 vol% in the final product with a variance of 0.0351. However, operating the column under MPC reduced the variance by 96.62% and allowed closer operation to the maximum specification line with an overall average ethane content in final propane product of 1.98 vol%. The superiority of the MPC in reducing the variation is largely because of the employment of model in the background that predicts future behavior while taking variables interaction into consideration, hence optimizes the reboiler duty and bottom draw rates, which are the manipulated variables. It does so by minimizing an objective that seeks to reduce the difference between the target output and the predicted future output and with least possible control efforts to reach the desired

state. The MPC manages to improve the performance of the system by predicting how the system will behave after a certain prediction horizon.

5.1.2. Feed increase scenario

A scenario has been simulated whereby feed to the unit has been increased by 10% to evaluate the capability of the MPC in adjusting the reboiler and bottom column draw rates which are the manipulated variables as to effectively maintain the ethane slippage as close as possible to the specified constraint limit.

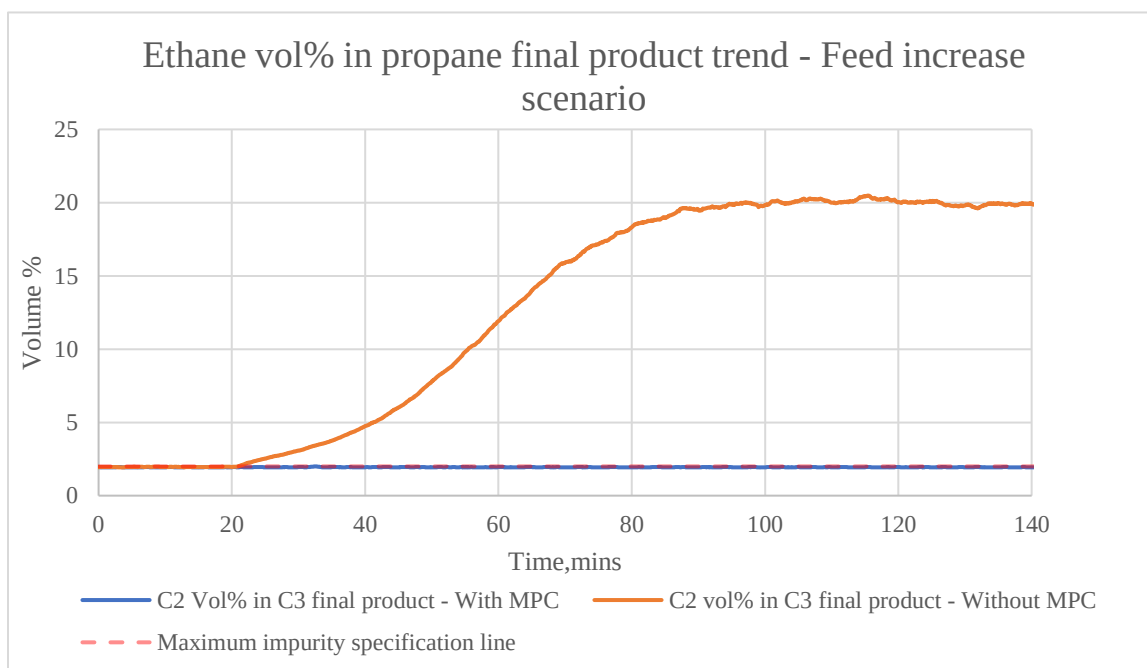


Figure 49 Deethanizer unit performance comparison - Deethanizer feed increase scenario

As can be seen above, the Deethanizer unit under MPC manages to maintain maximum slippage of ethane into the bottom product while respecting constraint limits by effectively navigating the reboiler duty and bottom product draw-rate. The MPC achieves this by optimizing

the reboiler duty and bottom draw by utilizing the prediction model in the background in order to decide an optimal course of action. The Deethanizer unit under MPC manages to achieve an overall ethane content of 1.94 vol% in the final propane product compared to a final steady state of 20 vol% ethane in case of operation without MPC thereby resulting in significant off specification production. Increasing the feed to the column inevitably increases the total hydraulic load in terms of total ethane content in the column. This means that more reboiler duty and potentially more liquid residence time will be required to order to maintain performance. The MPC successfully manages to calculate new set points regularly for both the reboiler and liquid level controller that controls that bottom draw rate indirectly.

It is important to note that this comparison is done in order to present the superiority of MPC in reacting to feed increases compared to the case of no MPC.

5.1.3. Feed decrease scenario

A scenario involving a feed decrease has been simulated. The feed has been decreased by around 10% to evaluate the MPC system capability in attempting to control the reboiler heat duty and bottom draw rate in order to maintain maximum ethane slippage while respecting constraint limits.

As can be seen in figure 50, as with the previous scenario, the Deethanizer unit under MPC successfully manages to maximize slippage of ethane into the bottom product while respecting constraints by effectively controlling the reboiler duty and bottom draw rate. The unit under MPC manages to maintain an overall ethane content of 1.96% vol in the propane product while the without MPC the slippage smooths out to almost nil in the final propane product which represents lost opportunity.

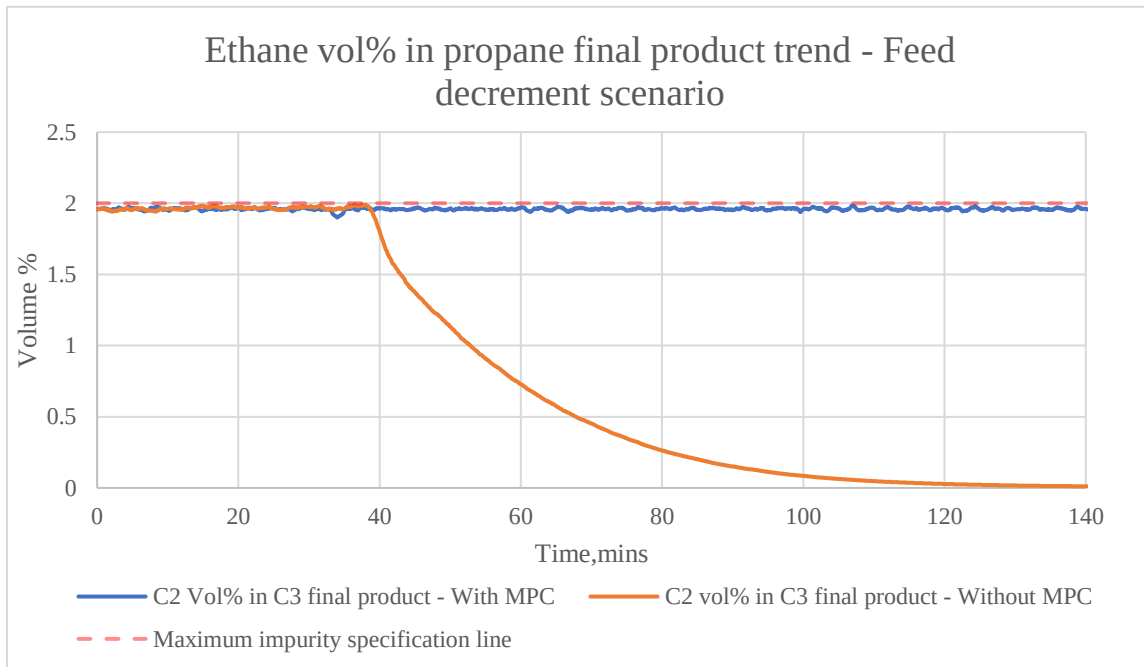


Figure 50 Deethanizer unit performance comparison- Feed decrease scenario

Decreasing the feed into the column inevitably means that hydraulically, total ethane content in the column is lesser and hence effectively more than required heat and residence time exists at the current condition, which results in stripping of almost the entire ethane product. This is evident in the case of Deethanizer unit operating without MPC.

5.1.4. Feed composition change – Increase of ethane content in feed

One of the most important scenarios identified from the plants operating history is the feed gas quality changes especially in the ROG quality coming from a nearby refinery. The quality of the ROG is very dependent on the operational performance of several key upstream units in the refinery. Therefore, upsets in any of these key units in the refinery will result in propagation of these upsets to the NGL under study. The ROG from the refinery is generally quite heavy in composition with significant amount of butane and naphtha content. It has been observed that in situations where for any reason there is a reduction or complete cut of the ROG, the ethane composition in the feed gas increases, which in turn will pose operational challenges to the unit. The Deethanizer unit is very susceptible to changes in the feed gas composition and especially the composition of ethane in the feed. The plant has been originally designed in 1979 for gas compositions very different from what is being processed with currently. All the design and operating parameters have been developed taking into consideration gas conditions in 1979. Through time, the plant facility engineering and operation personnel have developed a way in tacking these gas changes, which largely depends on the experience of the control man who is directly controlling the unit. This makes the entire operating process, hence, the plant performance subject to the experience of the control man, which in the absence of it; it may lead to economic losses that could have been otherwise avoided.

An increase in the ethane content from 6.38 to 9.97 vol% in the feed has been simulated in order to evaluate the capability of the MPC to handle changes in the feed composition. The basis for the incremental change has

been deduced after reviewing the plant's operating data historian to reflect some realism on the considered scenario.

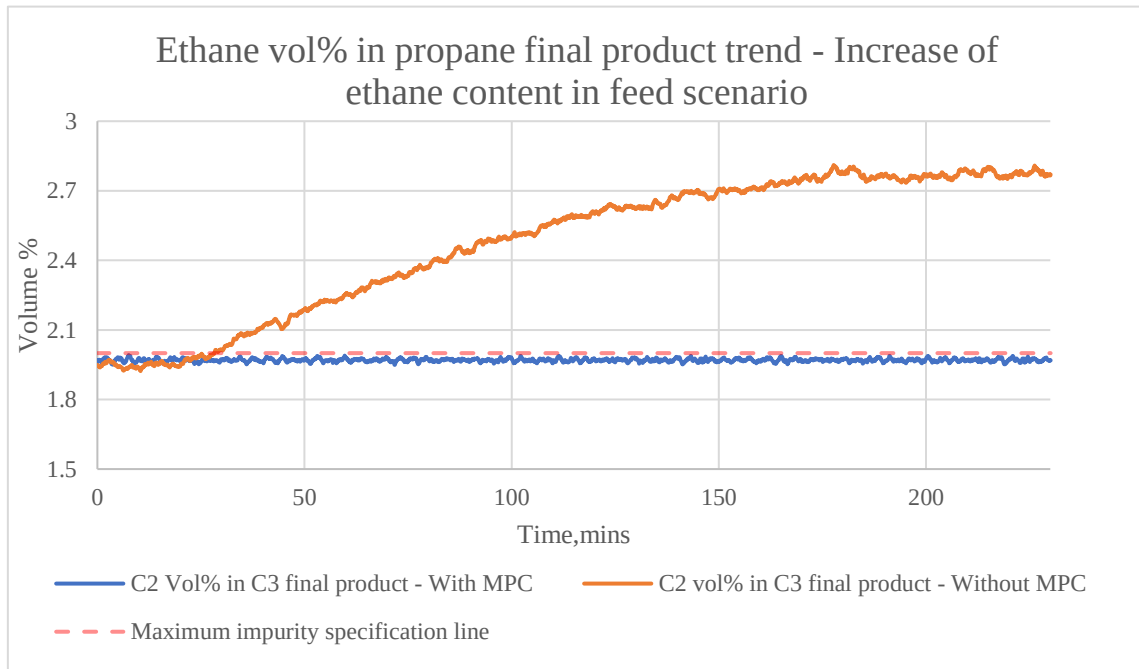


Figure 51 Deethanizer unit performance comparison – Increase of ethane content in feed scenario

As can be seen from the above figure, the unit under MPC control effectively manages to maximize the ethane slippage content in the final propane product while respecting specification constraints.

The unit under MPC manages to achieve an overall ethane content of 1.96 vol % while operating the unit without MPC results in a steady state ethane content of around 2.46 vol% in the final propane product.

5.1.5. Feed composition change – Decrease of ethane content in feed

Similarly, another scenario equally important has been simulated where decrease in ethane composition in the feed is occurring. This usually happens when the quantity of the ROG from the refinery increases or if the ROG composition becomes more heavy due to operational upsets in the refinery which will require a revising the operating profile of the unit.

A decrease in the ethane composition from 6.38 vol% to 3.11 vol% has been simulated.

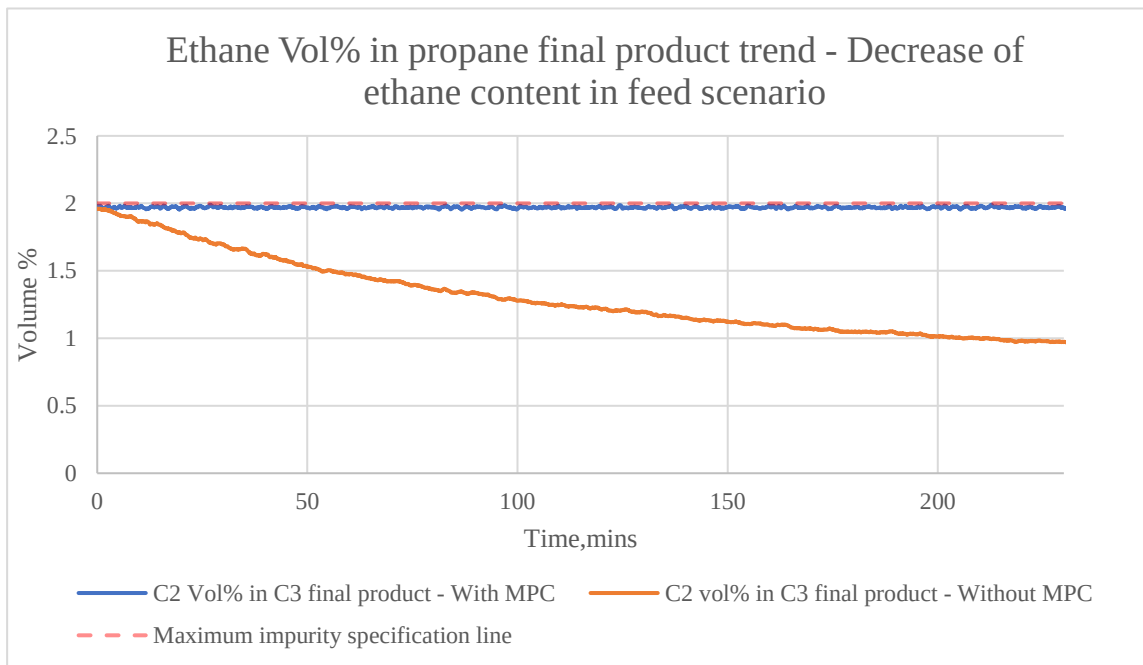


Figure 52 Deethanizer unit performance comparison – Decrease of ethane content in feed scenario

As can be seen from the above, the unit under MPC successfully manages to maintain the ethane content in the final propane product at the most economical point which is as close as possible to the maximum

allowed concentration. The MPC manages to achieve an overall ethane content of around 1.97 vol% compared to a final steady state of around 0.93 vol% without MPC. Decreasing the ethane slippage into the propane product does not result in any off-specification production but results in a lost opportunity for profit maximization.

5.2. Depropanizer unit

The results obtained from the simulation of the Depropanizer unit under MPC shall be presented below. The results include the details obtained from simulating different scenarios – the details and basis for the scenarios shall be stated along together with a comparison between operating the unit with and without the MPC.

As mentioned in earlier chapters, the objective of the Depropanizer unit is to extract propane from the feed, which primarily comes from two units – The Rich Oil Still unit and the gas condensate stabilization unit, and to purify the product to market acceptable specification.

An economic opportunity is identified by maximizing ethane content slippage and minimizing butane content in the final propane product. The basis for this lies in the fact that ethane is not sold as part of the products and does not hold any monetary value for the company while butane is sold at a higher price than propane. Based on this, the optimum target composition identified for the propane product is 98 vol% propane and 2 vol% ethane.

The propane product is extracted as a top product. The main control points for the Depropanizer unit are TIC-115 that controls temperature of tray no. 35 and QIC DEC3 reboiler that controls the heat input to the column. It has been observed that the top product purity is not very sensitive to controller operations of the reboiler, which constitutes an important control point in the unit.

Through analysis of the plant operator records, it has been observed that there exists an opportunity for profit maximization in terms of reducing product giveaway especially in the case when the ROG from the refinery is predominantly changing in composition with respect to butane content.

5.2.1. Variance handling capability – Propane content in final butane product

An important advantage that can be attributed to the MPC is its ability to reduce variation in product quality by employing a prediction model within an objective function that determine the best course of action. As mentioned, there is incentive to slip more propane content with the butane product. One of the most efficient way to achieve this is by reducing product quality variation. This would naturally allow the system average to be pushed closer to the specification limit without violating it and hence would present an economical method of operation.

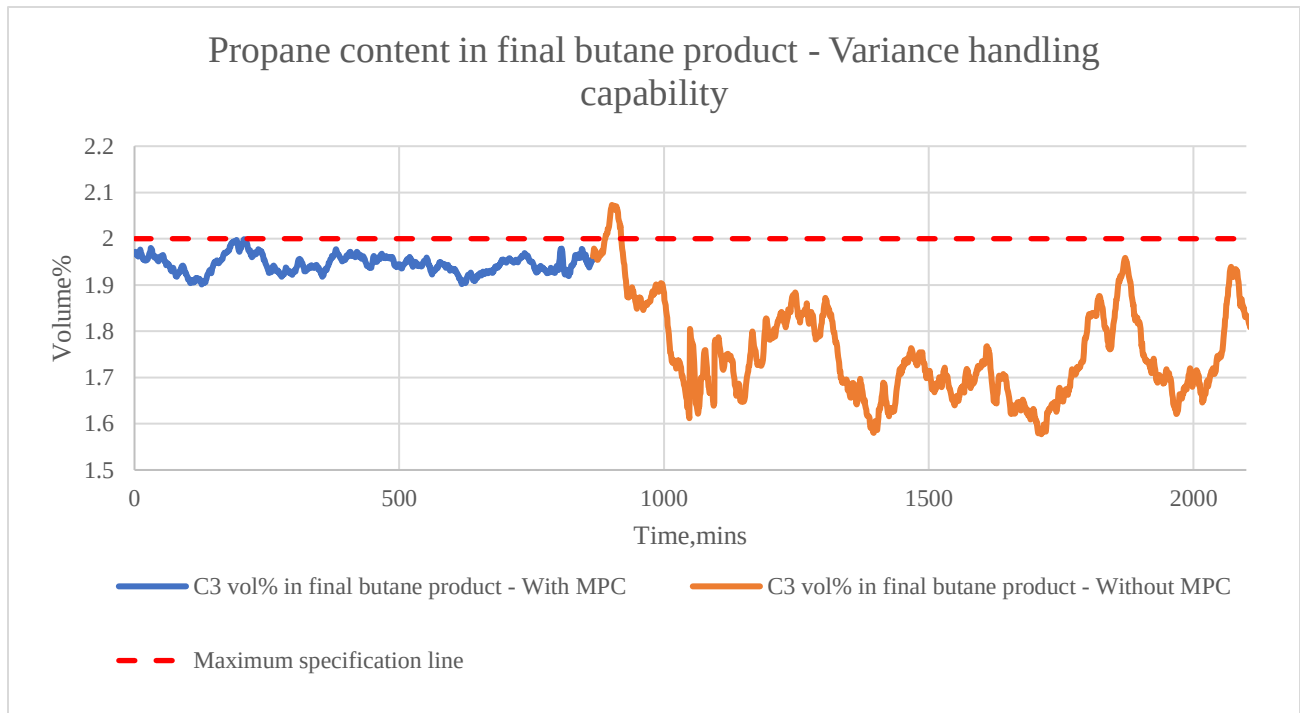


Figure 53 C3 vol% in final butane product – Operating closer to the constraint line. Variance handling capability MPC vs NO MPC

As can be seen from the graph, the MPC performs better in attempting to reduce production variations. The overall average of propane content in final butane product without MPC is 1.75 vol% with a standard deviation of 0.104. However, the unit under MPC manages to increase the overall average to 1.95 vol% with a standard deviation of 0.029. This presents an improved operating point and will have a positive impact on the profit margin.

5.2.2. Feed increase scenario

A scenario has been simulated whereby feed to the unit has been increased by 15% to evaluate the capability of the MPC in adjusting the reboiler and tray no. 35 temperature which are the manipulated variables as to effectively maintain the propane purity as close as

possible to the optimal concentration while at the same time maximizing propane slippage to the butane final product without violating the specification limit.

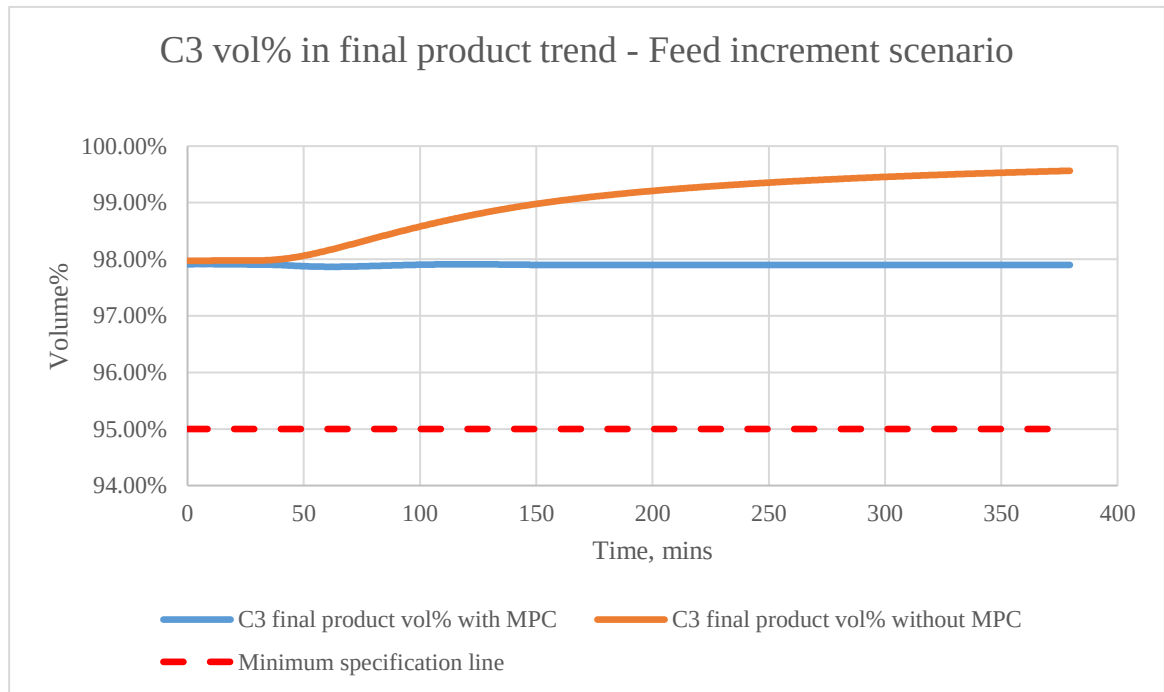


**Figure 54 Depropanizer unit performance comparison
– Feed increase scenario**

Figure 54 shows the result of the simulation scenario. It can be seen that the MPC effectively manages to maintain propane content as close as possible to the optimal 98 vol% purity line by manipulating the reboiler duty and reflux to the column which both have an impact on the target purity. The impact of increasing the feed to the column is of twofold. First, it will increase the overall propane content in the column, which tends to increase the purity of the final propane product and lowers the overall temperature profile. Secondly, it will inevitably result in more propane escaping in the bottom product and hence may result in off

specification butane production without appropriate intervention. Prompt action is required in order to ensure optimal temperature profile of the column. The overall average propane purity from the unit under MPC for the subject scenario is 97.90 vol%. The system manages to achieve this by manipulating both the reboiler and reflux to the column based on the minimization of an objective function that includes the control moves and difference between the predicted and target outputs. The MPC in essence calculates a new optimized operating profile at each interval and hence maintains the propane purity close the specification limit. It should be noted that in all scenarios under the MPC control cases, the effect of the Deethanizer unit MPC attempting to maintain as high as possible ethane content in the final propane product should be taken into consideration while viewing the results. This has been factored in the conversion factors to provide good representation of the system response.

Operating on the unit without the MPC on the other hand results in an increase in product purity to a final state of 99.56 vol%. Though this does not result in off specification production, this represents a loss in profit opportunity.

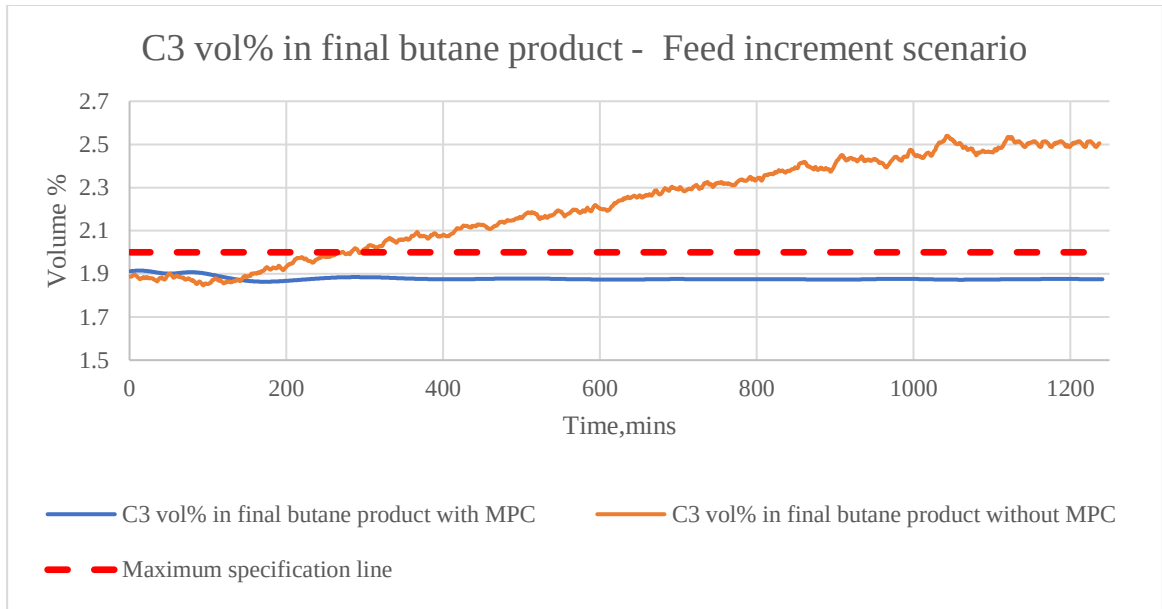


Figure 55 Depropanizer unit performance comparison for propane vol% in final butane product - Feed increase scenario

The above figure shows the system response in attempting to maximize propane slippage into the final butane product without violating the quality specification of 2 vol%. As mentioned earlier, increasing the feed will result in a lower temperature profile and leads over condensation of propane thereby causing more propane slippage in the bottom product. Due action will be required to continuously optimally adjust the temperature profile in order to ensure propane content in butane below the maximum permissible.

It can be seen that the unit under MPC manages to effectively maintain the propane content in the final butane product below the maximum with an average of 1.88 vol%. While the unit under no MPC sees the propane content increasing beyond the allowable specification a final purity of 2.50 vol%.

5.2.3. Feed decrease scenario

A scenario involving a feed decrease has been simulated. The feed has been decreased by around 15% to evaluate the MPC system capability in attempting to control the reboiler heat duty and reflux to the column in order to maintain the optimal propane product purity while maximizing propane slippage to butane product without violating the constraints.

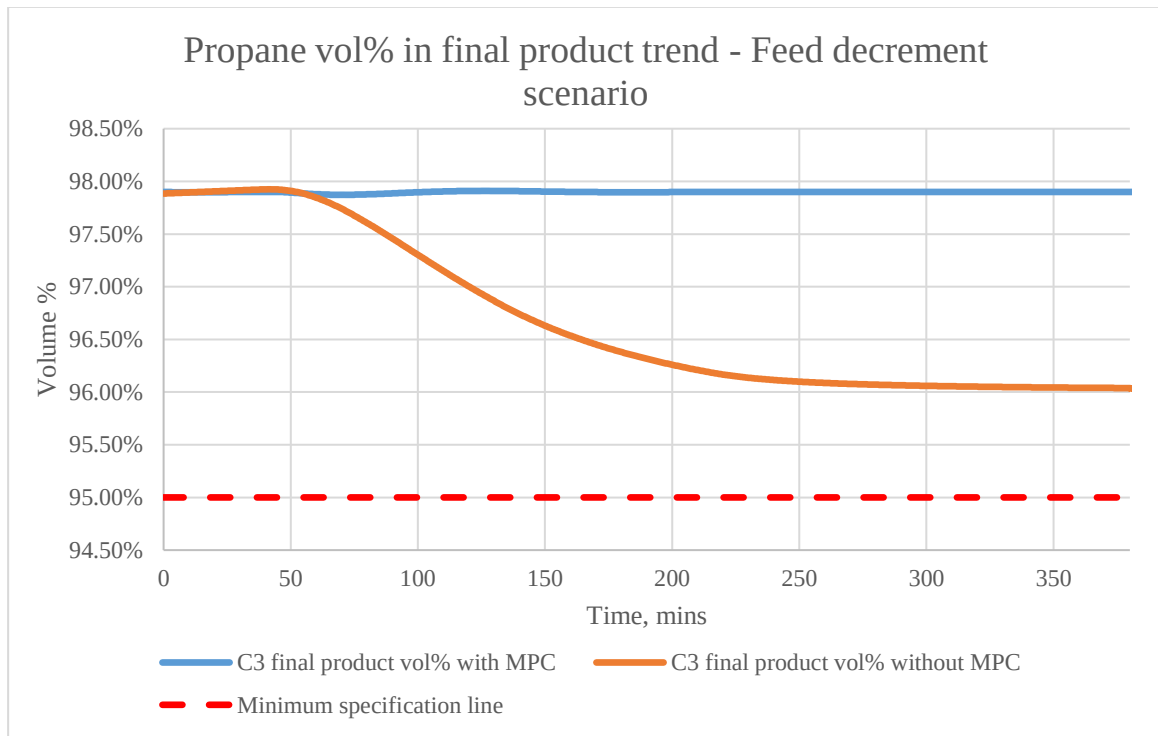


**Figure 56 Depropanizer unit performance comparison
– Feed decrease scenario**

Decreasing the feed suddenly to the column results in the temperature profile increasing. This leads to over vaporization of the butane components in the bottom of the column resulting in more butane products concentrating in the final propane product. This presents as an

uneconomical method of operation since butane is sold at a higher price. Figure 56 shows the comparison of the unit in terms of maintaining optimal propane purity with and without MPC. Under MPC, the unit achieves to maintain the target outlet purity of propane as close as possible to the optimal concentration. The overall purity achieved is 97.93 vol%. The MPC attempts to optimize this by continuously altering the TIC-115 and QIC DEC3 to lower the temperature profile of the column. This is done by recalculating the most optimal set points based on solving an optimization problem which attempts to find the best route that ensures lowest difference between the target setpoint and the predicted state while ensuring minimal moves.

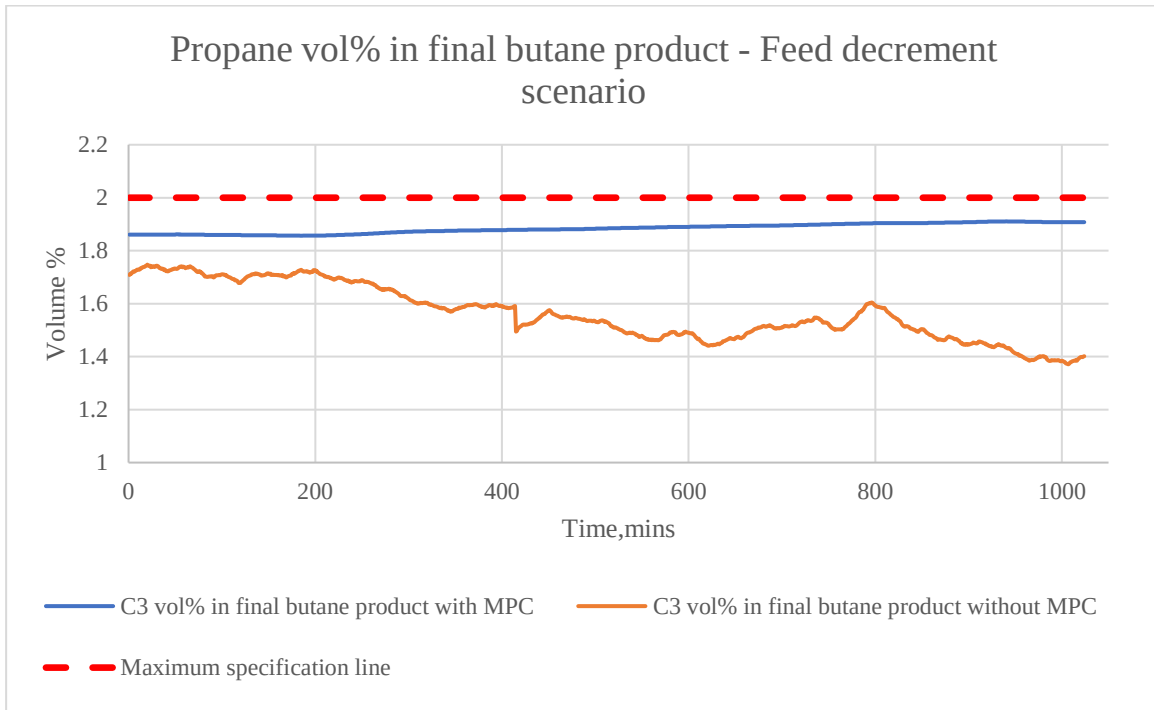


Figure 57 Depropanizer unit performance comparison for propane vol% in final butane product - Feed decrease scenario

Similarly, an increase in the temperature profile of the column results in excessive vaporization of the light ends in the bottom of the column which results in lesser slippage of propane in the bottom stream thereby reducing propane content in the final butane product. This presents as an uneconomical method of operation since this is considered as an economic opportunity loss.

Figure 57 shows the performance of the unit in terms of maximizing propane slippage in the final butane product. It is quite evident that the unit without MPC is not able to continuously maintain maximum propane slippage. Propane content in final butane product constantly decreases – due to the increasing temperature profile, to a final state of around 1.40 vol%. This presents as a loss of profit opportunity since increasing propane content in the butane product results in improved profit margin. However, the unit under MPC manages to continuously maximize propane slippage with the bottom product hence resulting in maximization of propane content in the final butane product. An overall average propane content in butane product of 1.88 vol% is maintained throughout the simulation scenario thereby achieving a more superior economic performance.

5.2.4. Feed composition change – Increase of propane content in feed

A scenario whereby propane content in the feed increased from 34.56 vol% to 40.56 vol% is simulated. Such disturbances occur where there are upsets or process changes happening especially in the ROG coming from the nearby refinery.

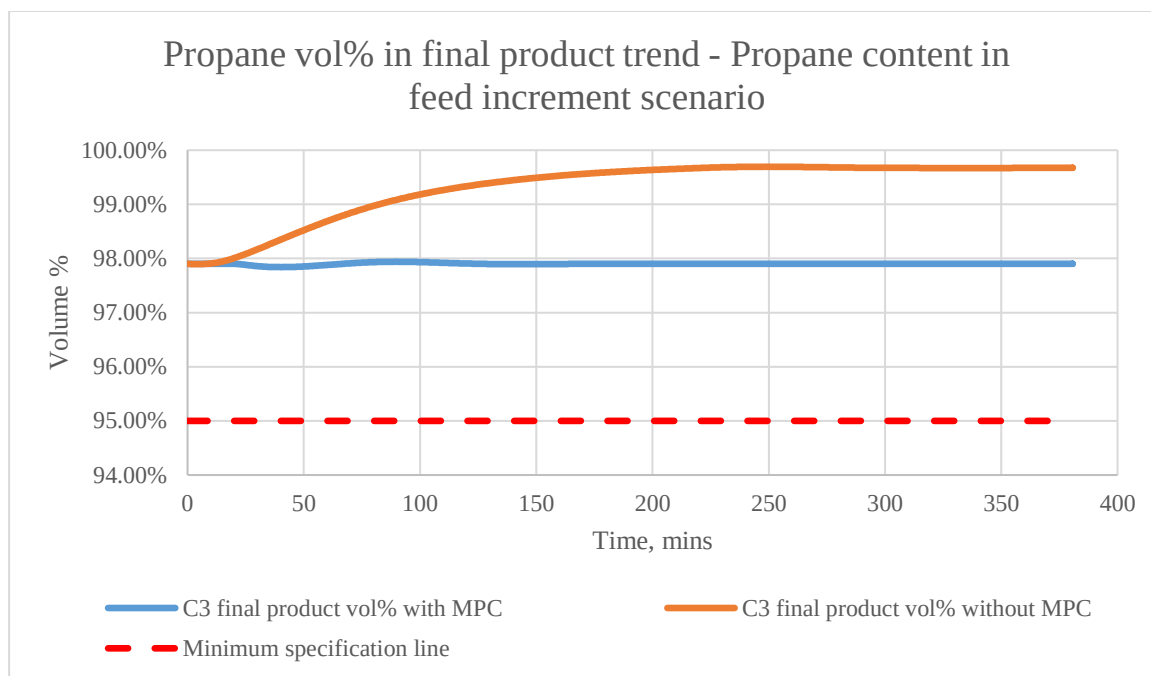


**Figure 58 Depropanizer unit performance comparison
– Increase of propane content in feed scenario**

The above figure shows the difference in performance of the unit between MPC and no MPC. Increasing the propane composition in the feed naturally means that more propane content will be present in the vapor liquid traffic inside the column. Appropriate adjustments will be required in order to ensure operation at the optimal point. From the figure, the unit under MPC successfully manages to maintain propane at an average of 97.85 vol%. Given that the upstream Deethanizer unit MPC attempts to maximize ethane slippage close to 2 vol% in the propane product naturally means that minimal slippage of butane product in the overhead propane product. In contrast, the unit under no MPC does not manage to maintain operation at the optimal point. It is evident from the figure that propane content increases to around 99.67 vol% hence presenting a product giveaway and loss of economic opportunity.

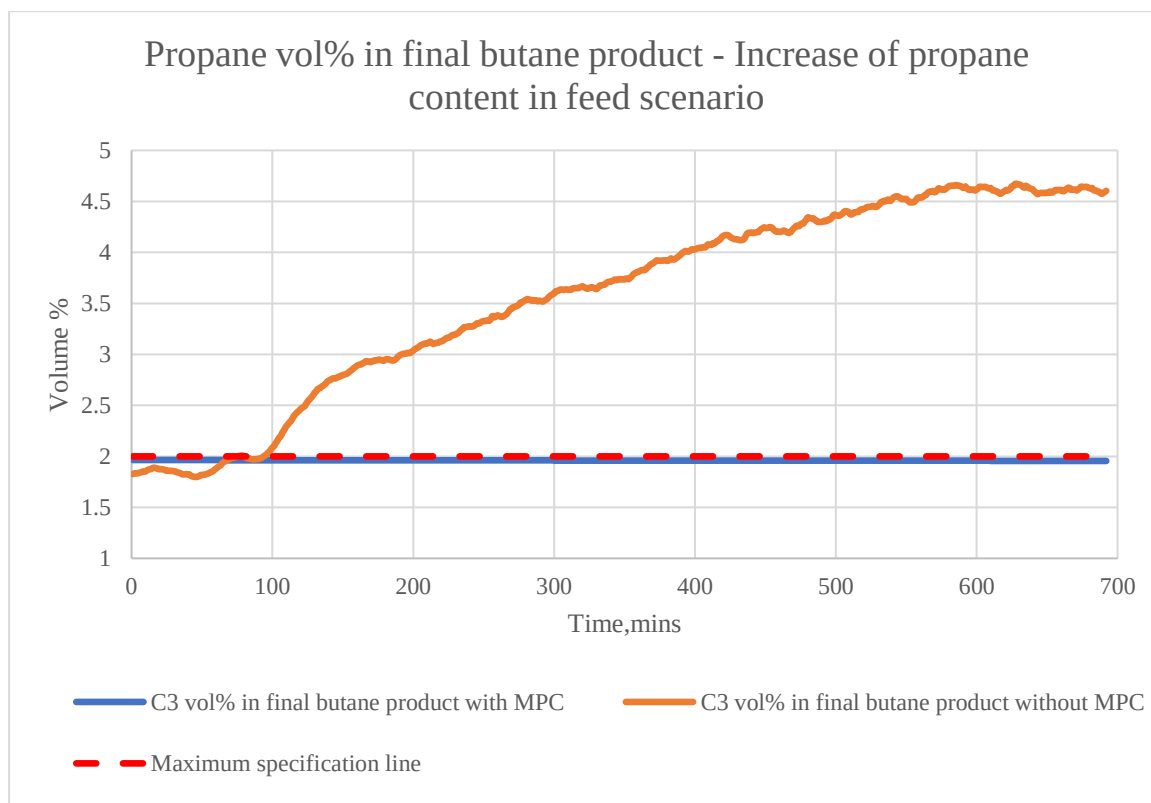


Figure 59 Depropanizer unit performance comparison for propane vol% in final butane product – Increase of propane content in feed scenario

Similarly, increasing propane content in the feed will definitely impact the amount of propane slippage with the bottom product and due adjustment in the lower column temperature will be required in order to maintain maximum slippage while respecting quality constraints for the butane product. From the above figure, it is evident that the unit without MPC cannot manage to keep propane content within acceptable limits. The propane content under no MPC increase to 4.60 vol% in the final butane product which results in off specification production. However, under MPC the average propane content in the final butane product

averages 1.93 vol% which translates to an increase in profit margin since propane can be sold at the higher price of butane.

5.2.5. Feed composition change – Decrease of propane content in feed

Similarly, a scenario whereby propane content in the feed decreases from 34.56 vol% to 30.45 vol% has been simulated. As in the previous case such disturbances mainly occur due to upsets in the nearby refinery or a change in their daily operating plan which causes the quality of the ROG to change.

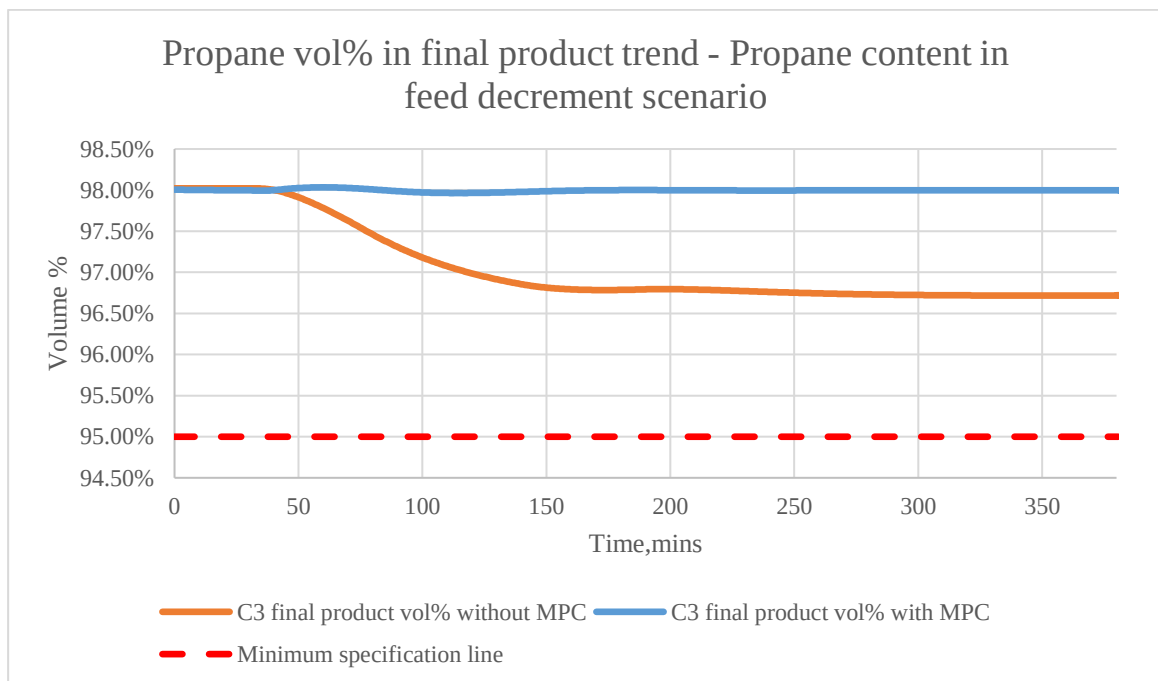


**Figure 60 Depropanizer unit performance comparison
– Decrease of propane content in feed scenario**

Figure 60 shows a comparison between the performance of the unit with MPC and without MPC. Decreasing the propane content in the feed naturally means that lesser propane content will be available in the

equilibrium vapor and liquid traffic phase inside the column. Under such conditions, it is imperative that in order to maintain optimal operation due adjustments will be required to lower the temperature profile. As can be seen from the figure, under no MPC the purity of the final propane product decreases. This is mainly explained by the fact that lower propane content in the column will require lower temperature profile to maintain performance. If no adjustments are made, then more butane will be vaporized and end up with the final propane product hence reducing the propane purity. This presents an uneconomical method of operation. With no MPC to constantly alter the column parameters, the propane purity proceeds to decrease to around 96.72 vol% which even though does not result in off specification production, represents an uneconomical method of operation due to more slippage of more valuable butane content in the propane product. In contrast, the unit under MPC manages to keep an overall propane purity of 98.21 vol% through the simulation scenario and hence as close as possible to the most optimal operating point.

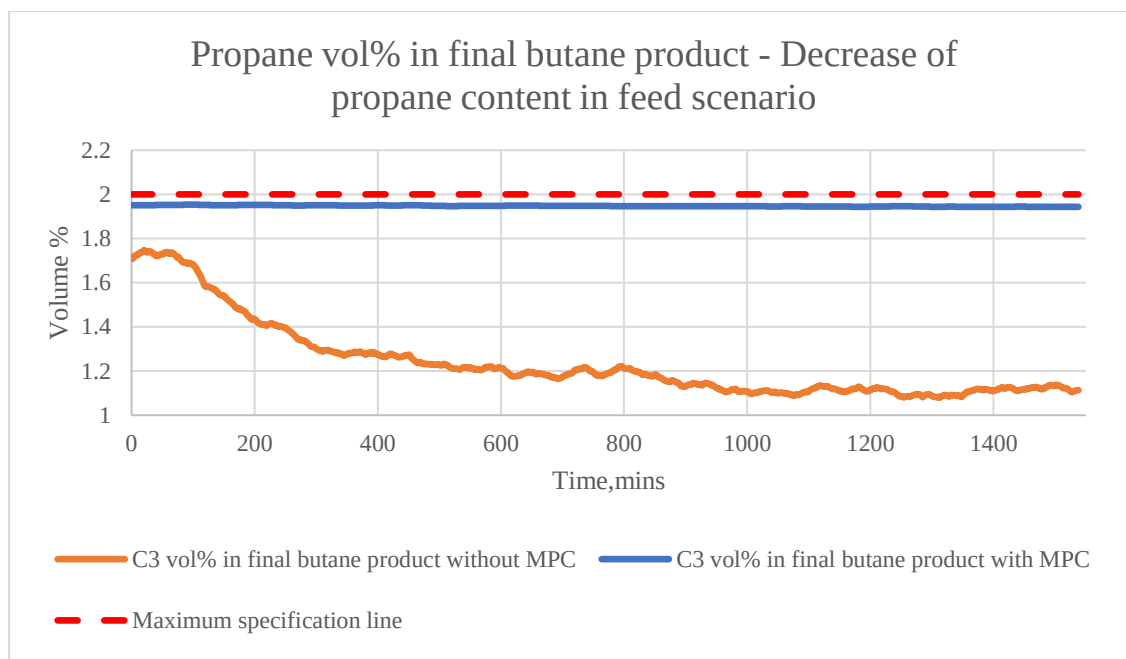


Figure 61 Depropanizer unit performance comparison for propane vol% in final butane product – Decrease of propane content in feed scenario

Decreasing propane content in the feed will most definitely impact the propane slippage performance of the column. Decreasing propane content will require lowering the bottom temperature of the column in order to continue maximizing propane slippage into the final butane product. From figure 61, it is clear that the unit operating without MPC cannot manage to keep the propane slippage as close as possible to the constraint line. From the simulation scenarios, it is seen that the propane content in the final butane product decreases to around 1.12 vol% and hence presents an economic opportunity loss since propane can be sold at higher price if managed to slip constantly with the butane product as long as quality constraint is met. However, the unit under MPC manages to maintain overall average of propane content in butane product of 1.94 vol% hence maximizing the propane slippage and therefore profit margin.

It is important to note that the effect of the propane content in final propane content in all of the scenarios appear to take longer time to react and reach a final steady state. This is due to the fact the Depropanizer column is a tower comprising of 42 trays with the feed inlet located at tray 27 and hence any effect on the bottom section will take a longer time to be realized.

It is imperative to note that while controlling product quality, there are certain degree of freedom that needs to be dealt with. Attempting to maximize ethane slippage into the propane product while at the same time maintain propane product purity at 98 vol% and assuring maximization of propane slippage along with the butane product automatically means that butane slippage along with the propane product is minimized and such benefits will be documented in the economic benefit analysis chapter.

5.3. Debutanizer unit

The results obtained from the simulation of the Debutanizer unit under MPC shall be presented below. The results include the details obtained from simulating 4 different scenarios – the details and basis for the scenarios shall be stated along together with a comparison between operating the unit with and without the MPC. A comparison shall also be done to demonstrate the variance handling capability of the unit with and without MPC.

The Debutanizer is the final unit in the fractionation train with the objective of separating both butane and naphtha and achieving purity specification acceptable by the market. Through observation of the plant operating records, it has been seen that there exists an opportunity for maximization of butane slippage into the naphtha product. There is no incentive in achieving higher purity naphtha since this would constitute a product giveaway and hence lost in profit opportunity. The main controlling points in this column is the reboiler duty and reflux to the column which controls temperature of tray no. 37.

5.3.1. Variance handling capability

One of the most important advantages that can be attributed to the MPC is its ability to reduce variation in product quality by employing a prediction model in the background that would participate in an objective function with the aim of determining the best course of action. One of the simulation scenarios conducted was to evaluate whether or not the naphtha RVP quality can be increased without violating the maximum specification line. Operating the unit away from the constraint line represents an economical loss since this would represent a reduction of butane slippage into naphtha – bearing in mind that naphtha is sold at a higher price.

It has been noticed from the plant operating records that plant operators tend to operate away from the maximum RVP to provide

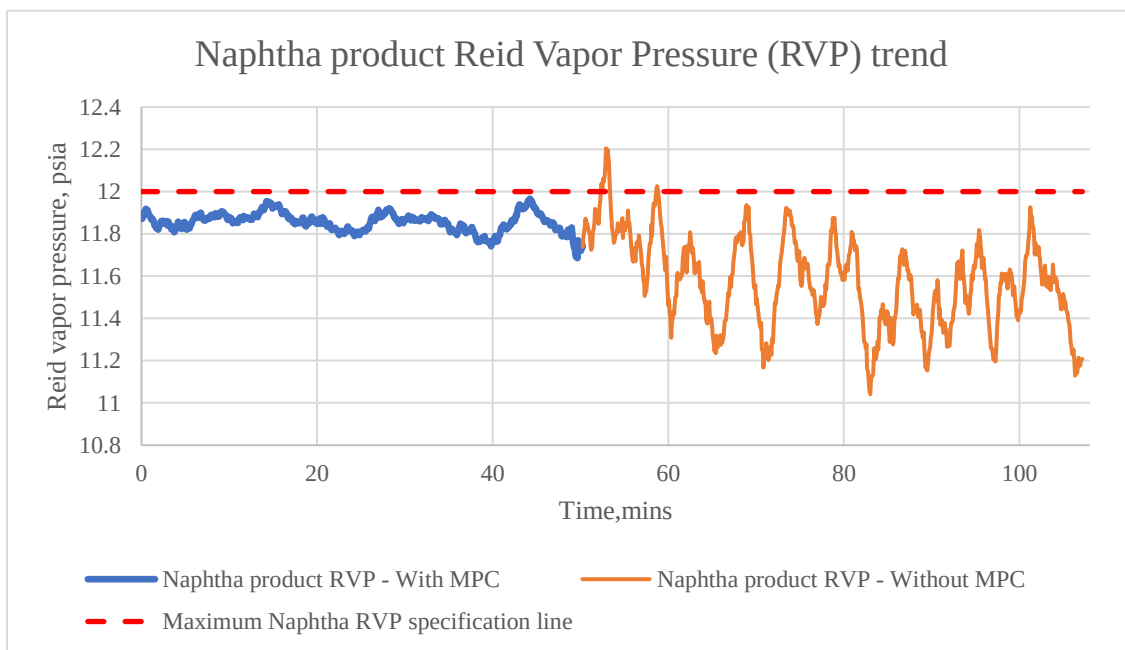


Figure 62 Naphtha RVP – Operating closer to the constraint line. Variance handling capability MPC vs NO MPC

a safety margin due to the fluctuating nature of the unit in terms of RVP product quality. Though this would represent a safe way of unit operation, this would also entail significant economic loss.

Figure 62 shows the difference in the variance handling capability of the unit with and without MPC. Due to the employment of the prediction model which is used to decide an optimal course of action the unit is able to reduce fluctuation in RVP quality of naphtha and hence naturally would allow the unit to operate closer to the constraint line. The unit under MPC manages to maintain an overall RVP quality of 11.90 psia while without MPC an overall RVP quality of 11.55 psia is maintained in order to provide sufficient margin to accommodate for the higher RVP fluctuation.

This entails a significant economic gain which would have been otherwise lost.

5.3.2. Feed increase scenario

A scenario has been simulated whereby feed to the unit has been increased by 7% to evaluate the capability of the MPC in adjusting the reboiler and reflux to the column which are the manipulated

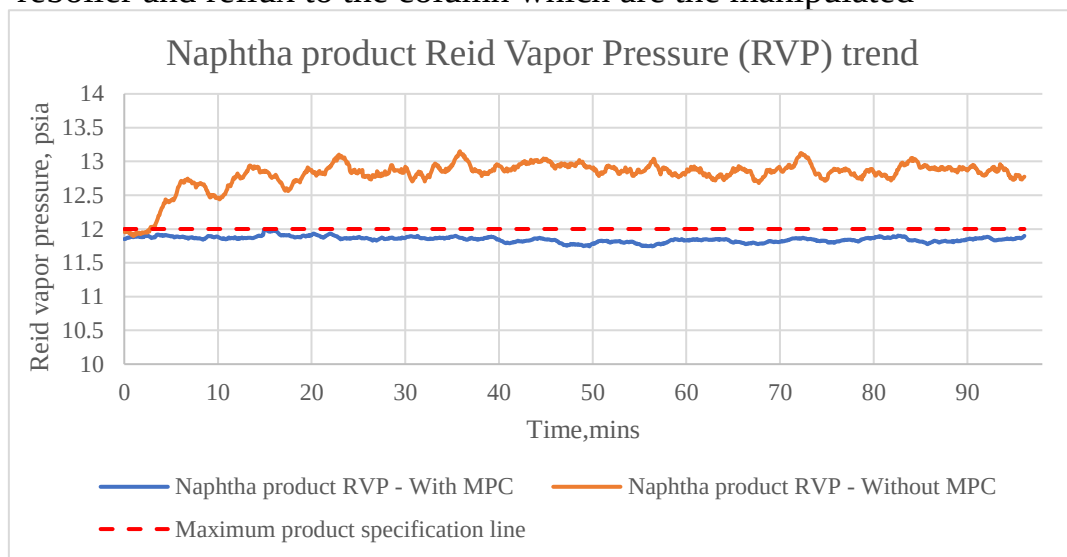


**Figure 63 Debutanizer unit performance comparison
– Feed increase scenario**

variables as to effectively maintain RVP of the naphtha product.

As can be seen above, the Debutanizer unit under MPC manages to keep RVP of the naphtha product as high as possible while respecting constraint limits by effectively navigating the reboiler duty and column reflux. The MPC achieves this by optimizing the reboiler duty and column reflux by utilizing the prediction model in the background in order to decide an optimal course of action. The unit under MPC manages to achieve an overall RVP of 11.84 psia in the final naphtha

product compared to a final steady state of 13.02 psia in case of operation without MPC thereby resulting in significant off specification production. Increasing the feed to the column inevitably increases the total hydraulic load in terms of total butane content in the column and reduces temperature profile of the column. This means that more reboiler duty and higher column temperature profile will be required to maintain performance. The MPC successfully manages to calculate new set points regularly for both the reboiler and tray no. 37 temperature controller that directly alters the reflux rates to the column.

5.3.3. Feed decrease scenario

A scenario involving a feed decrease has been simulated. The feed has been decreased by around 7% to evaluate the MPC system capability in attempting to control the reboiler heat duty and column reflux in order to maintain RVP of the naphtha product as high as possible without violating the specification limit.

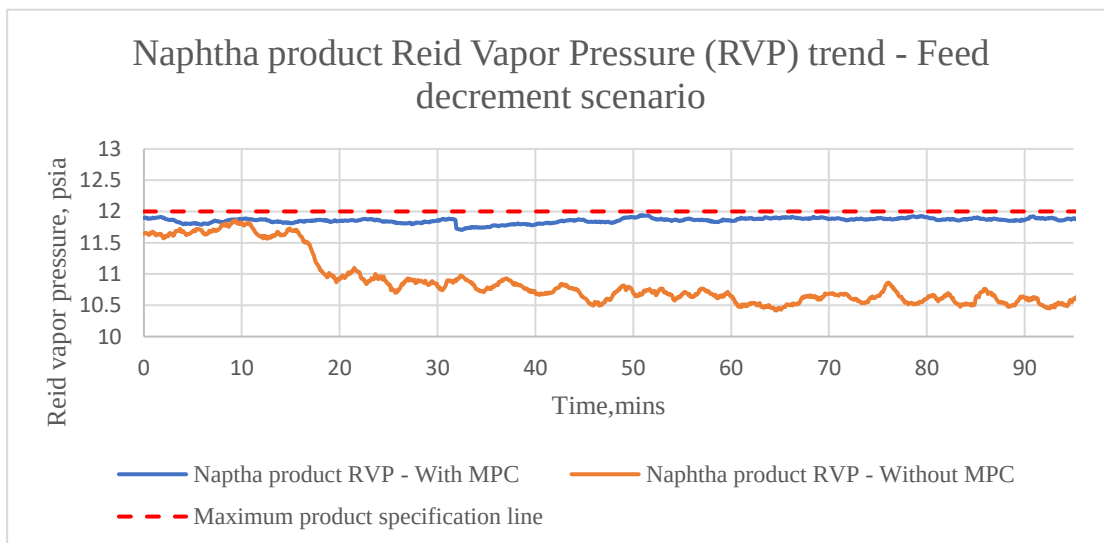


**Figure 64 Debutanizer unit performance comparison
– Feed decrease scenario**

As can be seen in figure 64, as with the previous scenario, the unit under MPC successfully manages to maximize RVP of the naphtha product while respecting constraints by effectively controlling the reboiler duty and column reflux. The unit under MPC manages to maintain an overall RVP of 11.86 psia in the naphtha product while without MPC the RVP decreases to 10.58 psia which represents an economical loss.

Decreasing the feed to the column inevitably means that lesser amount of butane enters the column which would require lowering the reboiler duty and maintaining a cooler temperature profile in order to maximize slippage of butane into the naphtha product and, hence optimizing economic performance.

5.3.4. Feed composition change – Increase in butane content

Similar to previous units, the performance of the Debutanizer is also affected by the change in composition primarily due to changes in the ROG coming from the nearby refinery. In case of changes in the ROG quality, the column operating parameters will need to adjust in order to maintain performance. The ROG is generally heavy in nature i.e., it contains more butane and naphtha. However, the split between these two can affect the performance of the unit. Currently, changes to the column operating parameters largely depend on the judgement of the operator controlling the unit and this inevitably means that different operators will control the plant differently with dissimilar results.

A simulation scenario has been conducted whereby butane content in the feed changes from 56.65 to 61.59 vol %. This basis has been obtained after reviewing the plants operating history in terms of variation in composition.

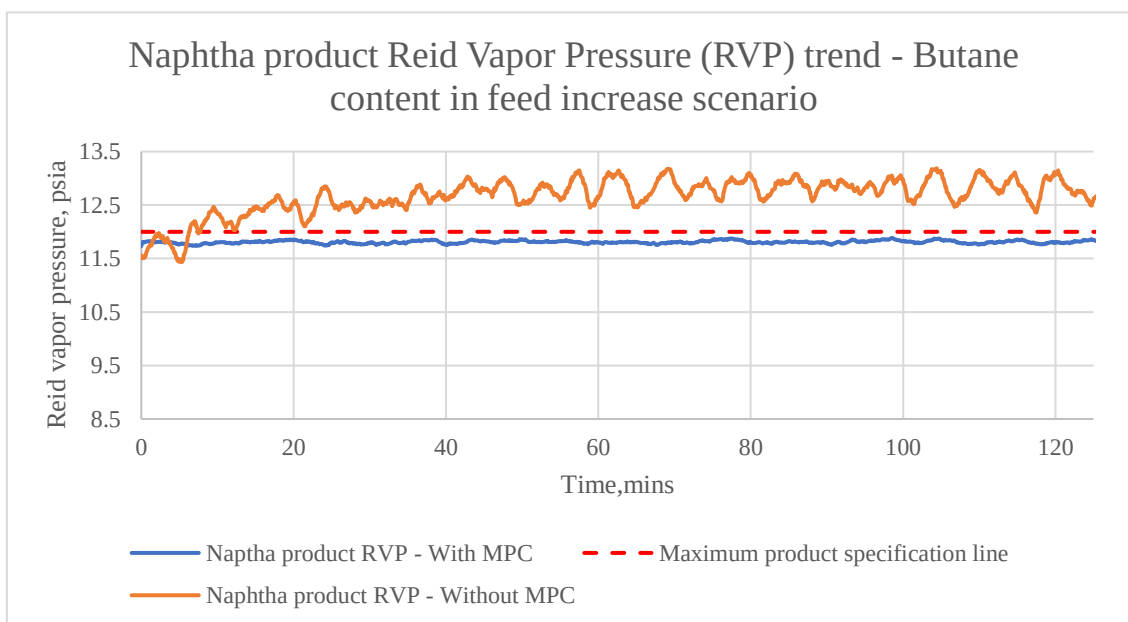


**Figure 65 Debutanizer unit performance comparison
– Increase of butane content in feed scenario**

As seen in figure 65, the MPC demonstrates an ability to maintain the RVP of the naphtha product as high as possible without violating the constraint. Increase the butane content in the feed naturally means that more heat and higher temperature profiles will be required to order to ensure more stripping of butane from the naphtha product. The MPC manages to ensure high RVP by effectively optimizing the reboiler duty and tray no.37 temperature at each control interval.

The MPC manages to operate the unit under the subject scenario to effectively keep the overall RVP in the naphtha product at 11.81 psia while without the MPC the RVP increases to around 12.75 psia and hence producing off specification naphtha.

5.3.5. Feed composition change – Decrease in butane content

Similarly, a scenario has been simulated whereby a decrease in the butane content in the feed to the Debutanizer unit occurs. Again, such scenarios tend to happen mainly due to changes in quality of the ROG coming from the refinery. The butane content in the feed has been reduced from 56.65 to 49.75 vol%.

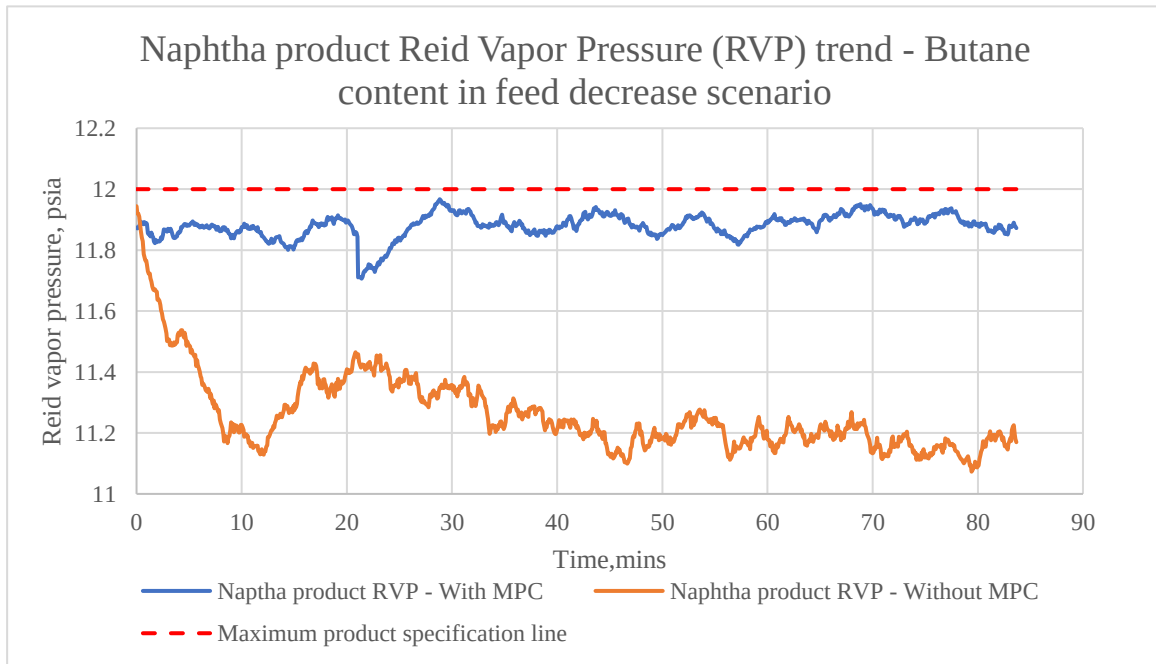


**Figure 66 Debutanizer unit performance comparison
– Decrease of butane content in feed scenario**

Figure 66 shows the results of obtained by simulating the subject scenario. The unit under MPC effectively manages to maintain the RVP of the Naphtha product as close as possible to the constraint achieving an overall RVP of 11.88 psia while without MPC the RVP decreases to around 11.20 psia which represents a loss of profit opportunity.

5.4. Conclusions

This chapter presented the results obtained from the MPC simulation of the different units under consideration from the plant, namely the Deethanizer, Depropanizer and Debutanizer unit from which significant economic benefits arising from improved control was expected. The MPC simulation is the final layer of simulation which comes above both the steady state and dynamic state simulation.

Different simulated scenarios to test the efficacy of the proposed MPC simulation were conducted. The basis of these simulations come from the actual situations faced by the plant operators. The simulation scenarios can be generally categorized as follows:

- (a) Feed increase scenario
- (b) Feed decrease scenario
- (c) Feed composition change scenario – Either increment of decrement of a key component in the feed
- (d) Variance handling capability of the controlled variable

For the Deethanizer unit, the main aspect that was measured in order to see the extent of improvement was the ability of the unit to maximize slippage of ethane into the bottom stream hence to the propane product. It has been seen that the MPC possess superiority in reducing variation and thereby increasing the average line closer to the maximum specification line. The unit under no MPC maintained an average ethane in final propane content of 1.90 vol% while under MPC an average

value of 1.98 vol% was maintained with reduction in standard deviation by almost 97%. The MPC control also showed superiority in attempting to continuously maximize ethane slippage into the propane final product in the different scenarios conducted.

As for the Depropanizer unit, the two controlled variables that were selected were – propane purity in the final product and propane content in the butane product. The objective in this case was to maintain propane product at 98 vol% which was deemed the most economical operating point while maximizing the propane slippage into the butane product. From the results obtained, it was seen that the MPC proved effective in trying to maintain propane purity as close as possible to the optimal point while maximizing propane slippage into the final butane product throughout all test scenarios that were conducted. In terms of variance handling capability, the MPC system again proved reliable in reducing variance and pushing the average line close to the constraint line. In this case, propane content in the butane product before implementing MPC averaged 1.75 vol% with a variance of 0.104 while under MPC the variance reduces to 0.029 thereby allowing the system to achieve a higher overall average of 1.95 vol%.

It is worth noting that even though not controlled directly, the butane slippage into the propane product is reduced since both the Deethanizer and Depropanizer MPC system work to maximize ethane content and maintaining of propane purity at 98 vol% hence naturally optimizing the reduction of the more valuable butane content in the propane product.

For the debutanizer system results were compared based on the ability of the unit to maintain maximum RVP of the naphtha product. It has been seen that in terms of variance handling capability, the MPC manages to reduce the variance from 0.19 to 0.060 thereby increasing the overall average from 11.55 psia to 11.90 psia, which will definitely increase profitability of the operation.

Chapter 6 - Economic benefit and payback analysis

6.1 Product prices – Propane, Butane and Naphtha

6.2 Method for estimating benefits

6.3 Plant historical data – Identification of improvement opportunities

6.4 Revenue increment based on improved operating line

6.5 MPC system cost and payback analysis

6.6 Conclusions

This chapter shall discuss an approach to quantify the economic benefits for the subject NGL plant that can be achieved by employing the proposed MPC. The benefits shall be quantified purely based on reduction of product quality giveaway (i.e., difference between the quality of the product and the quality specification) and on increasing of impurity slippage up to an acceptable level. First an economic basis shall be presented that will take into account historical prices from a baseline year. Actual historical records of the plant shall be presented and analyzed in order to identify the potential economic maximization opportunity. Further analysis will be done in order to quantify the identified economic opportunities. Then an estimate cost analysis shall be done for the MPC project in order to conduct the payback evaluation to justify the required investment.

6.1. Product prices – Propane, Butane and Naphtha

In order to estimate and quantify the economic benefits out of improved plant operation due to employment of MPC, it is first vital to establish the baseline prices of the products that would be used in the economic analysis. The table below presents the average contract trading prices for the months of 2019 as per the plant's records. The global energy sector has been affected by the COVID-19 pandemic with prices at certain times trading at an all-time low especially for Naphtha. Naphtha is mainly consumed by the aviation sector as aircraft fuel. Enormous negative impact on the tourism and aviation industry especially in 2020 resulted in low prices. Hence, fuel prices for the year 2020 has not been considered since this year represents an exceptional year that witnessed unprecedented wide scale global economic disruption and will not provide for a representative reflection of the product prices.

Table 9 2019 Monthly prices of LPG and Naphtha products \$/Metric tons – Company actual contract figures

Month	Propane \$/MT	Butane \$/MT	Naphtha \$/MT
19-Jan	430.00	420.00	448.00
19-Feb	440.00	470.00	493.00
19-Mar	490.00	520.00	526.00
19-Apr	515.00	535.00	560.00
19-May	525.00	530.00	529.00
19-Jun	430.00	415.00	454.00
19-Jul	375.00	355.00	491.00
19-Aug	370.00	360.00	442.00
19-Sep	350.00	360.00	472.00
19-Oct	420.00	435.00	473.00
19-Nov	430.00	445.00	503.00
19-Dec	440.00	455.00	527.00
Average Price	434.58	441.67	493.17

Hence, basis for economic benefit estimations shall be as follows: Propane 434.58 \$/MT, Butane 441.67\$/MT and Naphtha 493.17\$/MT.

6.2. Method for estimating benefit

From the literature review conducted in section 2.5.1., one of the methodologies that is used to assess the extent of benefits that arises from improved control methodology is the best operator technique. The premise of the best operator technique is that an improvement resulted from an advanced control system can be estimated by looking at the historical best operation previously attained. This is one of the methods used in industrial applications to justify required investments (SETPOINT Inc., 1986).

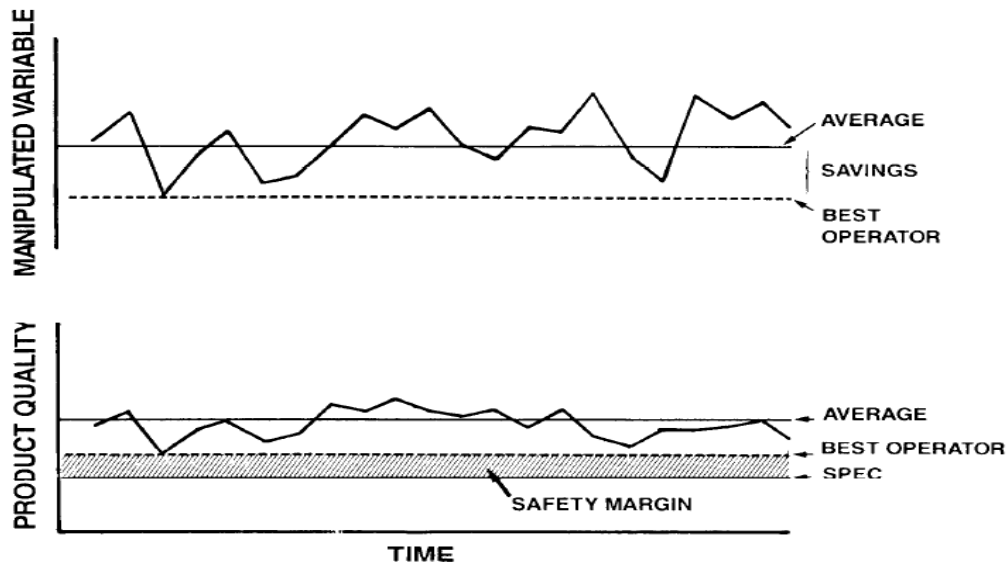


Figure 67 Best operator technique - Benefit estimation

This technique makes use of the actual historical records of the plant in trying to assess the benefits. This technique employs the idea of using the best actual operating point achieved with a reduced variance and is compared to the actual average of the current system. The difference between them is regarded as the benefit that can be achieved by advanced process control.

This technique is considered conservative, since it does not give any credit to the MPC in case the system performs better than the best operator practice.

6.3. Plant historical data – Identification of improvement opportunities

A review on the plant historical data shall be presented along with identification of improvement opportunities. The historical record shall mainly be the impurity slippage into the main products – Propane, Butane and Naphtha.

6.3.1. Ethane impurity in Propane Product

Propane product from the plant should contain a maximum ethane content of 2 vol% in order to meet international market specification.

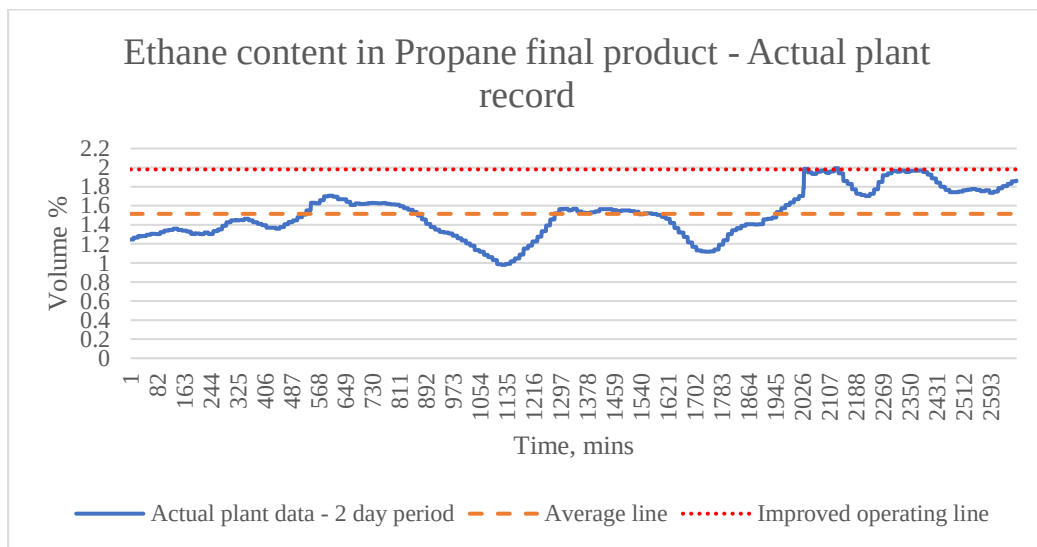


Figure 68 Ethane content in propane final product - Actual plant record

Figure 68 shows the historical data of the plant production quality concerning ethane slippage to propane product. The trend shows a two-day trend which is quite representative of the control strategy usually employed by the operators. The figure shows the behavior of the ethane in propane content over 6 shifts.

The performance of the unit in terms of ethane slippage depends on many factors such as gas quality, ambient temperatures, process upsets and mainly on the experience of the control operator in trying to maintain maximum slippage under all these variable conditions. Upon discussion with the plant operators, it was found that there exist instructions in order to attempt to maximize ethane slippage into the final product. However, no structured monitoring system exists for monitoring this operation and hence is almost up to the convenience of operator in trying to maximize the slippage over the shift. Now this mainly depends on the experience and the ability of the control operator in attempting to slip maximum ethane under the changing conditions of gas quality, process upsets and other environmental considerations.

The overall average ethane content was found to be 1.51 vol% over the 6-shift period. The highest sustained ethane in propane content was found to be 1.98 vol% which is evident in the last shift characterized by the two peaks observed in the end. This represents a significant economic opportunity loss as more ethane impurity can be pushed and sustained into the propane product. The dashed red line represents the possible improved operation line based on the concept of best operator technique for benefit estimation which is sustaining of 1.98 vol% ethane in propane product.

As mentioned previously, ethane is considered a byproduct from the plant and is not sold and hence there is significant incentive in trying to slip ethane into the propane product to be sold as propane (while

maintaining propane product specification) which otherwise would be been sent back for well re-injection or burned as utility fuel for which there exists no economic incentive since other types of gas, mainly methane are available for such purposes. Hence basis for economic benefit estimation shall be the following: current average ethane content in propane product is 1.51 vol% while the improved operating line shall be 1.98 vol%.

6.3.2. Propane in Butane product

The butane product from the plant has a maximum specification for propane content of 2 vol% in order to meet international market specification.

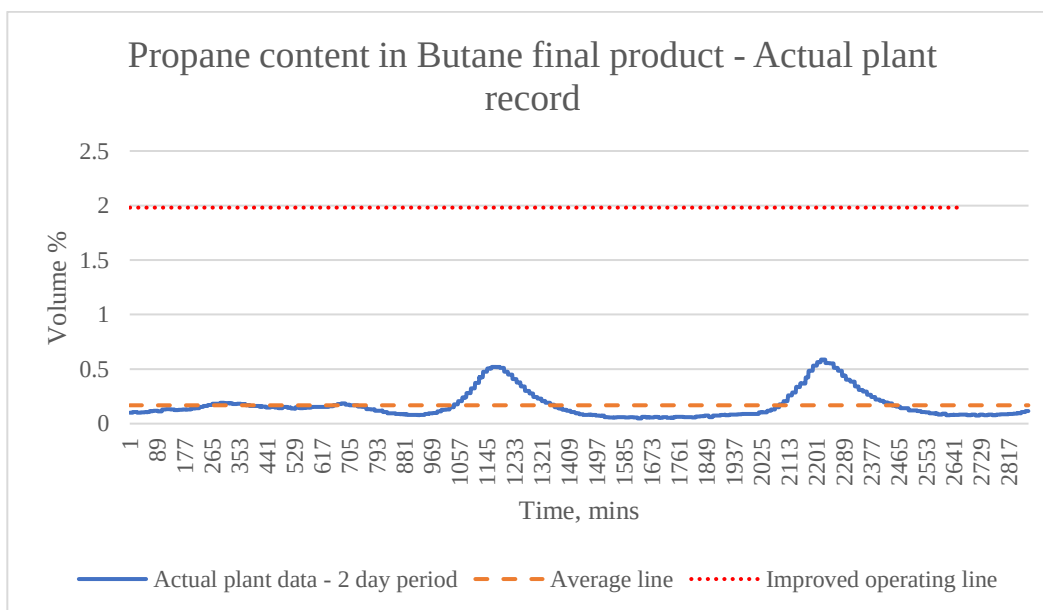


Figure 69 Propane content in butane final product - Actual plant record

Figure 69 shows the operational record of the production quality specifically for the propane content in butane product. Both propane and butane are considered products sold by the plant however, butane on an

average trade higher than propane by \$ 7.10 per metric ton (according to 2019 average prices). Hence there exists incentive in attempting to slip more propane into the butane final product.

Controlling the Depropanizer unit in order to slip more propane is less problematic and easier compared to that of the Deethanizer unit in attempting to maximize ethane slippage. This is mostly due to the fact the Deethanizer unit contains more operational constraints especially the reboiler capacity which is very evident when gas quality received is leaner than usual where the unit struggles to strip ethane without rerouting propane product back into the feed in order to artificially increase the molecular weight of the incoming feed to the unit.

Controlling the propane slippage to butane on the other hand is less problematic and the operation of which can be handled more smoothly, which is evident from the graph, as the trend appears to be relatively flat with no significant fluctuations. However, from the figure it was observed that the average propane content in butane is around 0.17 vol% which is very far away from the maximum allowable specification of 2 vol%. Upon discussing the issue with the concerned department personnel in the subject plant as to why there hasn't been at least some attempts to constantly target maximum slippage of propane into butane, it was revealed that there is no system in place that would monitor this operation performance such as a KPI or even standing instructions. As such, plant operators overlook this step and do not attempt to maximize propane slippage. This represents a significant loss of profit opportunity. From the simulation results, it has been seen that that the MPC can systematically and regularly optimize the control moves that would

enable maximum slippage of propane into butane even under changing feed conditions which it does by primarily reducing variation in the product quality by taking advantage of the process model in the background.

In this case, the best operator technique will not give a representative picture on the potential benefit estimation since from historical data, plant operators were never attempting to constantly maximize propane slippage in the first place due to lack of performance KPI monitoring and instructions in this regard.

It is more representative to consider the theoretical possible improvement in operation with adequate safety margin. The theoretical improvement in performance as seen from the MPC simulation is around 1.95 vol%. Hence basis for economic benefit estimation shall be the following: current average propane content in butane product is 0.17 vol% while the improved operating line is 1.95 vol%.

6.3.3. Improved recovery of butane from propane product

Another expected improvement due to the implementation of the MPC system is that arising from reduced butane slippage into the propane product. Butane on an average trades higher than propane and hence an uneconomical mode of operation arises if butane slips into the propane product.

The below is the graph representing actual plant performance in terms of butane content in the propane product.

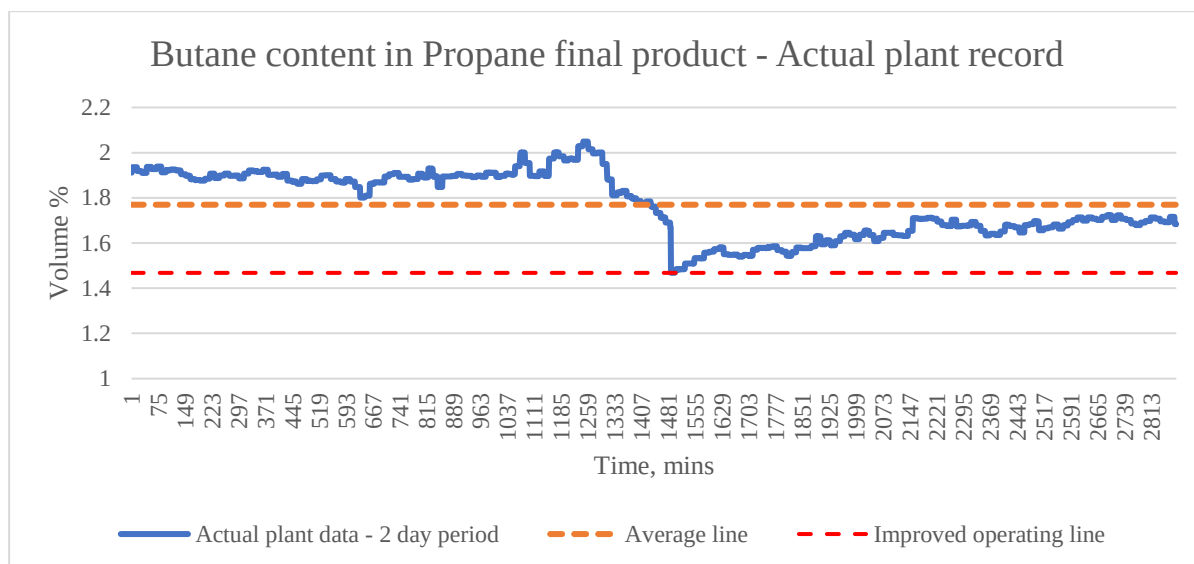


Figure 70 Butane content in propane final product - Actual plant record

From the records, the average butane content through the observation period is around 1.77 vol%. This presents a wasted opportunity since butane is sold at a higher price. Assuming the more conservative estimation in this case, by the best operator technique it is estimated that an improved operating average of 1.41 vol% is achievable. Hence, basis for economic benefit estimation shall be the following: current average butane content in propane final product is 1.77 vol% while improved operating line is 1.41 vol%.

6.3.4. Naphtha Reid Vapor Pressure RVP

Naphtha product from the plant has a maximum RVP specification of 12 psia as per market specification and lifting contracts. The below figure shows the two-day record for the actual RVP trend of the naphtha product from the plant.

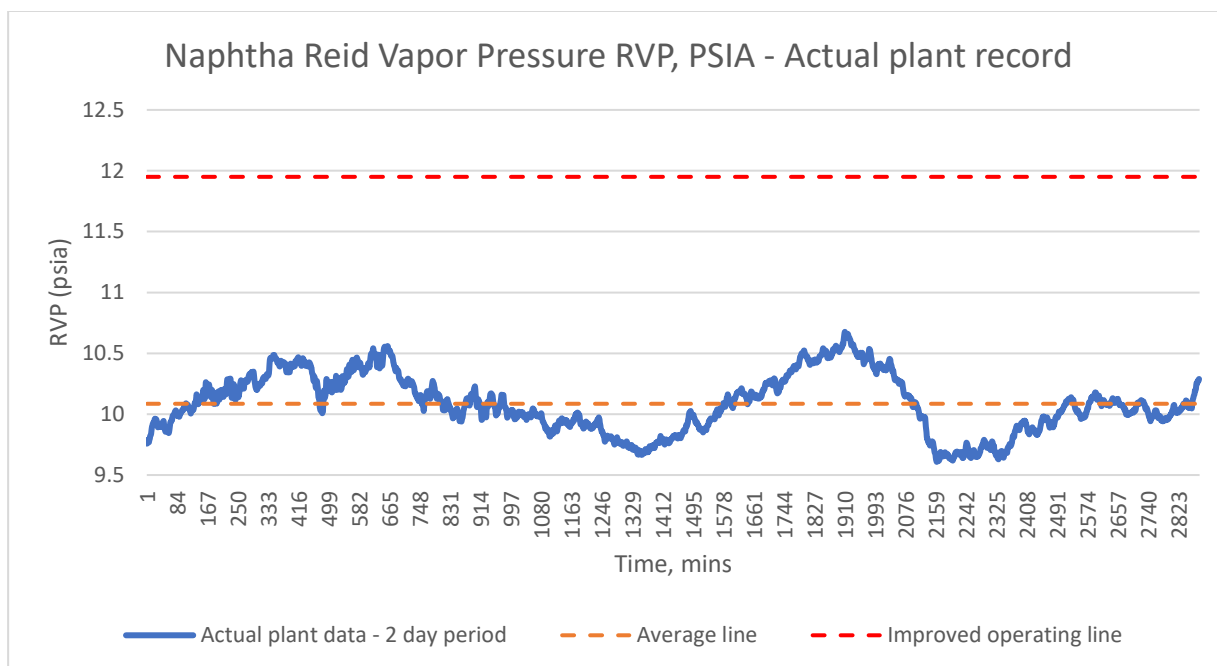


Figure 71 Reid Vapor Pressure RVP in Naphtha final product - Actual plant record

The RVP is a measure of how much low boilers is contained in the naphtha product. A higher RVP reading means that more low boilers (butane) is contained in the product. Naphtha prices are higher than that of butane with a difference of \$51.50 per metric ton (according to 2019 average prices) and hence provides an incentive to increase butane slippage in naphtha. Controlling the RVP is mainly done by altering the temperature profile of the Debutanizer column. Lower reboiler duty and adequate reflux rates should be able to increase butane slippage and hence higher RVP. However, the right temperature profile is a function of the current feed quantity and composition for which continues adjustment must be made in order to optimally and ensure maximum RVP of the product.

From figure 71 it is evident that the control operators of the Debutanizer unit do not systematically attempt to sustain the highest possible RVP. From discussions with the personnel from the concerned department, it has been revealed that similar to the previous unit, no KPI, monitoring system or standing instructions is in place in order to attempt to sustain higher RVPs. The discussions also reveal that the performance of the unit in terms of RVP largely depends on the control operator personnel on whether the control man chooses to slip more butane in order to increase RVP. Careful control operators would attempt to maximize RVP as much as possible leaving lesser safety margin while other control operators prefer to take the safer route and leave more safety margin in order not to produce off spec products in case of any disturbance.

Such different strategies employed by the control operators have two main problems:

- (a) Although certain operators attempt to increase the RVP until it reaches a certain high value, the overall average of the RVP over the course of the period under study is around 10.20 psia, which is quite far from the constraint limit of 12 psia.
- (b) The certain high value to which the RVP is allowed to increase before dropping is a value selected with a lot of safety margin. This together with the fact that there is no systematic method or controls in place that would incentivize the operator to continuously monitor and attempt to maximize

naphtha RVP means that the historical data does not represent any instance of best operator demonstration.

From the simulations conducted to evaluate how the MPC can control improve the RVP operating line by reducing variation it was seen that an average of 11.90 psia was sustained over the simulation period. Hence, basis for the economic benefit estimation shall be the following: current average RVP in naphtha product is 10.20 psia while the improved operating line is 11.90 psia.

6.4. Revenue increment based on improved operating line

An analysis of the estimated increase in financial revenue from the implementation of the improved operating line shall be presented here. All production figures used for the analysis have been obtained from the subject plant's annual performance review which is publicly available (Bahrain National Gas Company B.S.C, 2019).

6.4.1. Increase of ethane slippage into propane product

As mentioned earlier, ethane is considered an impurity and is not sold as a product. Majority of ethane leaves the plant with the residue-gas is either reinjected back into the reservoir in order to maintain wells pressure or send along with methane as fuel gas. Therefore, there is an economic incentive in attempting to slip more ethane into propane (while maintaining product quality constraint). The below table presents an estimated economic benefit analysis for the subject case.

Table 10 Economic benefit analysis - Increase of ethane slippage into propane product

Total annual propane production, Barrels	2,313,627.00
Average ethane vol% in product	1.51
Ethane barrels sold as propane	34,935.77
Improved ethane slippage vol%	1.98
Total annual production under improved ethane slippage	2,324,720.70
Ethane barrels sold as propane under improved ethane slippage	46,029.47
Difference in ethane barrels sold as propane	11,093.70
Propane conversion (Barrel/Metric ton)	12.38
Price of propane \$/ MT	\$434.58
Differential revenue \$ / year	\$389,426.58

The current average ethane content in propane is taken as 1.51 vol%. As demonstrated by the best operator technique, this average can be pushed further to 1.98 vol%. Under such this condition and assuming average product prices for 2019, it is expected that an incremental annual revenue of \$389,426 will be achieved. This is mainly due to the fact that increasing ethane slippage to 1.98 vol% would result in an increase of around 11,093 barrels of ethane sold as propane annually and hence increasing profit margin. This represents in an increase in propane production by 0.48%.

6.4.2. Increase of propane slippage into butane product.

Even though propane and butane are both considered products, butane on an average trades at a higher price and hence presents an economic

opportunity for maximizing profit by increasing slippage of propane into butane.

From analysis and discussions earlier, there appears to be no formal structured system or even standing instructions in place to continually monitor the maximization of this slippage. This is reflected in the relatively low propane content in butane. This presents another significant opportunity for profit maximization. The current average propane content in butane is around 0.17 vol% with the limit being 2 vol%. As mentioned earlier, due to the fact that no formal system is in place that would monitor or incentivize the slippage of propane into butane, obtaining an estimate of the possible improvement by the best operator technique will not provide a representative picture of the improvement since historical trends does not reflect such “*best operator*” moments. Hence an improvement using the simulation results has been considered.

The below table presents an economic analysis for the subject scenario -

Table 11 Economic benefit analysis - Increase of propane slippage into butane product

Total annual butane production, Barrels	2,084,647.00
Average propane vol% in butane product	0.17
Propane sold as butane, Barrels	3,543.90
Improved propane slippage vol%	1.95
Total annual production under improved propane slippage, Barrels	2,122,491.69
Propane sold as butane under improved propane slippage, Barrels	41,388.59
Difference in propane sold as butane, Barrels	37,844.69
Propane conversion factor (Barrel/Metric ton)	12.38

Butane conversion factor (Barrels/Metric ton)	10.84
Price of propane, \$ / MT	\$434.58
Price of butane, \$ / MT	\$441.67
Differential revenue \$ / year	\$213,484.62

From the analysis an estimated annual economic benefit of \$213,484 is expected to be achieved due to the slippage of 41,388 barrels of propane into the butane product this represents an increase of 1.81% of total annual butane production. In comparison, under the current control configuration and using current composition analysis only 3,543 barrels of propane is blended with the butane product.

6.4.3. Improved recovery of butane from propane product

Similar to the above, even though both propane and butane are considered products sold by the plant, butane on an average trade at a higher price and hence there is no incentive in slipping butane along with the propane product.

From our analysis and discussion, there appears to be no formal system in place in order to check this performance and is left mostly to the convenience of the operator controlling the unit whether to reduce butane slippage or not. It is be to noted that even though the butane content in propane was not directly selected to be one of the controlled variables, it is by logic indirectly controlled since the MPC design maximizes both ethane slippage (around 1.95 vol%) and also maintains propane purity as

close as possible to the 98 vol% hence leaving minimal possibility for butane sipping along with the propane product.

Below table presents the economic benefit analysis expected to be attained from reduced butane slippage:

Table 12 Economic benefit analysis - Improved recovery of butane from propane product

Total annual propane production, Barrels	2,313,627.00
Average butane vol% in propane product	1.77
Butane sold as propane, Barrels	40,951.20
Reduced butane slippage vol%	1.41
Total annual propane production under reduced butane slippage, Barrels	2,305,178.82
Butane sold as propane under reduced butane slippage, Barrels	32,503.02
Total annual butane production current, Barrels	2,084,647.00
Difference in butane sold as propane, Barrels	8,448.18
Total annual butane production under reduced butane slippage, Barrels	2,093,095.18
Propane conversion factor (Barrel/Metric ton)	12.38
Butane conversion factor (Barrels/Metric ton)	10.84
Price of propane, \$ / MT	\$434.58
Price of butane, \$ / MT	\$441.67
Differential revenue \$ / year	\$47,656.78

The basis employed here is taken from the plant actual operating data. As per the data, the average butane content in the propane product is 1.77 vol% which presents a wasted opportunity. Taking a conservative approach of selecting a new operating average of 1.41 vol% based on the best operator technique, yields a reduction of butane slippage of 8,454 barrels. This results in an incremental revenue of \$47,693 annually.

6.4.4. Improvement of the naphtha product RVP

On average, the naphtha product trades higher by \$ 51.5/MT compared to butane. The maximum RVP specification set by contract clauses is 12 psia. From the plant historical records, it was seen that the average RVP was 10.20 psia. This represents significant opportunity loss in terms of revenue and hence been identified as a major area for improvement should the proposed MPC is implemented at the NGL plant, given the difference in trading prices of the product.

As presented earlier, from the plant historical data and through discussions with the plant operators it was observed that there does not exist any structured system or formalized instructions in place in order to increase the RVP of the naphtha. At most, this is mostly left for the operators in charge of the unit to decide about. Hence, like the previous case, the historical trend of the RVP does not signify any “best operator” moment and will not be representative of the extent of improvement in the operation. MPC simulation results indicate that the average RVP of naphtha can be increased to around 11.90 psia. This shall be considered for the economic analysis.

The below table presents an estimated economic benefit analysis for the subject scenario:

Table 13 Economic benefit analysis - Increase of naphtha product RVP

Total annual naphtha production, Barrels	3,113,824.00
Average RVP, PSIA	10.20

Butane barrels sold as naphtha	47,303.00
Improved RVP, PSIA	11.90
Total annual production under improved RVP, Barrels	3,226,533.00
Butane sold as butane under improved RVP, Barrels	160,012.00
Difference in butane sold as naphtha, Barrels	112,709.00
Butane conversion factor (Barrel/Metric ton)	10.84
Naphtha conversion factor (Barrel/Metric ton)	9.40
Price of butane, \$ / MT	\$441.67
Price of naphtha, \$ / MT	\$493.17
Differential revenue \$ / year	\$1,320,997.80

Increasing the RVP to 11.90 psia would result to a total slippage of 160,012 barrels of butane sold as naphtha annually. This is in contrast to 47,303 barrels of butane sold as naphtha as per 2019 production estimates. This in turn generates an additional revenue of \$1,320,997 annually, which is attributed by running tighter quality production.

6.4.5. Summary of the economic benefit analysis

From the economic benefit analysis conducted for the estimated operational improvement arising from continuously optimized performance due to the employment of MPC, a total annual incremental revenue of \$1,971,565 is expected to be achieved. This is based on evaluations conducted with the consideration of the plant's baseline production for the year of 2019 along with the average product prices of the same year. From the estimated incremental revenue identified from the analysis, the largest contributor to the incremental revenue comes

from the increase in the RVP quality of the naphtha product – contributing to 67% of the incremental revenue.

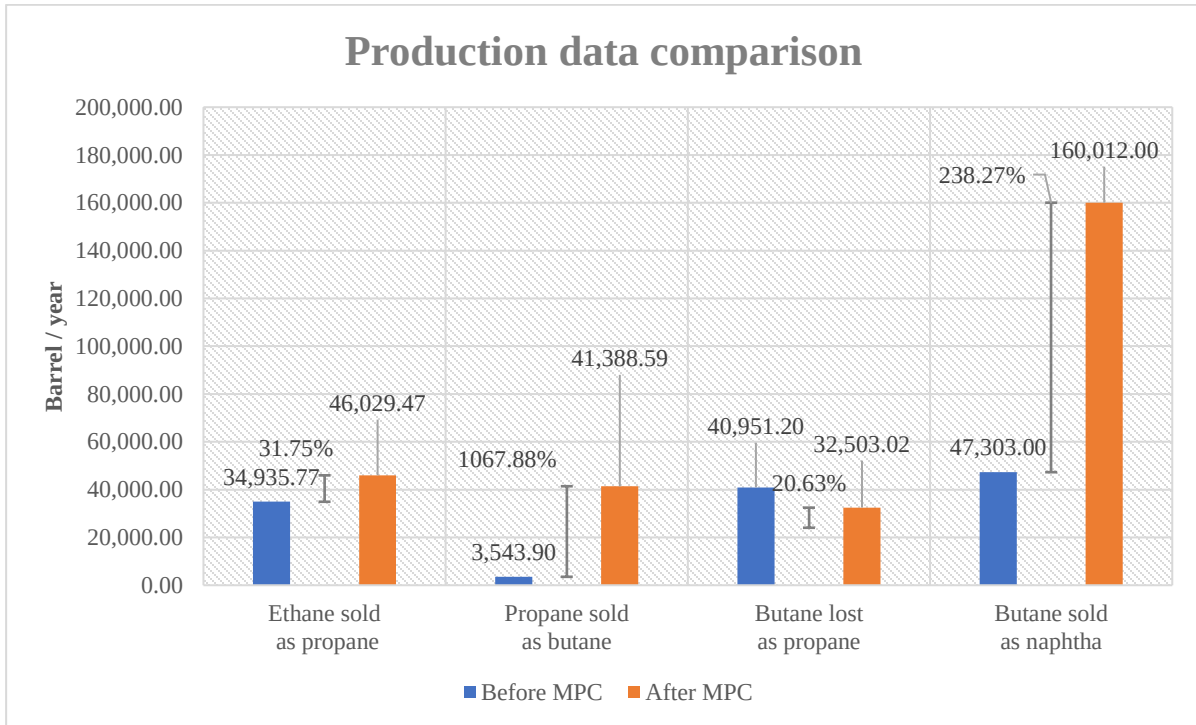


Figure 72 Production data comparison - Before and after MPC

From the figure, it is evident that in relative terms the increase is propane barrels sold, as butane is the most significant. An increase to 41,388 barrels of propane sold as butane annually is expected to be achieved, which is an increase by 1067%, compared to the baseline production of 3,542 barrels annually. However, in absolute terms the most significant increase in incremental production is achieved in naphtha where an incremental annual production of 112,709 barrels of butane sold as naphtha is estimated. This constitutes an increase of 238% increase compared to the baseline production considered in the analysis. As for the increase of ethane slippage in propane, the analysis shows that an increase of 11,093 barrels of ethane can be slipped into propane product which represents an increase by 31.75%. As for the recovered butane from

propane product, an estimated 8,448 barrels of butane can be recovered from the propane product which is a decrease by 20.65%.

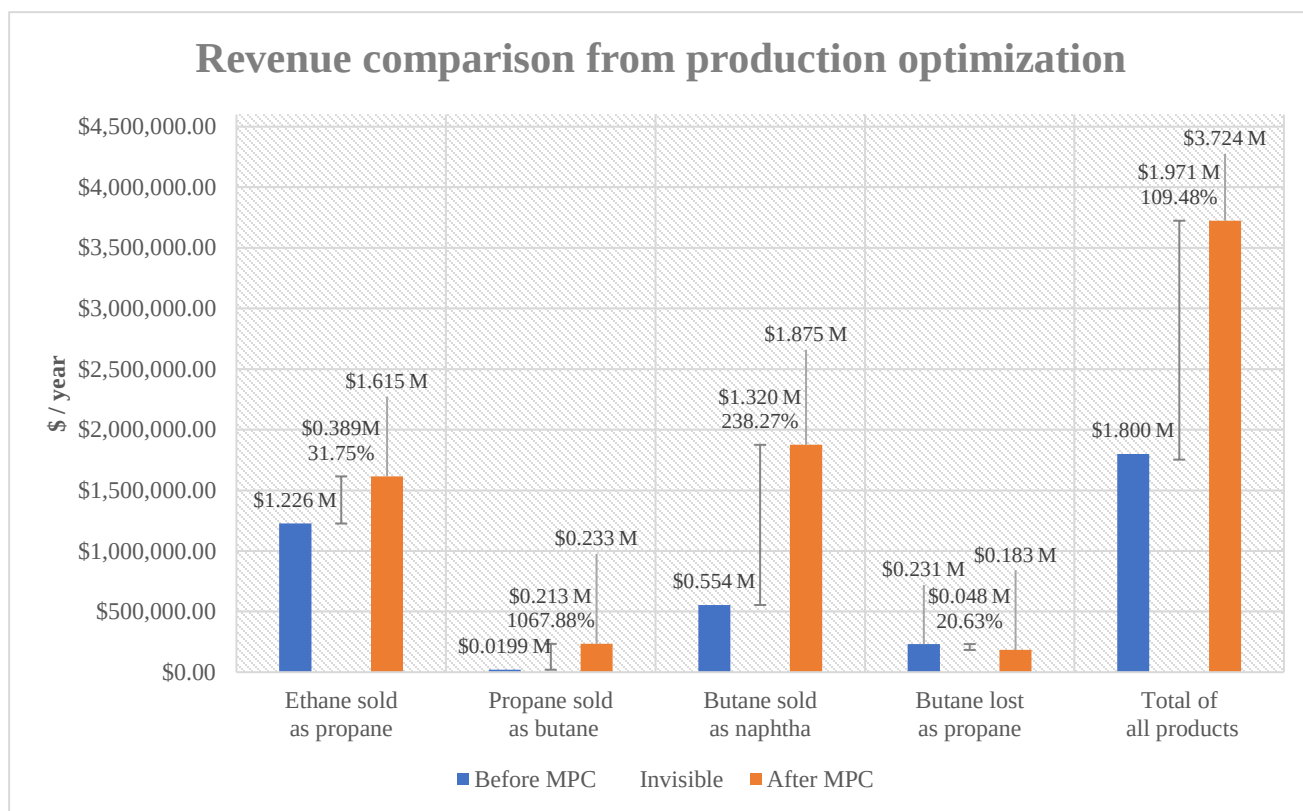


Figure 73 Revenue comparison - Production optimization

Figure 73 shows the comparative incremental increase in revenue estimated from the improvement in operational performance due to continuously optimized production. The total increase in revenue arising from improvement in production across all 3 products is estimated to be \$1,971,565 annually which represents an increase by 109.48% (across the scenarios considered) compared to the baseline years of 2019. The greatest contributor to this incremental revenue is identified to be from the improvement of the RVP quality of naphtha generating an additional \$1,320,998 annually. It is worth noting that even though the incremental

ethane quantity sold as propane is estimated to be 11,093 barrels annually, it generates an additional annual revenue of \$389,426. On the other hand, the incremental propane quantity sold as butane is estimated to be 37,844 barrels annually which is 3.4 times the quantity of incremental ethane barrels but only generates an additional annual revenue of \$213,484. This is due to the scenario difference in terms of opportunity lose. Ethane is considered an impurity and is not sold as one of the plants products. Hence any increase in propane production as result of additional slippage of ethane would contribute to revenue increase corresponding to a differential price equivalent to the price of propane which is taken as \$434.58/MT in the economic analysis basis. However, increasing butane production by slipping more propane does not assume an incremental revenue corresponding to the differential price equivalent to the price of butane. Increasing butane production by routing propane from one product portfolio to another will generate incremental income corresponding to differential price of \$7.1/MT which is the price difference between propane and butane. The calculation also considers density differences between propane and butane product which accounts to lesser volume of propane is required to make a metric ton of butane product compared to that of propane product.

As for recovered butane from propane product, the baseline data of the plant for the year 2019 shows that around 40,951 barrels of butane was sold along with the propane product. A reduction improvement of 8,448 barrels is expected which represents a reduction of loss by 20.63% thereby contributing an additional annual revenue of \$47,656.

6.5. MPC system cost and payback analysis

A payback analysis shall be conducted in order to evaluate the feasibility of installing an MPC system on the subject plant. For this, a cost estimate for all necessary equipment, software licenses and engineering services has been conducted. Table 14 summarizes an approximated cost breakup of the project as per the market prices.

Table 14 Cost breakup - MPC system implementation for the subject plant.

Sr. No.	Cost item	Cost USD	Remark
1	Software license	\$97,500.00	License cost per year including software maintenance and support
2	MPC implementation engineering services	\$291,000.00	One time service cost for system installation, configuration and tuning.
3	MPC hardware and communication server	\$50,000.00	Initial one-time cost for necessary hardware and servers.
Total		\$438,500.00	

1. Commercial costs obtained from a leading MPC technology supplier

The costs have been obtained after interacting with leading MPC technology suppliers and is specific to the subject plant requirements.

A discount rate of 5% which is the plant's Minimum Attractive Rate of Return MARR shall be considered. The project implementation typically takes around 7 months to 1 year and hence no benefit is expected to be attained in the initial year. Table 15 provides the payback analysis.

Table 15 Payback analysis - Discount rate at 5%

Year	Revenue	Cost \$	Cash flow \$	Discounted cash flow \$	Cumulative discounted cash flow \$
0	\$0.00	-\$438,500.00	-\$438,500.00	(\$438,500.00)	-\$438,500.00
1	\$1,971,565.78	\$97,500.00	\$1,874,065.78	\$1,784,824.55	\$1,346,324.55
2	\$1,971,565.78	\$97,500.00	\$1,874,065.78	\$1,699,832.91	\$3,046,157.46
3	\$1,971,565.78	\$97,500.00	\$1,874,065.78	\$1,618,888.48	\$4,665,045.94
4	\$1,971,565.78	\$97,500.00	\$1,874,065.78	\$1,541,798.55	\$6,206,844.49
5	\$1,971,565.78	\$97,500.00	\$1,874,065.78	\$1,468,379.57	\$7,675,224.06

From the analysis, it is estimated that the project returns back its initial investment and a desired return (at 5% MARR) at 1.25 years after the initial investment has been made. This further solidifies the project and makes the option of implementing an MPC system very lucrative.

6.6. Conclusions

The benefit estimation for the different application of the MPC system on the plants profitability has been demonstrated. It has been estimated that an improved control system based on constant parameter optimization can result in a total annual incremental revenue of \$1,971,565 of which the greatest contributor to this incremental revenue comes from the improvement in the RVP quality of the naphtha production which results in yearly additional revenue of \$1,320,998 – around 67% of the total estimated benefit. Increasing profit margin of the propane product by constantly maintaining as high as permissible ethane slippage results in an annual incremental revenue of \$389,426 while that achieved from increasing propane slippage into butane yields an additional yearly revenue of \$213,484. Improvement arising from reduction of butane sold as propane product yields an additional annual revenue of \$47,656 which is basically a reduction in lost butane production (lost to propane) contributing to additional butane product.

Costs of the necessary equipment, software and engineering services required to implement an MPC system on the subject plant have been obtained from a leading MPC technology supplier after having conducting site surveys and preparing a high-level estimation of the different requirements for the subject plant. A total initial investment of \$485,000 is required with a yearly license software fee of \$97,500. A payback analysis was conducted with a discount rate of 5% and found that the system pays back its initial investment at approximately 1.25 years after the initial investment hence presenting a very lucrative investment.

Chapter 7- Conclusion, recommendations, and future work

7.1 Conclusion

7.2 Recommendations

7.3 Future work

7.1. Conclusion

This research addresses the missing opportunities that hinder the economic profitability of a local NGL plant. This is largely attributed to the lack of an advanced control system that can consider the significant process interactions while systematically optimizing the setpoints to the existing regulatory control system, i.e., multi-loop controllers in the DCS. To answer the objective questions of the research, three necessary levels of simulations of the subject plant were developed. A steady state simulation using Aspen HYSYS® was first built using plants' equipment and process data, which was later verified using official plant references. In order to capture the plant's transient state behavior, a dynamic model of the plant using Aspen HYSYS Dynamics was built and later verified by comparing plant record data. Finally, a layer of MPC was designed over the dynamic model for which a functional design of the MPC was carried out in order to demonstrate and study its effect on the performance of the fractionation train. An economic quantification of the potential benefits was presented along with the payback analysis for the possible implementation of the project.

A conclusion on the thesis contribution shall be presented below.

7.1.1. Dynamic model of the plant

The Dynamic model of the plant was developed utilizing the steady state simulation model as a pre-requisite. The dynamic model serves to capture the transient behavior of the subject plant. The developed dynamic model consisting of the – Deethanizer unit, Rich oil still unit, Depropanizer unit

and Debutanizer unit was subjected to scenarios derived from the plants records in order to compare responses and behavior for which a good representation of the real dynamics has been demonstrated. The validation of the dynamic model is an important step in order to realize faithfully the merits of the proposed MPC. The latter is of significant importance, as this will ensure the realism of the estimated benefits for the subject NGL plant. While building a validated dynamic simulation is a curial step in the proposed workflow for demonstrating the benefits of MPC, its capability to mirror real processes transient behavior render it as a digital twin of the plant, which in itself has multiple benefits, to the operation and troubleshooting of the plant.

7.1.2. The impact of MPC on the plant's performance

A functional design of the MPC on the plant's main profit contributing units has been carried out and presented in Chapter 5. The effect of the MPC on the plant's performance has been demonstrated and the results indicate a superior ability in achieving the objectives of the different units. The results indicate that the MPC can reduce quality variations in different products and hence can push the operating level further to the constraint limit, from where additional benefits can be realized. This has been demonstrated specially for the ethane content in propane product, propane content in butane product and naphtha RVP. The MPC simulations have also been subjected to real life scenarios derived from the plants experiences, which include changes in feed rates and compositions. The simulations indicate the MPC outperforms the existing regulatory control in terms of maintaining the required objective of the

different units and hence provides an economical method of operation due to the employment of a prediction model that is used to decide the optimal course of action.

7.1.3. Economic benefit analysis

An economic analysis was conducted to get insight of the extent of benefit that can be achieved from implementing a MPC on the subject plant. First, average contract prices for the year 2019 were presented which would serve as a baseline. Quantification of the estimated benefits was later carried out by means of two methods: 1) employment of the “best operator” technique, that is, reviewing the respective historical operation data, 2) by utilizing simulation results after implementing the MPC, whenever the former is deemed inapplicable. The analysis concludes that an estimated incremental annual revenue of \$1,971,565 is achievable. The greatest contribution arises from the improvement in the naphtha RVP, which is around 67% of the identified increment revenue, while improvement arising from increased ethane in propane product, propane in butane product and recovered butane from propane product contribute each around 19.75%, 10.83% and 2.42%, respectively. To explore the feasibility of implementing MPC on the subject plant, a cost estimation was conducted after having approached leading MPC technology providers. Preliminary costing of the required hardware and software for the subject plant indicate that \$438,500 of initial investment is required along with an annual software license fee of \$97,500. The payback analysis conducted at 5% MARR indicate that the system will yield back its investment at 1.25 years after the initial payment, which proves very lucrative.

7.2. Recommendations and future work

On the basis of the findings and results presented in this thesis, the following shall serve as recommendations –

1. Implementing of the MPC above the regulatory layer of the subject plant will enhance profitability by its superior ability to achieve the economic objectives for the units studied. This has been demonstrated extensively in Chapter 5 and 6.
2. The MPC as has been demonstrated in Chapter 5, has the ability to meet target quality more consistently and reduces product quality even at heavy disturbances. This will ultimately lead to reduction in off specification production.
3. As cited by Forbes et al., 2015 and observed in the building of the simulation models used for the thesis, poor prediction models can have a negative impact on the performance of the MPC and hence the operation of the unit. It is vital that the step testing of the plant should be done carefully in order to have a complete picture of the process interactions. Judicious selection of the MVs and CVs is crucial to implement a successful MPC design. The functional design of the MPC presented in section 4.4 can serve as guideline in this regard.
4. The subject NGL plant has been observed actively changing plant and process conditions such as debottlenecking and increasing process capacities. Such changes may result in a degrading MPC performance due to poor prediction as the model no longer provides a good representation of the system. It is vital that process models are updated regularly to reflect the plants process and enhance maintain benefits.

5. It is recommended to establish an appropriate MPC monitoring system to evaluate the performance of the system post implementation. This would ensure that any degradation of the MPC performance due to model-plant mismatch would be caught and rectified thereby maintaining the benefits achieved by the system.
6. The dynamic models developed as part of the thesis' work have been seen to exhibit good representation of the real plant and as such renders itself as a digital twin of the plant. Such models can be employed to explore process troubleshooting operations, plant scenario testing, revamping studies and review of startup, shutdown and safety procedures. This would prove beneficial for the subject plant given that it does not have such dynamic models available.
7. The MPC simulation and consequent work was carried out with the objective of maximizing economic benefits. However other objectives can be explored concerning safety and environmental aspects. The scope of work could further be expanded to include utility section of the plant where most of the environmental emissions are released. The potential of reducing environmental emissions can be studied.
8. Further work on the integration of the MPC layer with the upper RTO layer which is the plant wide and individual unit optimizer layer thereby providing the directions to be achieved by the MPC.
9. An economic analysis justifies the investment required for such projects. Available literature on the benefit estimation techniques mainly revolves around rule of thumbs or estimation techniques. More rigorous financial models could be explored in order to quantify economic benefits resulting from MPC improvements.

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Appendix

A-1 Details for the Steady State simulation

A-1.1 Feed gas condition – Basis of simulation

Gas flowrate – 161.4 MMSCFD

Table 16 Feed gas composition – Design basis for the steady state simulation

Components	Mol%
Hydrogen	1.52
Nitrogen	10.68
Hydrogen Sulphide	0.06
Carbon dioxide	6.12
H ₂ O	0.30
Methane	72.28
Ethane	2.89
Propane	2.47
i-Butane	0.84
n-Butane	1.39
i-Pentane	0.35
n-Pentane	0.44
n-Hexane	0.32
n-Heptane	0.19
n-Octane	0.12
n-Nonane	0.02

Note

1. Composition as per plant design feed case.

A-1.2 Column details – Steady state simulation requirement

Table 17 Plant column details – Steady state simulation basis

Column	Number of stages	Column type	Top (Temp., F / Pressure, psia)	Bottom (Temp., F/ Pressure, psia)	Feed entry stage
204C-002 Absorber	24	Absorber	-20.58 / 384	-29.4 / 385	Semi lean oil – 24 th stage Gas feed – 1 st stage
204C-003 Deethanizer	24	Re-boiled absorber	26 / 410	349 / 415	Top feed – 24 th stage Preheated feed – 16 th Stage
204C-005 Rich oil still	36	Distillation column	209.6 / 210	520 / 215	25 th stage
204C-006 Depropanizer	42	Distillation column	145 / 300	275 / 315	Cold feed – 27 th stage Hot feed – 9 th stage
204C-007 Debutanizer	42	Distillation column	144 / 105	256 / 110	24 th stage

A-1.3 Heat Exchanger details

Table 18 Plant heat exchanger details - Steady state simulation basis

Heat exchanger	Duty, MMBtu/hr	Temperature (In / Out), F
204E-003 Cold gas/gas HE	9.44	59.96 / 14.90
204E-004 Inlet chiller	12.87	11 / -32
Pre-saturator chiller, 204E-005	3.38	-13.97 / -30
Absorber bottom HE, 204E-006	7.56	-28.32 / 127
Deethanizer feed heater, 204E-500	8.32	127 / 273

Deethanizer reboiler, 204E-008	18.60	282 / 349
Rich oil still condenser, 204E-052	15.96	209 / 161
Depropanizer reboiler, 204E-009	9.85	275 / 298
Depropanizer condenser, 204E-053	9.21	145 / 142
Debutanizer reboiler, 204E-011	7.00	256 / 283
Debutanizer condenser, 204E-054	9.62	144 / 141

A-2 Details for the Dynamic Simulation model

A-2.1 Column details – Dynamic simulation requirement

Table 19 Column details - Dynamic simulation

Column	Column geometry and internal
204C-002 Absorber	Internal diameter – 2.439 m 24 Flex type valve trays – 1 pass trays
204C-003 Deethanizer	Top internal diameter – 1.372 m 8 flex type valve trays – 1 pass Bottom internal diameter – 1.982 m 16 flex type valve trays – 2 pass
204C-005 Rich oil still	Top internal diameter – 1.829 m 18 flex type valve trays – 2 pass Bottom internal diameter – 2.744 m 18 flex type valve trays – 2 pass
204C-006 Depropanizer	Internal diameter – 1.829 m 42 flex type valve trays – 1 pass
204C-007 Debutanizer	Internal diameter – 1.524 m 42 flex type valve trays – 1 pass

A-2.2 PID loop tuning parameters

Table 20 Dynamic simulation controller tuning parameters

Controller	Kc	Ti, mins	Td, mins
FIC- 100	0.107	0.0733	0.0052
TIC 204E-003	0.138	0.367	-
LIC-031	2	-	-
FIC-018	0.426	0.0733	-
TIC 204E-005	0.226	0.0367	-
PIC-186A	5.86	0.0733	-
FIC-014	0.327	0.0674	-
LIC-023	2	-	-
FIC-034	4.78	0.0733	0.0059
TIC DEC2 reboiler	7.04	0.348	-
LIC-036	2	-	-
PIC-056	2	10	-
PIC-031	2	10	-
TIC-086	5.50	0.642	-
LV-032	2	-	-
TIC-115	7.86	4.03	0.291
FIC-052	0.456	0.0733	-
PIC-086	2	10	-
PIC-089	2	10	-
QIC DEC3 Reboiler	0.527	0.0733	0.00529
LIC-051	2	-	-
LIC-048	2	-	-
TIC DEC4 Reboiler	0.407	0.0594	-
TIC 128 DEC4	6.89	4.40	-

LIC-055	18.1	0.807	-
LIC-053	2	1.04	-

A-3 Heat and Mass Balance – Steady State simulation

Table 21 Heat and mass balance - Steady state simulation

	<i>Unit</i>	INLET GAS FW 2011	109	1	21**	22	deC2 upper feed	deC2 lower feed
Vapour Fraction		1.0000	0.8983	0.0000	1.0000	0.0000	0.0000	0.4825
Temperature	<i>F</i>	59.30	-32.00	-29.22	-20.58	-29.43	-28.32	273.00
Pressure	<i>psia</i>	408.00	385.00	384.00	384.00	385.00	398.00	388.30
Molar Flow	<i>Kmole/hr</i>	7954.00	8778.00	532.20	7731.00	1094.00	547.10	547.10
Mass Flow	<i>lb/hr</i>	374525.25	465146.00	103921.33	336873.93	160934.01	80467.01	80467.01
Liquid Volume Flow	<i>barrel/day</i>	65358.66	76491.01	10671.28	61150.97	18028.53	9014.27	9014.27
Heat Flow	<i>Btu/hr</i>	-6.73E+08	-8.25E+08	-1.15E+08	-6.60E+08	-1.95E+08	-9.74E+07	-8.15E+07
Components								
Comp Mole Frac (Hydrogen)		0.0152	0.0140	0.0003	0.0159	0.0003	0.0003	0.0003
Comp Mole Frac (H2S)		0.0006	0.0008	0.0016	0.0007	0.0019	0.0019	0.0019
Comp Mole Frac (CO2)		0.0612	0.0662	0.0539	0.0674	0.0561	0.0561	0.0561
Comp Mole Frac (Nitrogen)		0.1068	0.0989	0.0052	0.1115	0.0057	0.0057	0.0057
Comp Mole Frac (Methane)		0.7228	0.6921	0.1538	0.7625	0.1658	0.1658	0.1658
Comp Mole Frac (Ethane)		0.0289	0.0386	0.0564	0.0337	0.0683	0.0683	0.0683
Comp Mole Frac (Propane)		0.0247	0.0320	0.0477	0.0082	0.1539	0.1539	0.1539
Comp Mole Frac (i-Butane)		0.0084	0.0107	0.0000	0.0000	0.0597	0.0597	0.0597
Comp Mole Frac (n-Butane)		0.0139	0.0172	0.0000	0.0000	0.0954	0.0954	0.0954
Comp Mole Frac (i-Pentane)		0.0035	0.0038	0.0000	0.0000	0.0212	0.0212	0.0212
Comp Mole Frac (n-Pentane)		0.0044	0.0045	0.0000	0.0000	0.0248	0.0248	0.0248
Comp Mole Frac (n-Hexane)		0.0032	0.0022	0.0019	0.0000	0.0127	0.0127	0.0127
Comp Mole Frac (n-Heptane)		0.0019	0.0058	0.1918	0.0001	0.0965	0.0965	0.0965
Comp Mole Frac (n-Octane)		0.0012	0.0088	0.3223	0.0000	0.1572	0.1572	0.1572
Comp Mole Frac (n-Nonane)		0.0002	0.0010	0.0359	0.0000	0.0175	0.0175	0.0175
Comp Mole Frac (n-Decane)		0.0000	0.0026	0.0954	0.0000	0.0464	0.0464	0.0464
Comp Mole Frac (H2O)		0.0030	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (TEGlycol)		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (n-C11)		0.0000	0.0008	0.0282	0.0000	0.0137	0.0137	0.0137
Comp Mole Frac (n-C12)		0.0000	0.0002	0.0056	0.0000	0.0027	0.0027	0.0027

	<i>Unit</i>	14	28	ROS OUTLET.	29	TO STAGE 9 DEC3	C3 R/D	TO TRAY 24 DEC4
Vapour Fraction		1.0000	0.3523	0.0000	0.0000	0.3172	0.0000	0.5786
Temperature	<i>F</i>	26.06	317.36	519.00	162.53	402.43	143.68	212.16
Pressure	<i>psia</i>	410.00	211.90	250.00	316.40	317.80	357.00	106.60
Molar Flow	<i>Kmole/hr</i>	339.20	755.00	363.00	392.10	94.15	166.80	319.50
Mass Flow	<i>lb/hr</i>	19350.03	141583.98	93691.83	47892.15	18184.20	16215.83	49860.52
Liquid Volume Flow	<i>barrel/day</i>	3147.63	14880.91	9054.66	5826.25	1883.63	2193.46	5516.42
Heat Flow	<i>Btu/hr</i>	-4.40E+07	-1.16E+08	-5.98E+07	-5.07E+07	-1.37E+07	-1.82E+07	-4.50E+07
Components								
Comp Mole Frac (Hydrogen)		0.0010	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (H2S)		0.0048	0.0007	0.0000	0.0013	0.0000	0.0030	0.0000
Comp Mole Frac (CO2)		0.1809	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Nitrogen)		0.0183	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Methane)		0.5350	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (Ethane)		0.2138	0.0030	0.0000	0.0058	0.0000	0.0136	0.0000
Comp Mole Frac (Propane)		0.0348	0.2074	0.0000	0.3995	0.0511	0.9671	0.0005
Comp Mole Frac (i-Butane)		0.0049	0.0843	0.0000	0.1623	0.0514	0.0153	0.2064
Comp Mole Frac (n-Butane)		0.0054	0.1359	0.0000	0.2617	0.1200	0.0010	0.3561
Comp Mole Frac (i-Pentane)		0.0005	0.0306	0.0000	0.0588	0.0673	0.0000	0.0921
Comp Mole Frac (n-Pentane)		0.0004	0.0358	0.0000	0.0690	0.1076	0.0000	0.1164
Comp Mole Frac (n-Hexane)		0.0001	0.0183	0.0026	0.0329	0.1721	0.0000	0.0911
Comp Mole Frac (n-Heptane)		0.0002	0.1398	0.2814	0.0086	0.1774	0.0000	0.0628
Comp Mole Frac (n-Octane)		0.0001	0.2278	0.4737	0.0000	0.1803	0.0000	0.0532
Comp Mole Frac (n-Nonane)		0.0000	0.0253	0.0526	0.0000	0.0566	0.0000	0.0167
Comp Mole Frac (n-Decane)		0.0000	0.0673	0.1400	0.0000	0.0162	0.0000	0.0048
Comp Mole Frac (H2O)		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (TEGlycol)		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (n-C11)		0.0000	0.0199	0.0414	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (n-C12)		0.0000	0.0040	0.0082	0.0000	0.0000	0.0000	0.0000

	<i>Unit</i>	BUTANE R/D	NAPHTHA R/D	CONDENSATE TO 4C-006	RESIDUE GAS	LEAN OIL	LEAN OIL TO PREABSORBER
Vapour Fraction		0.0000	0.0000	0.0000	1.0000	0.0000	0.0000
Temperature	<i>F</i>	141.63	283.18	423.40	-30.00	525.81	-15.00
Pressure	<i>psia</i>	105.00	115.00	425.00	375.00	430.00	420.00
Molar Flow	<i>Kmole/hr</i>	184.60	134.90	94.15	7561	362.20	362.2
Mass Flow	<i>lb/hr</i>	23792.53	26067.99	18184.20	326394.03	93498.80	93498.80
Liquid Volume Flow	<i>barrel/day</i>	2825.40	2691.02	1883.63	59507.53	9036.00	9036.00
Heat Flow	<i>Btu/hr</i>	-2.55E+07	-2.26E+07	-1.37E+07	-6.39E+08	-5.95E+07	-9.12E+07
Components							
Comp Mole Frac (Hydrogen)		0.0000	0.0000	0.0000	0.0162	0.0000	0.0000
Comp Mole Frac (H2S)		0.0000	0.0000	0.0000	0.0006	0.0000	0.0000
Comp Mole Frac (CO2)		0.0000	0.0000	0.0000	0.0651	0.0000	0.0000
Comp Mole Frac (Nitrogen)		0.0000	0.0000	0.0000	0.1136	0.0000	0.0000
Comp Mole Frac (Methane)		0.0000	0.0000	0.0000	0.7689	0.0000	0.0000
Comp Mole Frac (Ethane)		0.0000	0.0000	0.0000	0.0305	0.0000	0.0000
Comp Mole Frac (Propane)		0.0008	0.0000	0.0511	0.0050	0.0000	0.0000
Comp Mole Frac (i-Butane)		0.3572	0.0000	0.0514	0.0000	0.0000	0.0000
Comp Mole Frac (n-Butane)		0.6163	0.0000	0.1200	0.0000	0.0000	0.0000
Comp Mole Frac (i-Pentane)		0.0251	0.1837	0.0673	0.0000	0.0000	0.0000
Comp Mole Frac (n-Pentane)		0.0006	0.2748	0.1076	0.0000	0.0000	0.0000
Comp Mole Frac (n-Hexane)		0.0000	0.2159	0.1721	0.0000	0.0026	0.0026
Comp Mole Frac (n-Heptane)		0.0000	0.1488	0.1774	0.0000	0.2814	0.2814
Comp Mole Frac (n-Octane)		0.0000	0.1260	0.1803	0.0000	0.4737	0.4737
Comp Mole Frac (n-Nonane)		0.0000	0.0395	0.0566	0.0000	0.0526	0.0526
Comp Mole Frac (n-Decane)		0.0000	0.0113	0.0162	0.0000	0.1400	0.1400
Comp Mole Frac (H2O)		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (TEGlycol)		0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Comp Mole Frac (n-C11)		0.0000	0.0000	0.0000	0.0000	0.0414	0.0414
Comp Mole Frac (n-C12)		0.0000	0.0000	0.0000	0.0000	0.0082	0.0082

A-4 Step test data

A-4.1 Deethanizer column

3

3

4-15-2021 18:1:31

60

121

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C2INC3CONV~~~C2INC3CONV~~~C2~~~

TICREBOILE~~~TICREBOILE~~~SP~~~kW

LIC-036~~~LIC-036~~~SP~~~%

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4-15-2021 18:1:31,1.952679,G,5170.000000,G,50.000000,G
4-15-2021 18:2:31,1.943226,G,5370.000000,G,53.000000,G
4-15-2021 18:3:31,1.886230,G,5370.000000,G,53.000000,G
4-15-2021 18:4:31,1.845441,G,5370.000000,G,53.000000,G
4-15-2021 18:5:31,1.809718,G,5370.000000,G,53.000000,G
4-15-2021 18:6:31,1.766778,G,5370.000000,G,53.000000,G
4-15-2021 18:7:31,1.741891,G,5370.000000,G,53.000000,G
4-15-2021 18:8:31,1.719627,G,5370.000000,G,53.000000,G
4-15-2021 18:9:31,1.708170,G,5370.000000,G,53.000000,G
4-15-2021 18:10:31,1.688273,G,5370.000000,G,53.000000,G
4-15-2021 18:11:31,1.666469,G,5370.000000,G,53.000000,G
4-15-2021 18:12:31,1.650386,G,5370.000000,G,53.000000,G
4-15-2021 18:13:31,1.640375,G,5370.000000,G,53.000000,G
4-15-2021 18:14:31,1.623742,G,5370.000000,G,53.000000,G
4-15-2021 18:15:31,1.597584,G,5370.000000,G,53.000000,G
4-15-2021 18:16:31,1.582814,G,5370.000000,G,53.000000,G
4-15-2021 18:17:31,1.563044,G,5370.000000,G,53.000000,G
4-15-2021 18:18:31,1.539743,G,5370.000000,G,53.000000,G
4-15-2021 18:19:31,1.517857,G,5370.000000,G,53.000000,G
4-15-2021 18:20:31,1.495167,G,5370.000000,G,53.000000,G
4-15-2021 18:21:31,1.476524,G,5370.000000,G,53.000000,G
4-15-2021 18:22:31,1.467100,G,5370.000000,G,53.000000,G
4-15-2021 18:23:31,1.454769,G,5370.000000,G,53.000000,G
4-15-2021 18:24:31,1.431280,G,5370.000000,G,53.000000,G
4-15-2021 18:25:31,1.414444,G,5370.000000,G,53.000000,G
4-15-2021 18:26:31,1.398475,G,5370.000000,G,53.000000,G
4-15-2021 18:27:31,1.371501,G,5370.000000,G,53.000000,G
4-15-2021 18:28:31,1.352687,G,5370.000000,G,53.000000,G
4-15-2021 18:29:31,1.339060,G,5370.000000,G,53.000000,G

4-15-2021 18:30:31,1.324728,G,5370.000000,G,53.000000,G
4-15-2021 18:31:31,1.308940,G,5170.000000,G,53.000000,G
4-15-2021 18:32:31,1.332573,G,5170.000000,G,53.000000,G
4-15-2021 18:33:31,1.353585,G,5170.000000,G,53.000000,G
4-15-2021 18:34:31,1.369535,G,5170.000000,G,53.000000,G
4-15-2021 18:35:31,1.375975,G,5170.000000,G,53.000000,G
4-15-2021 18:36:31,1.374731,G,5170.000000,G,53.000000,G
4-15-2021 18:37:31,1.368414,G,5170.000000,G,53.000000,G
4-15-2021 18:38:31,1.365611,G,5170.000000,G,53.000000,G
4-15-2021 18:39:31,1.364715,G,5170.000000,G,53.000000,G
4-15-2021 18:40:31,1.365179,G,5170.000000,G,53.000000,G
4-15-2021 18:41:31,1.363731,G,5170.000000,G,50.000000,G
4-15-2021 18:42:31,1.379041,G,5170.000000,G,50.000000,G
4-15-2021 18:43:31,1.370832,G,5170.000000,G,50.000000,G
4-15-2021 18:44:31,1.374346,G,5170.000000,G,50.000000,G
4-15-2021 18:45:31,1.355274,G,5170.000000,G,50.000000,G
4-15-2021 18:46:31,1.341270,G,5170.000000,G,50.000000,G
4-15-2021 18:47:31,1.348289,G,5170.000000,G,50.000000,G
4-15-2021 18:48:31,1.363988,G,5170.000000,G,50.000000,G
4-15-2021 18:49:31,1.359065,G,5170.000000,G,50.000000,G
4-15-2021 18:50:31,1.357135,G,5170.000000,G,50.000000,G
4-15-2021 18:51:31,1.360368,G,5170.000000,G,50.000000,G
4-15-2021 18:52:31,1.353201,G,5170.000000,G,50.000000,G
4-15-2021 18:53:31,1.358211,G,5170.000000,G,50.000000,G
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4-15-2021 18:57:31,1.350750,G,5170.000000,G,50.000000,G
4-15-2021 18:58:31,1.345489,G,5170.000000,G,50.000000,G
4-15-2021 18:59:31,1.345503,G,5170.000000,G,50.000000,G
4-15-2021 19:0:31,1.345018,G,5170.000000,G,50.000000,G
4-15-2021 19:1:31,1.355564,G,4970.000000,G,50.000000,G
4-15-2021 19:2:31,1.386828,G,4970.000000,G,50.000000,G
4-15-2021 19:3:31,1.419002,G,4970.000000,G,50.000000,G
4-15-2021 19:4:31,1.447430,G,4970.000000,G,50.000000,G
4-15-2021 19:5:31,1.479187,G,4970.000000,G,50.000000,G
4-15-2021 19:6:31,1.497177,G,4970.000000,G,50.000000,G
4-15-2021 19:7:31,1.525948,G,4970.000000,G,50.000000,G
4-15-2021 19:8:31,1.549762,G,4970.000000,G,50.000000,G
4-15-2021 19:9:31,1.564546,G,4970.000000,G,50.000000,G
4-15-2021 19:10:31,1.584282,G,4970.000000,G,50.000000,G
4-15-2021 19:11:31,1.608741,G,4970.000000,G,47.000000,G
4-15-2021 19:12:31,1.640232,G,4970.000000,G,47.000000,G

4-15-2021 19:13:31,1.657389,G,4970.000000,G,47.000000,G
4-15-2021 19:14:31,1.683910,G,4970.000000,G,47.000000,G
4-15-2021 19:15:31,1.693725,G,4970.000000,G,47.000000,G
4-15-2021 19:16:31,1.721688,G,4970.000000,G,47.000000,G
4-15-2021 19:17:31,1.748643,G,4970.000000,G,47.000000,G
4-15-2021 19:18:31,1.784064,G,4970.000000,G,47.000000,G
4-15-2021 19:19:31,1.814444,G,4970.000000,G,47.000000,G
4-15-2021 19:20:31,1.830640,G,4970.000000,G,47.000000,G
4-15-2021 19:21:31,1.855911,G,4970.000000,G,47.000000,G
4-15-2021 19:22:31,1.895372,G,4970.000000,G,47.000000,G
4-15-2021 19:23:31,1.928573,G,4970.000000,G,47.000000,G
4-15-2021 19:24:31,1.948076,G,4970.000000,G,47.000000,G
4-15-2021 19:25:31,1.963852,G,4970.000000,G,47.000000,G
4-15-2021 19:26:31,2.002542,G,4970.000000,G,47.000000,G
4-15-2021 19:27:31,2.045052,G,4970.000000,G,47.000000,G
4-15-2021 19:28:31,2.087026,G,4970.000000,G,47.000000,G
4-15-2021 19:29:31,2.107508,G,4970.000000,G,47.000000,G
4-15-2021 19:30:31,2.124092,G,4970.000000,G,47.000000,G
4-15-2021 19:31:31,2.154010,G,5170.000000,G,47.000000,G
4-15-2021 19:32:31,2.131653,G,5170.000000,G,47.000000,G
4-15-2021 19:33:31,2.133772,G,5170.000000,G,47.000000,G
4-15-2021 19:34:31,2.119050,G,5170.000000,G,47.000000,G
4-15-2021 19:35:31,2.121493,G,5170.000000,G,47.000000,G
4-15-2021 19:36:31,2.122707,G,5170.000000,G,47.000000,G
4-15-2021 19:37:31,2.124196,G,5170.000000,G,47.000000,G
4-15-2021 19:38:31,2.140384,G,5170.000000,G,47.000000,G
4-15-2021 19:39:31,2.137539,G,5170.000000,G,47.000000,G
4-15-2021 19:40:31,2.131936,G,5170.000000,G,47.000000,G
4-15-2021 19:41:31,2.140059,G,5170.000000,G,50.000000,G
4-15-2021 19:42:31,2.136766,G,5170.000000,G,50.000000,G
4-15-2021 19:43:31,2.150386,G,5170.000000,G,50.000000,G
4-15-2021 19:44:31,2.154632,G,5170.000000,G,50.000000,G
4-15-2021 19:45:31,2.157030,G,5170.000000,G,50.000000,G
4-15-2021 19:46:31,2.153626,G,5170.000000,G,50.000000,G
4-15-2021 19:47:31,2.157970,G,5170.000000,G,50.000000,G
4-15-2021 19:48:31,2.186296,G,5170.000000,G,50.000000,G
4-15-2021 19:49:31,2.191305,G,5170.000000,G,50.000000,G
4-15-2021 19:50:31,2.199851,G,5170.000000,G,50.000000,G
4-15-2021 19:51:31,2.207643,G,5170.000000,G,50.000000,G
4-15-2021 19:52:31,2.208502,G,5170.000000,G,50.000000,G
4-15-2021 19:53:31,2.216340,G,5170.000000,G,50.000000,G
4-15-2021 19:54:31,2.218529,G,5170.000000,G,50.000000,G
4-15-2021 19:55:31,2.225859,G,5170.000000,G,50.000000,G

4-15-2021 19:56:31,2.224397,G,5170.000000,G,50.000000,G
 4-15-2021 19:57:31,2.229058,G,5170.000000,G,50.000000,G
 4-15-2021 19:58:31,2.229707,G,5170.000000,G,50.000000,G
 4-15-2021 19:59:31,2.233916,G,5170.000000,G,50.000000,G
 4-15-2021 20:0:31,2.235768,G,5170.000000,G,50.000000,G
 4-15-2021 20:1:31,2.239957,G,5170.000000,G,50.000000,G

A-4.2 Depropanizer column

4

4

4-19-2021 18:15:0

60

62

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TOPPRODUCT~~~TOPPRODUCT~~~Master Comp Volume Frac (Propane)~~~

C3inC4conv~~~C3inC4conv~~~A5:~~~

QICREBOILE~~~QICREBOILE~~~SP~~~kW

TIC-115DEC~~~TIC-115DEC~~~SP~~~F

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4-19-2021 18:15:0,0.957464,G,1.871718,G,1309.938992,G,131.631914,G
 4-19-2021 18:16:0,0.957432,G,1.890755,G,1389.938992,G,134.091914,G
 4-19-2021 18:17:0,0.957379,G,1.855522,G,1389.938992,G,134.091914,G
 4-19-2021 18:18:0,0.957305,G,1.810941,G,1389.938992,G,134.091914,G
 4-19-2021 18:19:0,0.957225,G,1.762782,G,1389.938992,G,134.091914,G
 4-19-2021 18:20:0,0.957145,G,1.732574,G,1389.938992,G,134.091914,G
 4-19-2021 18:21:0,0.957066,G,1.697719,G,1389.938992,G,134.091914,G
 4-19-2021 18:22:0,0.956974,G,1.670919,G,1389.938992,G,134.091914,G
 4-19-2021 18:23:0,0.956861,G,1.626559,G,1389.938992,G,134.091914,G
 4-19-2021 18:24:0,0.956740,G,1.604160,G,1389.938992,G,134.091914,G
 4-19-2021 18:25:0,0.956607,G,1.592253,G,1389.938992,G,134.091914,G
 4-19-2021 18:26:0,0.956462,G,1.586380,G,1389.938992,G,134.091914,G
 4-19-2021 18:27:0,0.956328,G,1.570962,G,1389.938992,G,134.091914,G
 4-19-2021 18:28:0,0.956204,G,1.539601,G,1389.938992,G,134.091914,G
 4-19-2021 18:29:0,0.956091,G,1.519788,G,1389.938992,G,134.091914,G
 4-19-2021 18:30:0,0.955985,G,1.507356,G,1389.938992,G,134.091914,G
 4-19-2021 18:31:0,0.955886,G,1.481884,G,1309.938992,G,134.091914,G
 4-19-2021 18:32:0,0.955797,G,1.529285,G,1309.938992,G,134.091914,G
 4-19-2021 18:33:0,0.955716,G,1.574978,G,1309.938992,G,134.091914,G
 4-19-2021 18:34:0,0.955638,G,1.593309,G,1309.938992,G,134.091914,G
 4-19-2021 18:35:0,0.955557,G,1.600043,G,1309.938992,G,134.091914,G
 4-19-2021 18:36:0,0.955471,G,1.605989,G,1309.938992,G,134.091914,G

4-19-2021 18:37:0,0.955381,G,1.617002,G,1309.938992,G,134.091914,G
4-19-2021 18:38:0,0.955283,G,1.618324,G,1309.938992,G,134.091914,G
4-19-2021 18:39:0,0.955191,G,1.617076,G,1309.938992,G,134.091914,G
4-19-2021 18:40:0,0.955113,G,1.622620,G,1309.938992,G,134.091914,G
4-19-2021 18:41:0,0.955044,G,1.609492,G,1309.938992,G,131.631911,G
4-19-2021 18:42:0,0.954993,G,1.601295,G,1309.938992,G,131.631911,G
4-19-2021 18:43:0,0.954974,G,1.603125,G,1309.938992,G,131.631911,G
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4-19-2021 18:45:0,0.955020,G,1.611757,G,1309.938992,G,131.631911,G
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A-4.3 Debutanizer column

4

4

4-22-2021 18:33:28

60

68

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BUTANEPURI~~~BUTANEPURI~~~A3~~~

NAPHTHAR/D~~~NAPHTHAR/D~~~Calculator (API 5B1.1[RVP]-Overall)~~~

TIC-128DEC~~~TIC-128DEC~~~SP~~~F

QICDEC4REB~~~DEC4REBOIL~~~SP~~~kW

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A-5 MPC tuning parameters

Table 22 MPC Tuning parameters

Units Parameters	Deethanizer unit MPC	Depropanizer unit MPC	Debutanizer unit MPC
Control interval	120	150	160
Control horizon	2	2	2
Step response length	121	62	68
Prediction horizon	25	15	15
Gamma U	1	0.5	0.5
Gamma Y	0.7	0.5	1

A-6 Aspen HYSYS MPC Block

Let $a_0, a_1, a_2, \dots, a_T$ represent values of the unit step response function obtained from open loop response.

Consider a a step response arising from a change in the input Δm . Let C_n be the actual output and $\hat{C}_n$ the predicted value of the output and m_n be the value of the manipulated variable at the n^{th} sampling time. Denoting the $\Delta m_i = m_i - m_{i-1}$, the convolution model of the single step respose function is given as follows

$$\hat{C}_{n+1} = c_0 + \sum_{i=1}^T a_i \Delta m_{n+1-i}$$

(A-1)

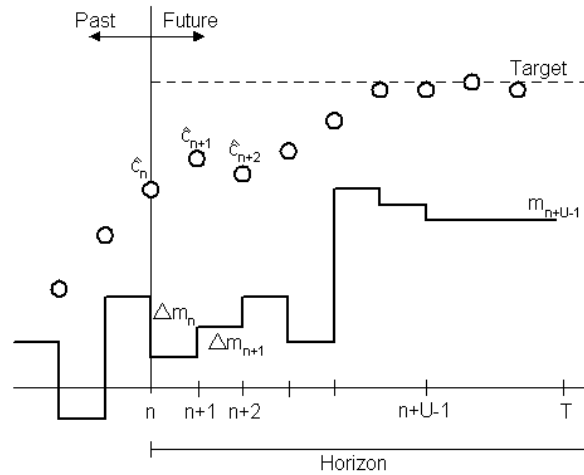


Figure 74 Model predictive control – Aspen HYSYS MPC Block

A recursive form of the model can be expressed in incremental change Δm as follows:

$$\hat{C}_{n+1} = \hat{C}_n + \sum_{i=1}^T h_i \Delta m_{n+1-i}$$

(A-2)

Where h_i is the impulse response coefficient.

To provide corrections for the influence of unmeasured loads and model errors, a corrected prediction term $\ddot{C}_{n+1}$ is introduced. The corrected value is obtained by comparing the actual value with the predicted and then shifting correction forward as follows:

$$\ddot{C}_{n+1} - \hat{C}_{n+1} = C_n - \hat{C}_n \quad (\text{A-3})$$

Substituting equation, A-3 in A-2, the following corrected prediction equation is obtained:

$$\ddot{C}_{n+1} = C_n + \sum_{i=1}^T h_i \Delta m_{n+1-i} \quad (\text{A-4})$$

Equation A-4 can be extended to include predictions for a number of future time steps hence allowing the model-based control system to anticipate future process behavior.

A prediction horizon parameter V is introduced that influences the control system performance and is expressed in terms of incremental changes in the manipulated variable:

$$\ddot{C}_{n+j} = \ddot{C}_{n+j-1} + \sum_{i=1}^T h_i \Delta m_{n+j-i} \quad (\text{A-5})$$

Where $j = 1, 2, 3, \dots, V$

Equation A-5 can be expressed in a vector matrix form as:

$$\begin{bmatrix} \ddot{C}_{n+1} \\ \ddot{C}_{n+2} \\ \ddot{C}_{n+3} \\ \vdots \\ \vdots \\ \ddot{C}_{n+V} \end{bmatrix} = \begin{bmatrix} a_1 & 0 & 0 & \dots & 0 \\ a_2 & a_1 & 0 & \dots & 0 \\ a_3 & a_2 & a_1 & \dots & 0 \\ \vdots & \vdots & \vdots & \ddots & \vdots \\ \vdots & \vdots & \vdots & \ddots & \vdots \\ a_V & a_{V-1} & a_{V-2} & \dots & a_{V-u+1} \end{bmatrix} \begin{bmatrix} \Delta m_n \\ \Delta m_{n+1} \\ \Delta m_{n+2} \\ \vdots \\ \vdots \\ \Delta m_{n+U-1} \end{bmatrix} + \begin{bmatrix} C_n + P_1 \\ C_n + P_2 \\ C_n + P_3 \\ \vdots \\ \vdots \\ C_n + P_V \end{bmatrix} \quad (\text{A-6})$$

Where -

$$a_i = \sum_{j=1}^i h_j \quad (\text{A-7})$$

$$P_i = \sum_{j=1}^i S_j \quad \text{for} \quad i = 1, 2, \dots, V \quad (\text{A-8})$$

$$S_i = \sum_{i=j+1}^T h_i \Delta m_{n+1-i} \quad \text{for} \quad j = 1, 2, \dots, V \quad (\text{A-9})$$

Denoting the set point trajectory as $r_{n+j}, j = 1, 2, \dots, V$, equation A-6 can be expressed as

$$\hat{E} = -A\Delta m + \hat{E}' \quad (\text{A-10})$$

Where –

A - the $V \times U$ triangular matrix, U being the control horizon i.e. the number of control moves calculated over the prediction horizon.

Δm – the $U \times 1$ vector of future control moves.

$\hat{E}$ and $\hat{E}'$ - the closed loop and open loop predictions respectively.

The best solution can be obtained by minimizing the following performance index

$$j[\Delta m] = \hat{E}^T \Gamma_U \hat{E} + \Delta m^T \Gamma_y \Delta m$$

(A-11)

Where Γ_U and Γ_y are positive weighting matrices for the predicted errors and control moves respectively. These are the penalties that is placed on the predicted errors resulting in a better tuned controller.

The resulting control law that minimizes j

$$\Delta m = (A^T \Gamma_U A + \Gamma_y)^{-1} A^T \Gamma_U \hat{E}'$$

(A-12)

ملخص

الهدف من أي منشئة صناعية سواء كانت تعمل في قطاع النفط أو الغاز أو البتروكيماويات هو إضافة قيمة في سلسلة العمليات من خلال تحويل المواد الخام إلى منتجات أكثر قيمة، مع الأخذ بعين الاعتبار القيود التشغيلية المختلفة على سبيل المثال لا الحصر، توافر المواد الخام، وأسعار المنتجات والسلامة والقيود البيئية.

لمواجهة هذه القيود على النحو الأمثل، يلجأ مشغلو هذه المصانع إلى استراتيجيات التحكم والاستراتيجيات التي من شأنها أن تساعد في تحسين الإنتاج وتلبية متطلبات الجودة.

تبلغ الطاقة الاستيعابية الإجمالية لمصنع الغاز الطبيعي المسال 660 مليون قدم مكعب في اليوم. يستخدم المصنع حاليًا نظامًا تقليديًا للرقابة والتحكم، وهو نظام المراقبة والتحكم التوزيعي للتحكم بالعمليات الرئيسية في المصنع. يفقر نظام التحكم الحالي إلى القدرة على استيعاب والتعامل مع القيود التشغيلية المترابطة والعلاقات المتداخلة بين دوائر التحكم ، وبالتالي يوفر حلاً دون المستوى الأمثل لمشكلة التحكم المفيدة.

تهدف هذه الأطروحة إلى استكشاف إمكانات أنظمة التحكم المبنية على نماذج التحكم كاستراتيجية تحكم مثالية ومقدمة لمعالجة بعض تحديات الجودة والإنتاج التي تواجه عمليات التشغيل حالياً في مصنع سوائل الغاز الطبيعي في مملكة البحرين ومن أجل توفير أرضية واقعية لهذه الدراسة، تمت محاكاة مصنع سوائل الغاز الطبيعي المدروس على ثلاثة مستويات مختلفة ضرورية لتحقيق الهدف من هذه الرسالة.

أولاً، تم بناء محاكاة الحالة التشغيلية المستقرة للمحطة مع دمج كافة تفاصيل التصميم ومن ثم التحقق من صحتها باستخدام بيانات تصميم المصنع الحقيقية.

ثانياً، تم إنشاء محاكاة ديناميكية للمحطة فوق طبقة محاكاة الحالة المستقرة من أجل تمثيل السلوك العابر لعمليات المصنع كاستجابة للتغيرات المحتملة الوقوع. بالإضافة إلى ذلك تم التحقق من المحاكاة الديناميكية التي تم إنشاؤها عن طريق إجراء عمليات محاكاة اختبارية مع سيناريوهات مستمدة من بيانات المصنع الفعلية والتي تمت مقارنة الاستجابة لها والتحقق منها.

بعد ذلك استخدمت المحاكاة الديناميكية التي تم التحقق من صحتها لإنشاء نموذج للمصنع وهو أمر ضروري لتصميم أنظمة التحكم المبنية على نماذج التنبؤ.

أخيراً، تم بناء نظام إشرافي فوق طبقة المحاكاة الديناميكية من أجل تحقيق المتطلبات الموضوعية المتعددة لوحدة المعالجة المختلفة.

تم تقديم إطار عمل لصياغة مشكلة السيطرة المثلى. توضح هذه الدراسة تفوق أنظمة التحكم المتقدمة المبنية على نماذج التنبؤ في تحقيق الأهداف الاقتصادية المختلفة لسلسلة من أعمدة التقطير في مصنع سوائل الغاز الطبيعي في مملكة البحرين وهي وحدات عمود تقطير نزع الايثان و عمود تقطير نزع البروبان عمود تقطير نزع البيوتان.

تم محاكاة العديد من السيناريوهات التشغيلية على غرار تجارب المصنع الحقيقية، من أجل تقييم قوة أنظمة التحكم المبنية على نماذج التنبؤ للحفاظ على أداء التحسين الخاص بها. كما تم النظر في العديد من سيناريوهات المحاكاة مثل التغييرات في معدلات التغذية والتركيبات. أظهر التحليل المقارن قدرة وتفوق نظام التحكم المتقدم المبني على نماذج التنبؤ على التحسين المستمر للمتطلبات الموضوعية المختلفة المحددة لكل وحدة من الوحدات. تقدم الأطروحة أيضًا تقديرًا للفوائد الكمية التي يجب تحقيقها.

تم تقدير الإيرادات الإضافية السنوية البالغة 1,971,565 دولارًا أمريكيًا عند تحسين إنتاج المنتجات الرئيسية الثلاثة لمصنع سوائل الغاز الطبيعي، وهي البروبان والبيوتان والنافثا. أشارت التحاليل الاقتصادية عن جدوى الاستثمار في نظام التحكم المتقدم المبني على نماذج التنبؤ للمصنع قيد الدراسة وعن فترة سداد تبلغ 1.25 سنة.

جامعة البحرين

قسم الهندسة الميكانيكية

كلية الهندسة



تحسين إنتاجية المنتج وجودته في مصنع محلي لسوائل الغاز الطبيعي باستخدام أنظمة التحكم المبنية على نماذج التنبؤ

بحث لاستكمال متطلبات الحصول على درجة (الماجستير) في برنامج الإدارة الهندسية

اعداد

يوسف احمد النفيعي

20160021

إشراف

د. محمد بن شمس

مملكة البحرين

مايو 2021