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Data based distillation column's steam consumption analysis

Master's Thesis

Process Engineering Master

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1. Introduction

The process of separation using distillation columns is one of the main utilities consumer in any chemical plant due to their high heat demand. In most cases, energy required for separation is provided using steam in the distillation reboiler. Approximately the same amount of energy is removed in the condenser but at lower temperature using a cold fluid like cooling water. Supplying more energy to the column than necessary will not only result in higher steam cost but also higher cooling utility demand and cost. As a result, distillation optimization is an essential step to optimize the total plant's utilities consumption.

Optimization of the energy consumption in distillation columns involves two aspects; the first aspect is the selection of number of trays and feed location and the second aspect is the selection of the operating conditions that result in minimum operating cost. Both aspects are determined during the design phase of the column. However, over the lifetime of a column, operating conditions often change due to equipment aging, fouling, process drift, etc.
Therefore, to determine the optimal operating conditions for a system that is currently in use, historical operating data such as temperature, pressure and inlet and outlet flow rates is collected and analyzed. The aim is to identify the actual state of the unit and its operation characteristics. By making use of the available data, operators and process engineers can gain insight into the running of the process. Consequently, data based decisions can be made to improve the unit's performance.

The unit studied in this work is composed of sequentially operated distillation columns which separate acetone, phenol, a-methylstyrene (AMS), cumene and traces of other components. The columns are thermally integrated to make use of the heat available. This means that the condenser of one column is used as the reboiler of another one. This reduces both the hot and cold utility requirements at the cost of increased complexity of the control task.

The first column C-01 which separates acetone in the overhead is considered the main consumer of steam in the separation process. A reduction in its steam consumption will have a notable effect on the total plant steam consumption. The steam consumption in this column is oscillating and exhibits flow disturbances whose causes are unknown. The historical data of the column operation will be studied to assess the current status of the column, the performance of the control system and identify the possible causes of steam oscillation and disturbances. The theoretical amount of steam required for the column will be calculated and modeled to be compared to the actual steam consumption. The operating periods with notable deviation between the model and the actual consumption will be investigated.

Process operating data from any industrial process is usually complex. Availability of large amount of data is beneficial for any analysis as it gives wider process insight. At the same time, one should take into consideration that the larger amounts of data will be accompanied by larger errors that could lead to acquiring false information, skewed results and as a result, wrong decisions. Therefore, a data assessment is required to limit the data that should be used in the model. The assessment must identify the bad quality data that needs to be eliminated or corrected before being used in the model.

The data assessment in this work is supported by basic process understanding. The measured data will be compared to the expected values. The operating data values are expected based on equipment operating principles. The readings from different sensors indicating similar physical quantities will be compared with each other for consistency. Cross-checks for violations of operating conditions or the physical constraints will be carried out. Heat and material balances will be performed and checked if they are satisfied.

For this work, the Seeq analytics software tool is used. Seeq analytics tool connects to the plant historian where historical process data is stored. The software facilitates continuous monitoring, searching and cleansing and modeling of process data as well as performing correlation and causality analysis.

1.1 Scope of Work

The scope of this thesis is to investigate the historical data of the plant and perform data analytics in relation to steam usage and oscillation. The aim of the investigation is to suggest possible areas of operation improvement based on the findings. Improving the operation require identifying the existing problems and its root causes. Once the problem is formulated and the data is obtained, the operation assessment can be performed.

A mathematical model of the distillation column under study C-01 is to be created as a first step to make sure the distillation is capable to perform what is desired. The model helps simplify the analysis, study the heat integration and calculate the heat provided through integration with other columns.

Afterwards, the steam required for this column can be calculated and modeled as a function in the interconnecting heat exchangers duties. The steam consumption model is compared to the actual historical steam consumption data. It can be used to find periods when the actual consumption highly deviates from the model. Operational changes during these periods will be studied.

Defining the causal relationships between signals helps to find the causes of the variation of the steam consumption. A causality analysis that is provided in the Seeq software will be conducted. The causality method employed here is based on Granger causality.

In addition, a correlation analysis can discover the relationship between signals. The analysis looks for the signals that are highly correlated to the steam consumption taking into consideration the time lags and time shifts.

The problems observed in the steam consumption (oscillation and disturbances) could be a result of poor control performance. Therefore, the performance of the control system and its impact on steam consumption will be assessed.

1.2 Work Methodology

The methodology applied in this work consists of the following steps: formulating the problem, acquiring a solid understanding of the system, collecting performance data and performing data analytics.

1.2.1 Problem Formulation

The main concern of the operators is that the plant suffers from steam consumption oscillation and steam consumption disturbances in the column C-01 that results in higher energy cost. The operator noticed that the steam is oscillating because of oscillations in the column control tray temperature TC-01. The operator's explanation was that the temperature oscillation was due to the change in feed composition. To determine the cause of oscillation and disturbances, the entire control configuration, operating condition disturbances and operating modes need to be considered in detail.

1.2.2 Process Knowledge Acquisition

The plant personnel provided DCS screens that display the operating information of the distillation columns under study. The screens served as an acceptable description of the process showing process interactions between columns. The process was then simplified in the flow sheet presented in chapter 3, Figure 16.

1.2.3 Data Collection

The raw measured time series data was obtained by connecting Seeq, an industrial time series analytical software, to the plant historian. Seeq allows reading data in near-real-time as the data arrives in the historian. Laboratory composition analysis reports for the feeds and overheads of the column C01/02/03/04/05 were provided.

Data related to operational constraints, operational modes, instrument maintenance schedules, and controller tuning were also obtained through meetings with the plant operators and process engineers.

1.2.4 Data Analysis

The analysis of the data is conducted through the following steps:

- Check the consistency of the provided data and its reliability.
- Performing Correlation and causality analysis
- Performing material and energy balance
- Modeling the historical steam consumption and comparing the model to the actual steam consumption
- Assessing the performance of the column's control

1.3 Thesis Structure

This thesis can be divided into 7 chapters as follows:

- In Chapter 1, the motivation for this thesis is explained, the scope of work is defined and work methodology is presented.

- In Chapter 2, the literature review for this work is presented. It includes the theoretical background about data mining and data analytics in the process industry, distillation mass and heat balance calculations, distillation control, controller oscillation and valve stiction.
- In chapter 3, the case study is described and the process flow sheet and the control philosophy including the advanced process control (APC) of the distillation column under study are presented.

- In chapter 4, the data set available is assessed, correlation and causality analyses are conducted, and mass and energy balance calculations are performed.

- In chapter 5, The APC performance assessment is performed using the time series data available. The assessment identify the abnormal operating conditions, the percent of time the APC works in automatic mode, the percent of time the constraints are violated and the APC's response to step changes. Possible causes for the APC poor performance such as disturbances, operating condition changes, model errors, and soft sensors errors are investigated.

- In chapter 6, the temperature control loop TC-01 reporting to the APC and its cascaded control flow loop FC-01 are assessed and possible causes of loop oscillation such as external disturbance, interacting loop, controller tuning and sticky valves are investigated.

2. Literature Review

This chapter presents the background on data analytics in industrial processes, distillation principles, different approaches to control composition in distillation column, model predictive control and control loops assessment.

2.1 Data Interpretation and Analysis in Operation Improvement

Industrial manufacturing processes usually have large volume of data that can reach 20,000 measurements of different variables for large scale plants. These data are complex, dynamic and highly dimensional .There are large number of correlated and interrelated variables which makes assessing the large volume of data challenging.

Over the last years, advances have been achieved in data interpretation and analytics. The goal is to use the collected data to provide information for plant operators and supervisors that help understand the process behaviour, monitor the performance, assess the current status, gain knowledge about normal and abnormal operation, identify problems and causes of poor performance and make the right decisions to improve operation. The work flow is shown in Figure 1 (Wang, 1999).

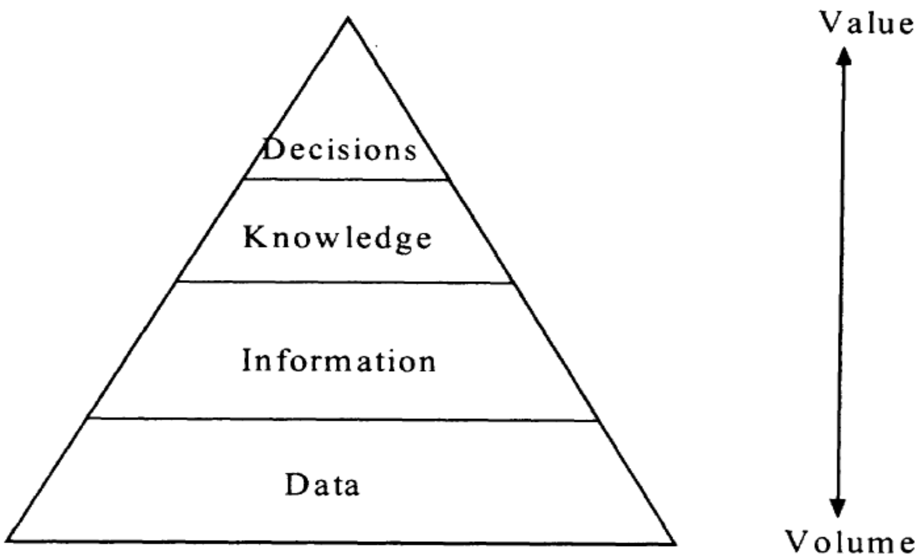

Figure 1: Value of Data (Wang, 1999)

2.1.1 Correlation Analysis

Process signals are characterized by the existence of mutual relationships between each two or more signals. Handling large number of process signals with measurement noise, dynamic responses and time lags makes it difficult to find the relationships between these signals. A root cause that started a plant-wide disturbance is difficult to be identified and the disturbance propagation path is difficult to be followed. A number of data-driven techniques have been developed for root cause analysis. These techniques use historical process data to perform root cause analysis. A very common used technique is correlation analysis. When correlation analysis is performed on two time series signals X and Y, it reveals the relationship between the two signals, determines whether X precedes Y or Y precedes X and estimates the time delay (Bauer & Thornhill, 2008).

Correlation analysis can be performed using Seeq tool to identify the best correlation between two signals in the presence of time shift and time lag effects, and quantifies the amount of lead or lag time in the signal relationship. The correlation results are visually displayed in "heatmap correlation matrix". The heatmap summarize the cross-correlations for each pair of signals in the dataset. The blue represent the direct (positive) correlation. The bluer, the higher the positive correlation. The red represent the indirect (negative) correlation. The more red, the higher the negative correlation.

2.1.2 Causality Analysis

Oscillating control loops are very common in industrial process. Oscillation in a loop propagates to the whole plant through interlinking process equipment and mass and heat transfer. The propagation can result in a plant-wide problem and negatively impact the process economics. Detecting the root cause is not an easy task as it requires good understanding of the causal relationships between process signals and cannot be performed using the standard loop performance measures. Yet, it is important to overcome the plant variations and upsets. So a strategy is required to analyze the cause and effect relationships in a plant data set (Bauer & Thornhill, 2008).

Several methods have been proposed to deal with cause and effect relationships to identify the root cause of plant-wide oscillation. The most common method to test the causal relationships is the Granger-Causality. Granger-Causality is a data based method to determine whether one time series is useful in forecasting another. It measures the causal effect from one time series to another based on linear predictions. According to Granger causality, X is said to Granger-cause Y if the inclusion of the past data of X and Y in the linear regression model of both X and Y reduces the prediction error of Y compared to the model using solely the past data of Y (Yuan & Qin, 2012).

The idea of Granger method illustrated in Figure 2 (Liu & Bahadori, 2012). When X Granger-causes Y, every time point of the effect time series Y is influenced by the cause time series X with a time delay. Thus the inclusion of the data of X time series in the prediction of Y time series gives better results.

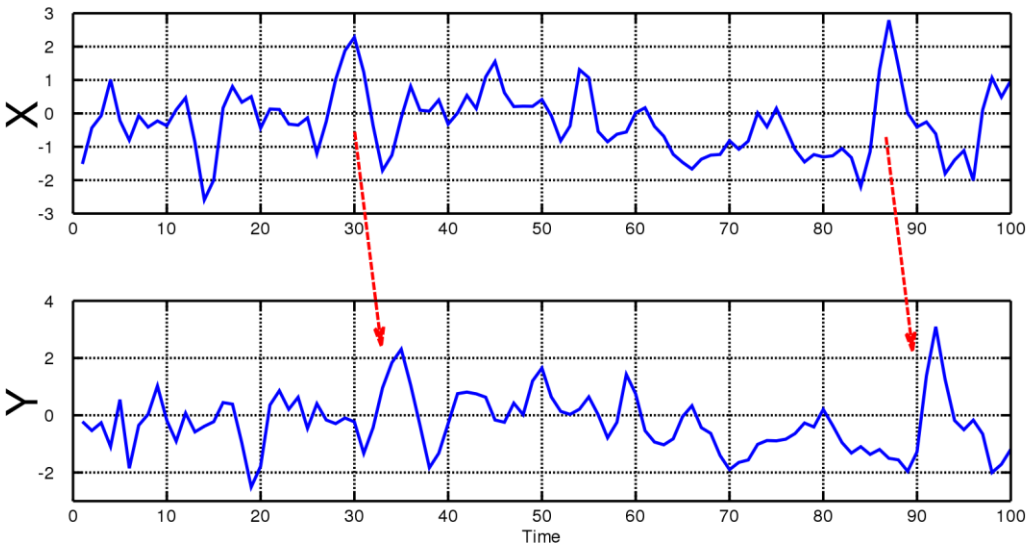

Figure 2: Granger causal relationship between time series X and Y (Liu & Bahadori, 2012)

Traditional Granger-Causality testing methods have its limitations. To ensure valid and effective Granger causality test, the data should be stationary that means that it should have constant statistical properties such as mean, variance and autocorrelation over time. Time series data is often non-stationary (Maddala, 2001) which limits the effectiveness of the traditional Granger causality. Accordingly, several variants of the Granger-Causality have been developed. A simple method that can be applied to non-stationary data is Toda and Yamamoto (T-Y) Granger-Causality method (Toda & Yamamoto, 1995) and it is the method used in Seeq causality analysis.

Seeq uses T-Y Granger-Causality to generate a causal map in which the signals are represented by circles and the causal influence between signals is shown through arrows. The dependent signals (effects) are represented by red circles. The smaller and the more red circles represent the most dependent signals. The independent signals (causes) are represented by blue circles. The larger and bluer circles represent the most independent signals. Uni-directional relationships are represented by darker arrows, whereas bi-directional relationships are represented by lighter arrows.

2.2 Distillation Operation and Control

Distillation is the most used equipment for separation in chemical industries, it is an application of mass transfer operation. It is based on the difference of volatility between the components of the feed mixture. The distillation is multistage separation, where vapor and liquid contact on a tray or packing to exchange mass and heat to reach equilibrium. The more volatile components are vaporized and geos up to the top of the column where they are partially or totally condensed in the distillation's condenser. Part of the condensed overhead product will be returned to the column as reflux. The less volatile components are condensed on the column's trays and goes down to the bottom of the column where part of the liquid will be vaporized in the reboiler and returned to the column as boil-up. A schematic drawing of a distillation column is shown in Figure 3 (Smith , 2012).

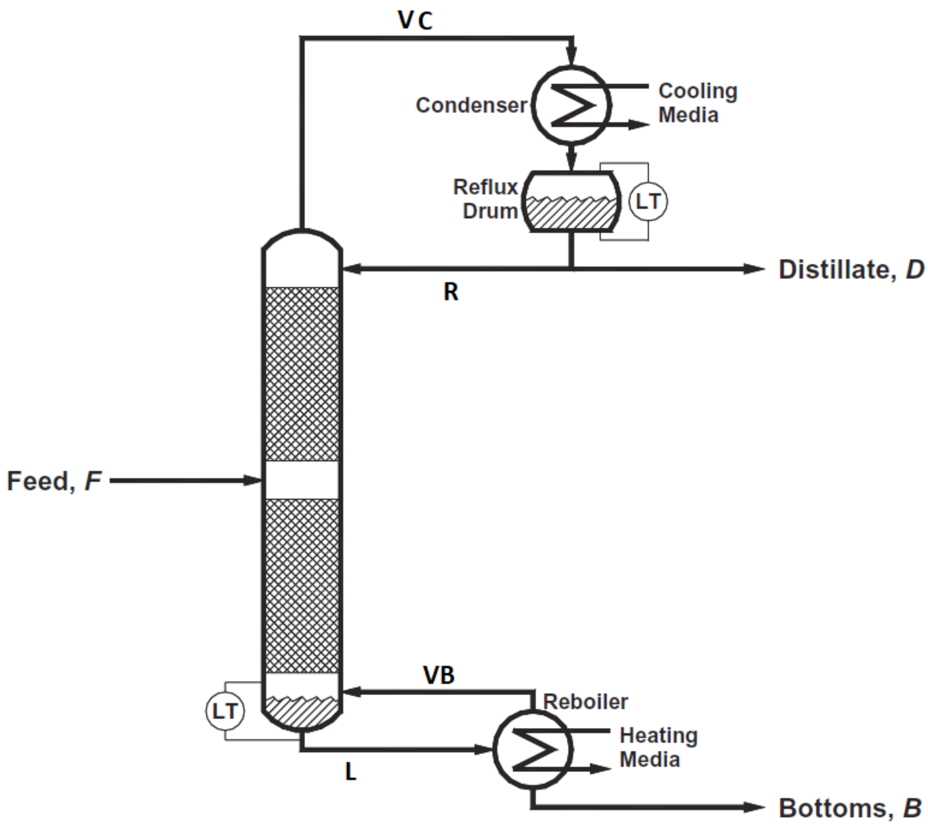

Figure 3: Distillation column (Smith , 2012)

2.2.1 Distillation Mathematical Model

The Mathematical model of the distillation consists of the mathematical relationships between its variables and is used to describe the behaviour of the system. The model is based on the first principles of mass, energy and component balance (Smith , 2012). The physical and thermodynamic properties used in the energy balance equations are obtained from Yaws chemical properties handbook (Yaws, 1999).

2.2.1.1 Model Equations

Material balances are the most fundamental equations used in the model of any process. The material balance equations for the distillation system are:

- The steady state total material balance equation, this equation must close (material balance must be satisfied) on the long run.

$$F = B + D \tag{1}$$

- Condenser material balance.

$$VC = R + D \tag{2}$$

- Reboiler material balance.

$$\mathrm{L} = VB + B \tag{3}$$

Where F is the feed flow rate, B is the bottom product flow rate, D is the distillate flow rate, VC is the column's overheads, R is the reflux, L is the column's bottoms and VB is the boil-up.

Component material balance equation can be written for each component n in the feed as follows:

$$F\ Zn = D\ Yn + B\ Xn \tag{4}$$

Where the composition of the component is Zn in the feed, Xn in the distillate and Yn in the bottom product.

The overall Energy balance equations can be written as follows:

$$\Delta H = Q_R + Q_C \tag{5}$$

Where Q_R is the reboiler heat input and Q_C is the heat removed by the condenser.

The performance of the distillation column depends on the composition of its products. Therefore, composition control is extremely important in the distillation column. Composition control can be single - end composition control or double - end composition control. In single- end composition control, only one product composition is controlled whereas in double- end composition control, the composition of both products is controlled.

The column studied in this work employs advanced process control to control the composition of the bottom product. In what follows, the different methods for composition control, its advantages and disadvantages are discussed. The aim is to understand the basic composition control principles based on which the advanced process control in this column was developed and to discuss the standard composition control shortcomings that should be overcome by the advanced process control.

2.2.2 Temperature and Composition Control

Column temperature control is the most common, reliable, easy and inexpensive method used to control product composition in a binary system operated under constant pressure. A binary distillation column is a column used to separate only two components from each other. As the temperature of a boiling mixture is thermodynamically related to composition and vapor pressure, keeping the system temperature and pressure constant implies constant composition. However, in multicomponent systems the temperature is not a reliable indicator of composition due to non-ideality nature of mixtures and also because of possible disturbances in feed composition. In such systems, even if the system is able to keep a constant temperature, this does not imply a constant composition or that the separation composition control objectives are being met (Kister, 1990).

Due to the limitations of temperature control in multicomponent systems, on-line composition analyzers were introduced to directly measure the product quality and feed results into the automatic control system. A sample is withdrawn from the product line and sent to an analyzer which runs analysis and gives electrical signals that are proportional to the percent of each component in the sample analyzed. The control component value is recorded and used as a control signal (Santon & Sterling , 1979). The sampling point location has great importance in determining the analyzer system dynamics, reliability and accuracy (Kister, 1990).

A study conducted on an acetic acid recovery column to compare how different control systems will react to changes in feed composition and how the product purity is affected showed that a system

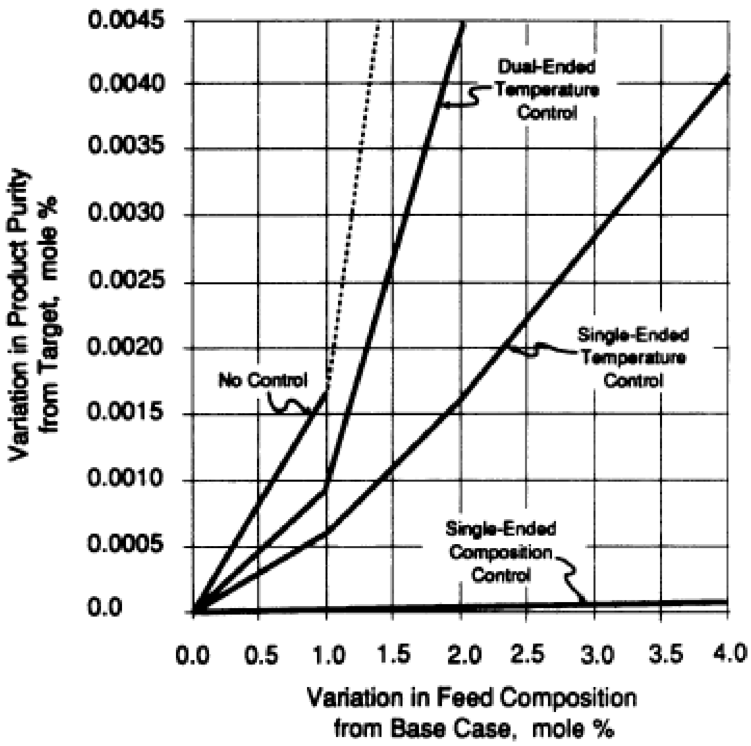

Figure 4: Disturbance rejection of various sensor based control systems for an acetic acid recovery column (Luyben, 1992)

using analyzer gives much better product quality control than conventional temperature based schemes. The results of this study are shown in Figure 4 (Luyben, 1992).

Although an analyzer gives better composition control, it has some drawbacks that makes its usage difficult. Large measurement lags are inherent in the measurement which result in response delays. This in turn can result in an unsatisfactory control system performance. These lags are mainly because of process lags such as the time lag when sampling from an accumulator, sample transfer lags or the time for the sample to travel from the sampling point to the analyzer, and analyzer transfer lags or the time from analyzer valve to detector and sample intervals. The overall lag can be 10-30 minutes. This long cycle time to generate composition results from the analyzer prevents early detection of process disturbances and causes slow responses to process changes leading to long disturbance recovery times (Kister, 1990).

To benefit from the analyzer control and the temperature control and at the same time mitigate some of their problems, a Temperature – composition parallel cascade control configuration can be used (Patke, Deshpande, & Chou, 1982).

2.2.3 Temperature – Composition Parallel Cascade Control

In this configuration the output of the analyzer controller is fed to a temperature controller whose output acts as the set point to a secondary output controller. In this configuration the slow response of analyzer is overcome by the rapid response of temperature controller as the main control action is taken by the temperature controller and the analyzer is only responsible for changing the temperature controller set point to keep the product composition within limit. When the analyzer is not in operation the temperature controller will take full responsibility for automatic column control. The fast action of the temperature controller helps in fighting process disturbances. In this configuration the reflux flow rate can also be controlled by the analyzer (Luyben, 1992).

For this temperature – composition control system to have a good performance, the temperature controller must have a fast control action (faster than its set point adjustment by analyzer) and its sensitivity to key components concentration must also exceed its sensitivity to any expected pressure variations. This is most important under vacuum. The control tray should be selected to give a good indication of composition. Selecting the best control tray can be challenging and is considered a main issue in the use of temperature control composition (Kister, 1990).

2.2.3.1 Drawbacks of Temperature Composition Control

Temperature - composition control has overcome most of the problems related to analyzer only control. However, its malfunction still holds an important spot in the list of distillation malfunctions. Mostly, temperature - composition control shortcomings are seen in chemicals or gas plants performing close separations. (Kister, 2006)

One of the shortcomings of temperature composition control is that the transient response is quite poor. A study was conducted to compare the closed-loop response of the single-temperature feedback scheme with the inferential and parallel cascade schemes (Patke, Deshpande, & Chou, 1982). In this study, a depropanizer column was simulated with the different control schemes. The primary control objective is to maintain the composition of isobutane in the top product. The response of each control scheme was tested for different six feed disturbances. The study results shows that;

- In the case of using single ended temperature control the composition highly deviates from the target. Closed-loop response of single ended temperature control is shown in Figure 5.

- In the case of using a parallel cascade scheme, the offset in isobutane composition is eliminated but results in overshoot. Closed-loop response of the parallel cascade control is shown in Figure 6.

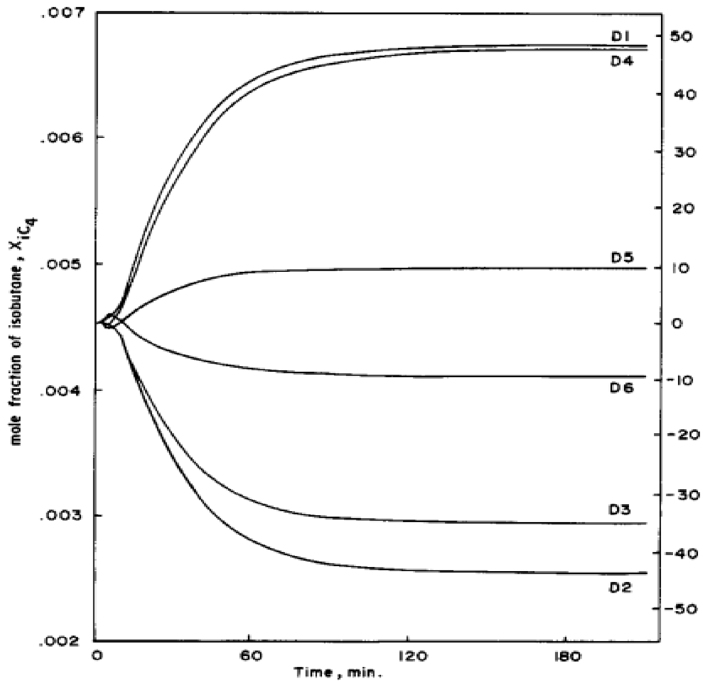

Figure 5: Closed-loop response of the feedback control system to disturbances in the feed (Patke, Deshpande, & Chou, 1982)

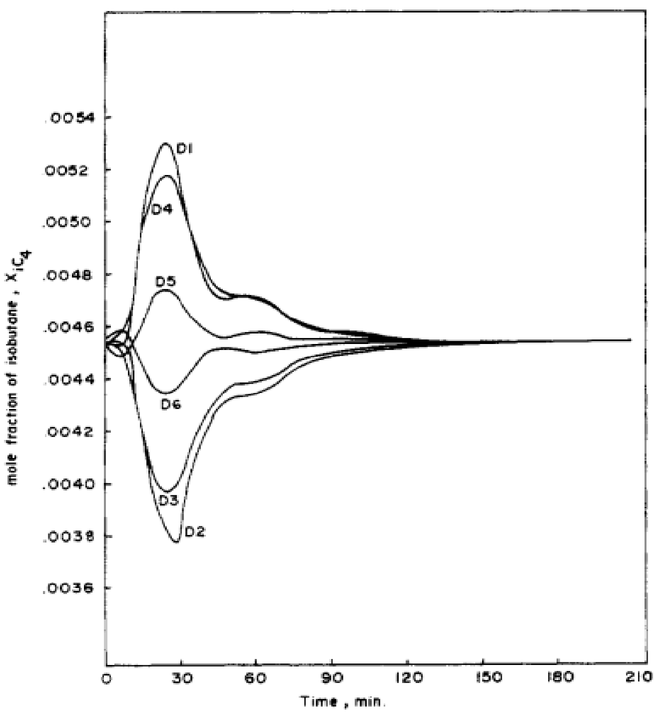

Figure 6: Closed-loop response of the parallel cascade control system to disturbances in the feed (Patke, Deshpande, & Chou, 1982)

Another serious problem of the temperature – composition control configuration is that it is not suitable to be used in systems where feed flow is not stable and enthalpy is not controlled. Instability in feed flow and enthalpy results in operating the controller in automatic mode less than half the time. (Kister, 1990)

Santon and Sterling studied temperature composition control using a depropanizer column (Santon & Sterling , 1979). Besides illustrating the economic benefits, they illustrated in-depth the requirements and the challenges facing this control scheme. These requirements are as follows:

- Sample time should be less than 5 minutes otherwise the automatic control will deteriorate rapidly. Figure 7 shows the integrated squared error corresponding to different sample times. The error signal is the difference between set point and measured process value.

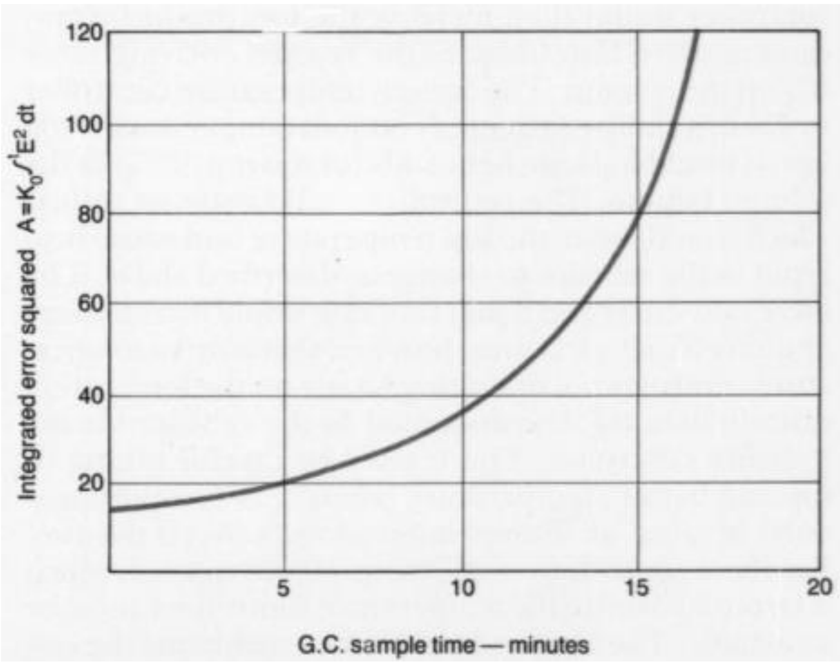

Figure 7: control performance corresponding to different analyzer sample time (Santon & Sterling , 1979)

- The conventional Temperature composition control shown in Figure 8 can only be used in case of slow feed flow rate changes. When the flow is highly changing they recommended using the local optimal control configuration shown in the Figure 9.

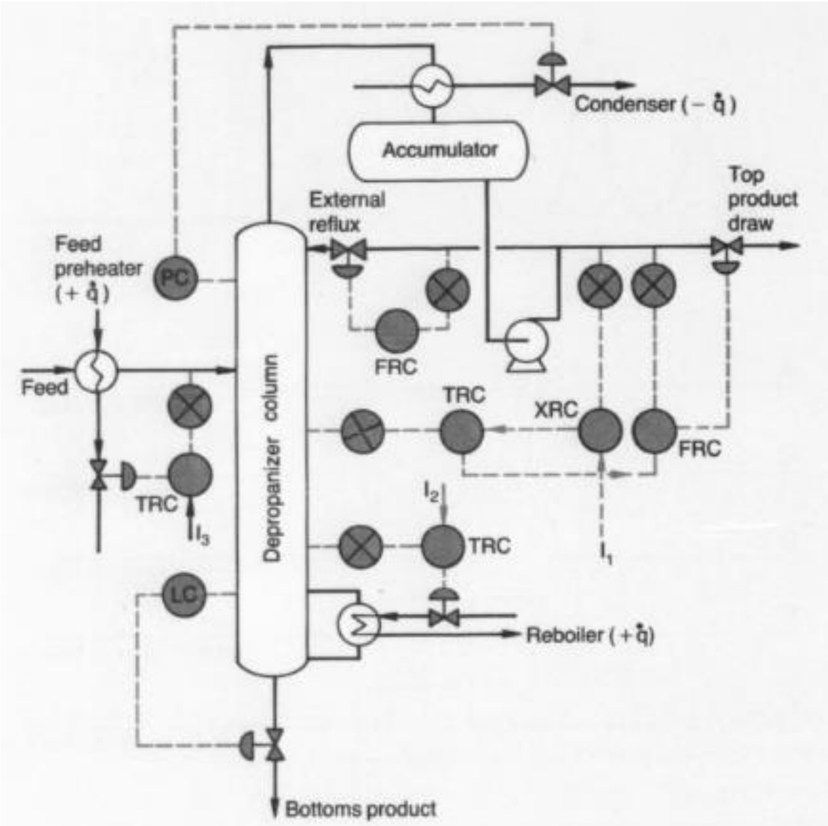

Figure 8: Depropanizer column with conventional temperature-composition control (Santon & Sterling , 1979)

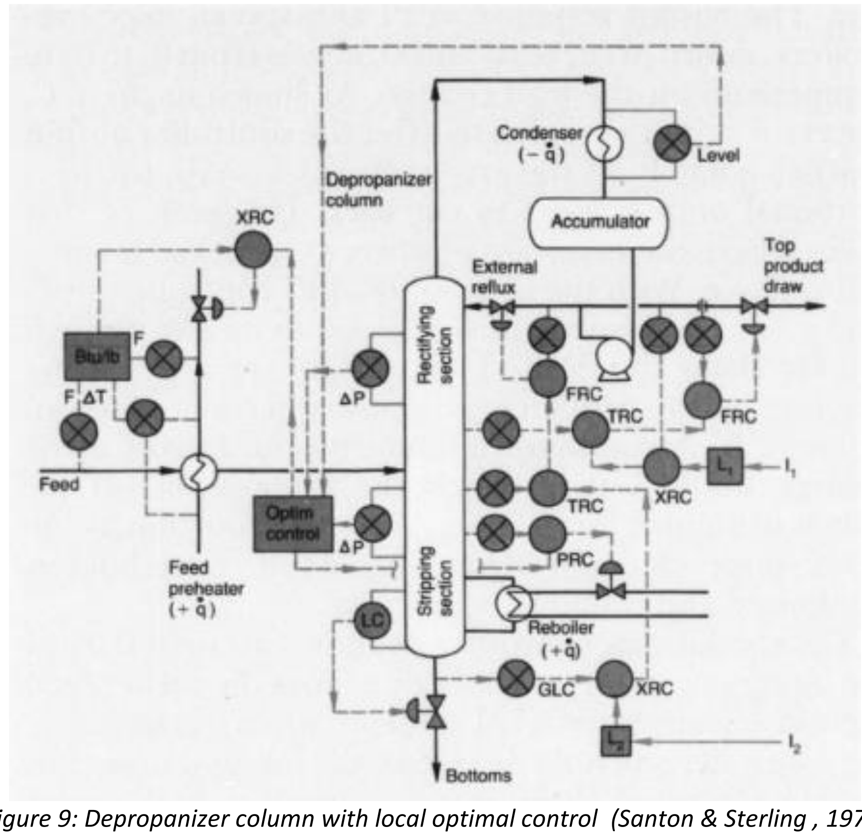

Figure 9: Depropanizer column with local optimal control (Santon & Sterling , 1979)

2.2.2.2 Effect of Feed Enthalpy on System Control

Feed enthalpy is the amount of heat supplied to the distillation column via the feed stream(s). Although this amount of heat is generally small compared to the total heat input, regulating feed enthalpy plays an important role in a distillation control system.

In a distillation using temperature – composition control, changes in feed enthalpy affect the performance of the primary control system (analyzer control loop). A change in feed will be considered as disturbance that the analyzer controller tries to compensate by changing the temperature set point resulting in a change in the steam supply.

To study the effect of feed enthalpy changes on control quality, an experiment was conducted on a distillation column using temperature – composition control (Lupfer & Oglesby, 1961). The aim of the experiment is to compare how the analyzer loop corresponds to feed changes in case of feed enthalpy control existence (A) and in case of its absence (B). The following conclusions were made:

- Changes in feed enthalpy affects the performance of the primary control system (analyzer control loop). A change in feed will be considered as disturbance that the analyzer controller tries to compensate for by manipulating the reboiler heat input. Changes in feed enthalpy and reboiler heat input will result in changes in liquid and vapor flows inside the column accompanied by composition changes at the analyzer location. The analyzer will try to maintain the composition by manipulating the reboiler heat input again. The measured value will oscillate and the primary control loop will be unstable. Eventually, the analyzer controller must be detuned for more stable operation.
- It is recommended to regulate the feed enthalpy as it gives better analyzer controller performance. Figure 10 compares the good controller responses obtained when feed enthalpy is controlled (response A) to the response obtained when the feed enthalpy is not controlled (response B).
- Analyzer controller performance is improved with increasing the sample rate. Figure 11 shows that the best results are obtained with a controlled feed enthalpy and high speed analyzer.

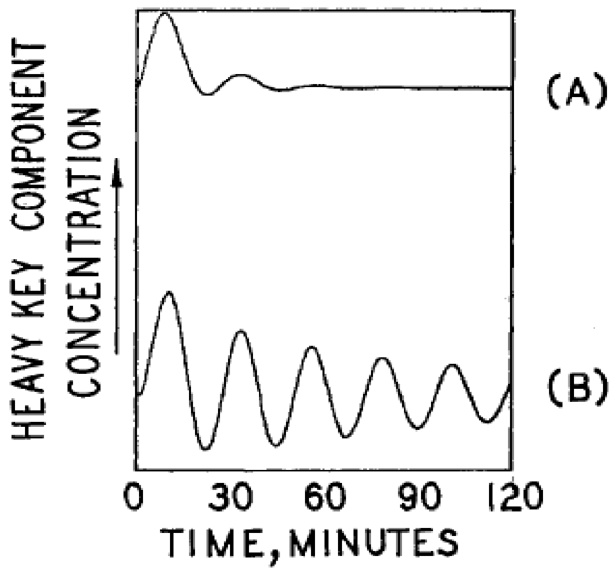

Figure 10: Analyzer controller response to load step changes (Lupfer & Oglesby, 1961)

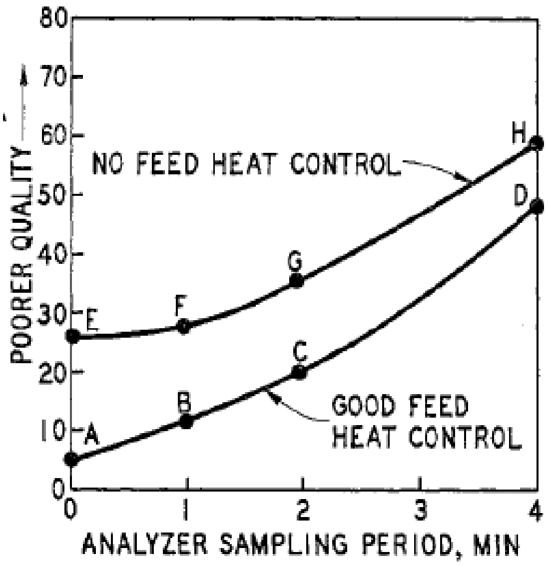

Figure 11: Effect of feed enthalpy control and sampling rate on controller performance (Lupfer & Oglesby, 1961)

Due to the problems associating the parallel cascade control system, the need to an advanced control system has arose to handle the interacting complex control loops in distillation columns.

2.3 Advanced Process Control (APC)

In the recent decades, advanced process control (APC) technology has gained a great importance and wide usage in petrochemical industries. The economic importance of the APC system is highly demonstrated in the control of distillation columns as they are highly non-linear system and always subjected to severe disturbances. These distillation characteristics makes the conventional control system performs unsatisfactory and fails to achieve the required product quality control with efficient energy consumption. It is reported that the energy consumed to achieve a certain product quality can be 30 to 50% more than is thermodynamically required by the system. But with the application of better column control, the energy consumption can be reduced to the optimum level. (Humphrey , Seibert, & Koort, 1991)

The recent control approach is to use a multi layers control structure. The structure consists of the base layer that regulates that plant using number of standard PID controllers. The higher control layer is the advanced control layer whose output determines the input to the base layer to keep the system operating within a certain optimal operating point. Then comes the real time optimizer that calculates the optimal operating point. On the top layer, enterprise planning is applied (Lu , 2003).

2.3.1 Model Predictive Control (MPC)

The most popular form of the APC used in distillation is the model predictive control (MPC). The popularity of the MPC in the industry mainly comes from its ability to predict the future process path and steer the multi-variable systems to the optimum condition while taking into considerations the constraints of the process variables. In case of using Model predictive control, the real time optimizer can be integrated with the advanced process control layer or can be implemented as a separate higher layer. The targets and limits of the MPC is determined by the real time optimizer or higher planning and scheduling layer. The MPC manipulated variables are the set points of the PID controllers executed in the distributed control system (DCS). The model predictive control structure is shown in Figure 12 (Darby, Harmse, & Nikolaou, 2012).

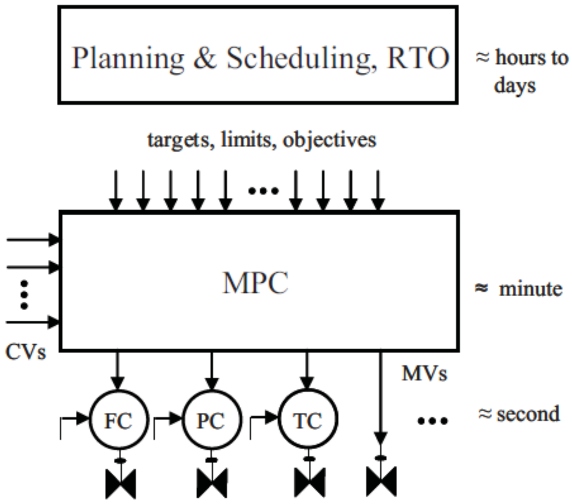

Figure 12: MPC structure (Darby, Harmse, & Nikolaou, 2012)

The basic idea of the model predictive control is to use a dynamic system model to predict the future process path based on the current situation. If the predicted path is not achieving the controlled variables' targets, unable to keep the process within constraints or at the optimal operating point, the APC solves a constrained optimization problem to give an optimum output. For solving the optimization problem, the APC takes into consideration the current plant measurements, the process variables targets and constraints. The problem output determines the trajectory of the manipulated variables' to steer the process to the optimum steady state operating conditions.

The optimization problem is solved on line in each time instant (t) and gives an optimal control sequence as shown in Figure 13: MPC working strategy scheme Only the first step in the sequence is applied to the plant and the remaining steps are discarded. In the next time instant (t+1) new measurements are collected from the plant and the optimization problem calculation is repeated. The goal is to minimize the prediction error, the difference between the predicted output and the desired reference, over a prediction horizon as the error from proximal forecast is smaller than the error of distant predictions (Bemporad & Morari , 1999) (Schwenzer, Ay, Bergs, & Ab, 2021).

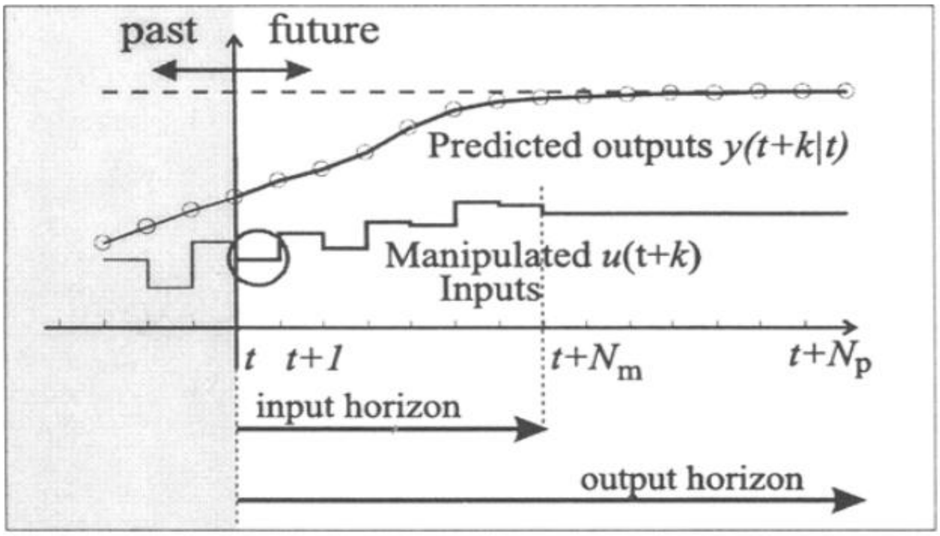

Figure 13: MPC working *strategy scheme predictions (Bemporad & Morari , 1999)*

Implementing a model predictive control is a challenging mission that require a lot of effort. The base layer controls should be evaluated and the instrumentation should be determined if they are adequately performing. The base layer problems should be found and fixed before implementing the MPC (Darby, Harmse, & Nikolaou, 2012). The system disturbances and constraints should be taken into consideration to guarantee feasibility, stability and robustness. The measurements of the controlled variables is not always available on frequent basis and sometimes it is subjected to large delays. Such delays deteriorate the control performance and lessen its ability to handle disturbances. The approach followed to come over this problem is to use soft sensors for inferring process output.

2.3.2 Soft Sensors

Soft sensors are used in many industrial applications to estimate any process variable or product quality. The estimation of process variables is done using an inferential mathematical model. This model estimates the output process variable using the information acquired from other measurable variables. Soft sensors offer many advantages such as reducing the need to hardware devices, allowing real time estimation of the data, overcoming hardware sensors' time delays and sampling limitation and realization of a reliable control. (Fortuna , Graziani, Rizzo, & Xibilia, 2007)

In distillation columns, the soft sensors are mostly used In place of online analyzers to control product quality. Online analyzers usually introduce time delays to the process and subjected to be taken out of operation for maintenance or re-calibration. Therefore, soft sensors represent a cost effective and reliable alternative that estimates the product quality using the available tray temperature. (Patke, Deshpande, & Chou, 1982)

2.3.3 MPC Constraints

Considering the process constraints in solving the optimization problem makes the APC outperforms the conventional control systems in its ability to handle inputs' and output's limits. Constrains are divided into hard constraints and soft constraints. Hard constraints are the limits that cannot be violated and have to be satisfied. Soft constraints are the limits that can be violated temporarily. Although the deviation of soft constrains is allowed, there are always undesirable consequences such as loss of profit or damages to valves or equipment. (Rossiter, 2018)

The constraints can be physical constraints such as actuator limits, safety constraints such as pressure and temperature, product specification constraints, economic and environmental constraints. Physical and Safety constraints are always hard constraints and given high priority. The MPC always tries to achieve them first, then comes the step to optimize other objects. (Rossiter, 2018)

The MPC ability to handle constraints is one of the important factors of its success and popularity in chemical plants, but at the same time, this feature adds much complication to its implementation. Having hard constraints turns the optimization problem non-linear, restricts the degrees of freedom of the optimization problem and can lead to infeasibility and instability. (Schwenzer, Ay, Bergs, & Ab, 2021)

2.3.3.1 Constraint Analysis

Constraint analysis plays an important role in the assessment of model predictive control performance. The analysis is performed by comparing the process values of the manipulated and the controlled variables against their upper and lower limits. Quality of the model predictive control performance is determined by its ability to keep the process variables within the control limits. (Gaoa, et al., 2003)

Statistical process control (SPC) techniques have been introduced to monitor the control performance, identify abnormal behaviour and detect control limits violation. These techniques use process control charts to judge whether the process is on or off target. The most popular SPC chart is Shewhart chart, in which the process values, the mean value, and the control limits are plotted. The target is usually specified as the grand mean. An example is shown in Figure 14 . The points that violate the limits are marked with red and it indicates the process is not well controlled. (Jelali, 2013)

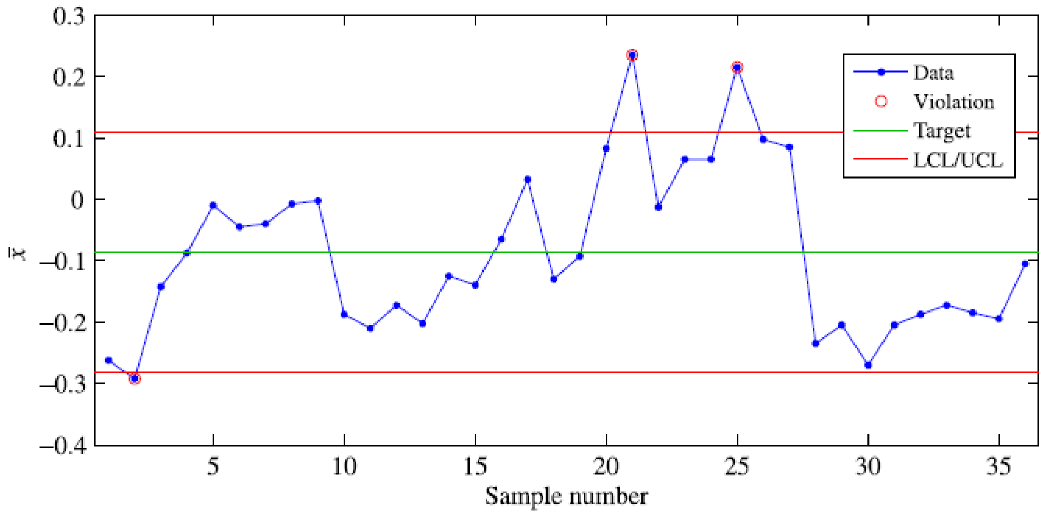

Figure 14: Shewhart control chart (Jelali, 2013)

2.3.4 MPC Feasibility

One of the challenges associating the MPC is ensuring its feasibility. The term feasibility concerns the MPC optimization problem. Feasibility of the optimization problem means that the problem results in feasible set points that enable achieving the targets and at the same time operating within constraints. In case of rapid disturbances, the existence of input and output hard constraints often render the optimization problem infeasible, and the set points become unreachable. Infeasibility and unreachable set points will always lead to system instability.

Thus, the MPC tend to relax the output constraints, as they usually defined as soft constraints, to guarantee the MPC feasibility and maintain the process in the desired operating regime. Constraints relaxation takes place by adding a slack variable to the optimization problem to increase the degree of freedom available to the optimizer and to penalize the violation. The size of the slack variable corresponds to size of the associated constraints violations. (Schwenzer, Ay, Bergs, & Ab, 2021) (Zhao, Lu, Zheng, & Huang, 2012)

2.3.5 MPC Stability and Robustness

Model predictive control stability is the feature concerned with resulting a system's bounded output from a bounded input. A stable controller should achieve the predetermined output despite the disturbances of the control loop. To guarantee the stability, the model predictive control feasibility has to be guaranteed.

System stability is followed by system robustness, which means the model predictive control is able to deal with uncertainties and meet the performance specifications. A real Plant has different types of uncertainties and the model of a model predictive control can never be perfectly representing the plant. Yet, the model should be designed to consider the uncertainties and external disturbances affecting the plant dynamics. Different approaches have been developed to deal with uncertainties and obtain a robust predictive model. (Zhao, Lu, Zheng, & Huang, 2012)

2.4 Control Loop Assessment

Base layer control loops are the main components regulating the operation of the plant. Operation safety, product quality and energy consumption are highly linked to the performance of the base control layer. Moreover, the performance of the model predictive control highly depends on the quality of the base control layer. However, it has been reported that a large portion of industrial control loops, as many as 60%, has performance problems. These problems found to be negatively impacting the performance of the whole control structure. So it is important to monitor and analyze the performance of the control loops to ensure good control performance. (Jelali, 2013) (Lahiri, 2017).

Poor control loop performance can be the result of the failure of one of its components; controllers, sensors and control valves. It is important to understand the failure of each component and its consequences to be able to assess the control loop performance. Poor performance can also be the result of external disturbances, poor process design, process oscillation and lack of maintenance. Performance of control system loops is presented in Table 1. (Lahiri, 2017)

Control Loop Status	**Percent**
control loops in manual	10% to 90%
loops-tuning is completely wrong	30%
loops oscillating	40%
Suboptimal tuning	85%

Table 1: Performance of Control Loops in Industry

As one can see in the table, one of the common problems in control loops is oscillation. Oscillation is exhibited in the control system under study. Therefore the focus in this section is on loop oscillation and its possible causes. There are several reasons for oscillation. The major reasons are external disturbances, aggressive tuning and control valves nonlinearity (Srinivasan, Nallasivam, & Rengaswamy, 2011).

2.4.1 External Disturbances

Disturbances are unwanted inputs that affect the measured process variable. The control loop can have many disturbances; internal and external. External disturbances usually come from the upstream process or from interacting loop (Jelali, 2013). In modern plants, the wide energy integration between different equipment contributes to the propagation of disturbances. If these disturbances are not properly measured and compensated for, it deteriorates the control performance (Lahiri, 2017).

2.4.2 Controller Aggressive Tuning

Aggressively tuned controller can quickly develop oscillation in the control loop and its response is characterized by the following: (Lahiri, 2017)

- ***High peak overshoot ratio (POR):*** aggressive tuning increase overshoot in response to set point change.
- ***Large decay rate:*** large decay rate is associated with oscillation in step response.

2.4.3 Control Valve Stiction:

Control valves are the final control elements in a control loop. They are mechanical devices that are subjected to wear and tear and require regular maintenance. With time, they develop several problems such as stiction, saturation, large deadband, backlash and corrosion. Stiction is the most common problem in control valves. If the valve has stiction, it may has oscillatory output which in return leads to oscillatory process variable. Stiction can be defines as the static friction that prevents the valve from movement. The stiction keeps the valve from moving in case of small controller output change, it only moves when there is enough force to overcome the static friction. Stiction can be detected by putting the controller in manual mode and make small controller output changes and monitor the process variable. If the process variable does not change with small controller output changes and it only change after considerable controller output change has been done, then the valve has stiction (Choudhury, Shah, & Thornhill, 2008).

Valve stiction can also be detected from closed loop operating data without the need of experimentation with the plant. In this non-experimental approach the valve input and output data are plotted and checked against the ideal input- output behaviour. An ideal valve behaviour is linear but a valve under stiction exhibits the behaviour illustrated in Figure 15 (Zabiri & Mazuki, 2009).

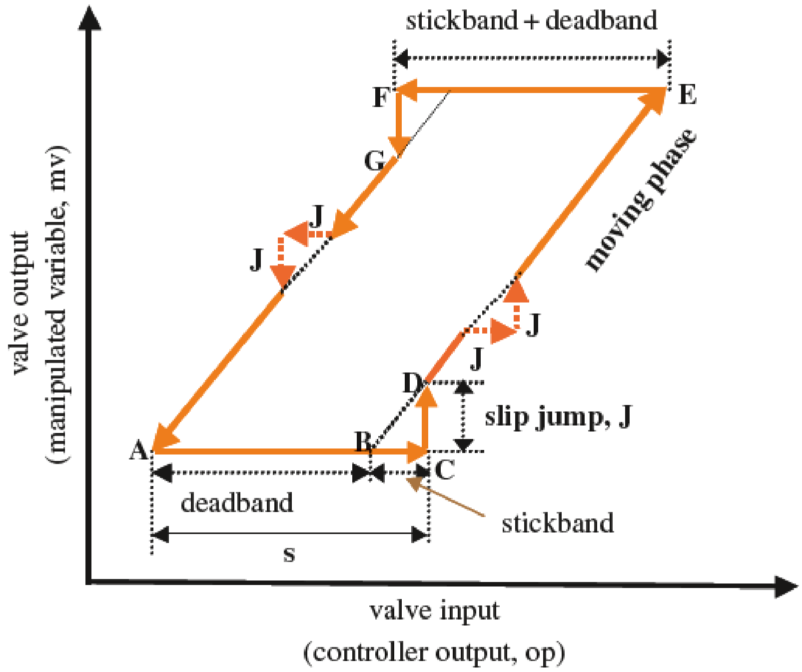

Figure 15: Input-output behaviour of sticky valve (Zabiri & Mazuki, 2009)

3. Case Study

The unit studied in this thesis is the first distillation column in a phenol plant. The plant produces phenol from cumene and separates the produced chemicals from non-converted raw material through subsequent distillation columns. These distillation columns are thermally integrated to make use of the heat available and use the condenser of one column as the reboiler of another one. In this chapter, the flow sheet and the control scheme of the first column that separates acetone from the reaction product mixture are described.

3.1 Process Flow Sheet

The distillation column under study (C-01) separates the acetone in the overhead and send the remaining mixture that contains traces of acetone and various other heavy components from the bottoms to column C-02. In column C-02, acetone is separated in the top product. An analyzer is installed on the overhead product line of column C-02 to measure acetone component percentage. Figure 16 shows the flow sheet of the columns C-01/02 with the detailed control scheme and measurement points of column C-01.

A multicomponent stream comes from an upstream unit in the process and is fed to column C-01. A second stream is being fed to the column to enhance the separation process. This stream is mainly water with phenol and cumene traces, separated from the downstream process. The bottom stream from the column C-01 is divided into three parts, one continues to the next column C-02, the second is fed to the reboiler R-01, and third part is sent to the heat exchangers HEX-01/02/03 that, through integrated heat transfer, assists the reboiler.

The column C-01 is therefore thermally integrated with three subsequent columns C-03/04/05 through the heat exchangers HEX-01/02/03. This means that the condensers of those other columns are used as the heating fluid in HEX-01/02/03 to vaporize the boil up in the column C-01.

The instrumentation for this column includes level indicator on the column LI-01, level controller LC-02 to control the condenser liquid level, pressure controller PC-01 that controls the overhead column pressure, pressure indicator PI-02 to indicate the bottom pressure, temperature indicator TI-02 that indicated the main feed temperature , temperature indicator TI-03/04 that indicate the temperature in different locations in the column, temperature controller TC-01 controlling the temperature in the stripping section, flow indicator FI-05 which indicates the overhead product flow, composition analyzer AI-01 analyzing the product composition, flow controllers FC-01/02/03/04/06 and control valves that are located on the steam, main feed, reflux, second feed and bottom product represent flow control loops that maintain flow rates at specified levels. The flow indicators FI-07/08/09 are located at the overhead product lines of the columns C-03/04/05.

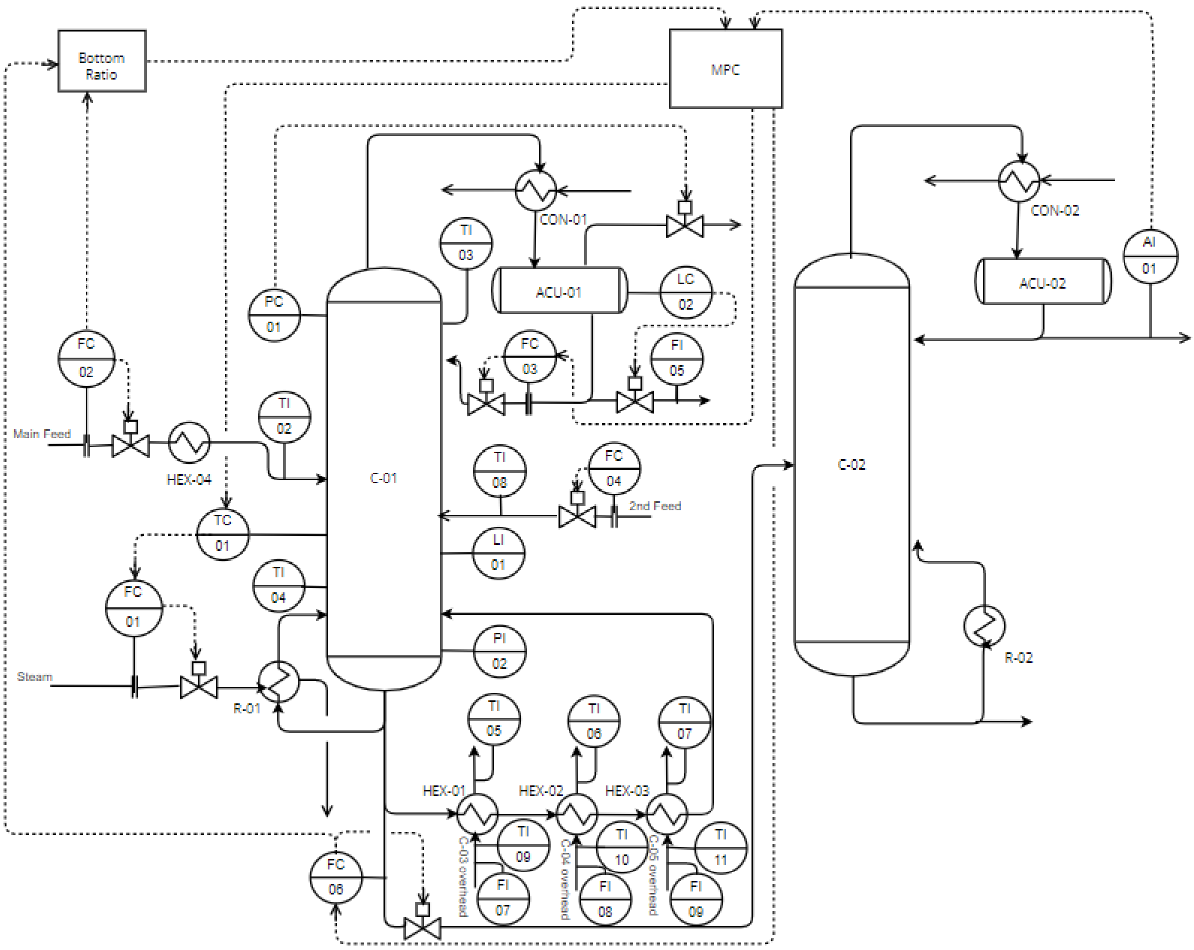

Figure 16: Process Flow Diagram for distillation columns C-01/02

3.2 Column C-01 Control Scheme

The feed to the column C-01 (FC-02) is cascaded to a controller in the upstream process section and its set point is manipulated based on the received signal from that master controller. The feed stream is not equipped with an analyzer so its composition is not measured. The set point of the second feed flow rate FC-04 is an operator input. The column overhead pressure PC-01 is controlled by manipulating the non-condensables flow from the condenser accumulator (ACU-01). The accumulator level (LC-02) is adjusted by varying the overhead product flow rate.

The stripping section temperature controller (TC-01) is cascaded to the reboiler steam flow controller (FC-01). The temperature controller is the master controller whose output is used to adjust the set point of the steam flow controller FC-01 and thus the heat input into the column.

The control objective is to control the acetone composition at the overhead of the column C-02 under a defined boundary conditions. Due to process complexity, difficult dynamics, and the various external disturbances affecting operation, the basic control level cannot perform satisfactorily to achieve the control objective. Therefore, an advanced process control (APC) system that includes an embedded model predictive control (MPC) function is installed and linked to the plant's distributed control system (DCS).

The APC has two controlled variables; the acetone concentration (AI-01) and bottom stream ratio to the feed (BR). The first controlled variable (acetone concentration) is more important than the second. The manipulated variables are the control tray temperature (TC-01), the reflux flow rate (FC-03) and bottom flow (FC-06). The manipulated variables' target values also have different weightings; the temperature target is given the highest weighting. The APC has 4 operating modes, the description of each mode is presented in Table 2.

APC operating mode	Automatic /manual
Mode 1	Fully automatic (all APC inputs and outputs are in automatic mode)
Mode 2	Partially automatic (the optimizer continues to calculate the controlled variables targets, the APC selects the set point for reflux and bottom flow controller but keeps the temperature set point constant at the last calculated value when the APC was in operating mode 1.
Mode 0	APC switched off
Mode -3	Limits are closed (upper and lower limits for each variable are equal)

Table 2: APC operating modes description

3.3 Data Description

The data set available contains 26 process variables (PV), 8 controller output (OP), 8 controller set points (SP), 4 MPC variables targets (T) , 4 MPC variables upper control limits (UCL) and 4 MPC variables lower control limits (LCL) and APC statues. The data is studied in the period from 1st January to 30th June, 2022. The measurements of column C-01 are shown in Table 3.

ID	tag name	Description	No. Data read	Comment
1	FC-01	Steam flow control	- PV: 1,304,699 - SP: 2,156,470 - OP: 2,773,359	- Oscillating SP and PV. - No steam supply for six hours on June 4th and for two days from June 5th to June 7th - Steam flow is significantly higher than normal flow although the feed does not significantly change from February 6th to February 12nd.
2	TC-01	Control tray temperature control	- PV: 2,258,347 - SP: 214,396 - OP: 2,633,645 - T: 84,210 - UCL: 6,032 - LCL: 6,035	- Changing controller's set point in operating mode 1. - Constant controller's set point in operating mode 2. - PV and OP time trends always show oscillation. There are variations in the oscillation period and amplitude. - Oscillation period ranges from 15 to 70 minutes. - The oscillation pattern suggest more than cause of oscillation. - It does not show the same exact oscillation pattern as LC-02, PC-01 and PI-02, FI-05. However, it can be observed that a change in the oscillation amplitude

				of one of these parameters is accompanied by changes in the amplitude of the others. - Target never exceed the limits. - The operators determined the set point to be higher than the upper limit on the days of 27th February, 1st March and 2nd March, 2022. - The set point time trend shows frequent and rapid changes in the set point in the period from March 21st to May 20th.
3	FC-02	Main feed control	- PV: 2,717,261 - SP: 2,803,686 - OP:3,090,583	- Feed flow disturbances are observed. - Feed cut off on June 4th from 10:00 to 14:00.
4	TI-02	Main feed temperature indicator	- PV: 916,951	- Readings' range is from 58-70°C which suggest the temperature is not controlled. - The temperature is inversely correlated to the feed flow rate.
5	PC-01	Overhead pressure control	- PV: 2,642,147 - SP: 6,192 - OP: 3,415,144	- PV and OP are always oscillating. There are variations in the oscillation period and amplitude. - Oscillation period ranges from 15 to 40 minutes.
6	TI-03	Overhead temperature indicator	- PV: 2,019,322	
7	FC-04	Second feed flow control	- PV: 2,560,868 - SP: 4,595 - OP: 3,904,308	- Set point determined by operator

8	TI-08	Second feed temperature indicator	- PV: 2,560,868	-
9	LI-01	Column Level indicator	- PV: 2,839,546	
10	TI-04	Bottom temperature indicator	- PV: 1,446,790	- This temperature reading is higher than TI-05/06/07 which means temperature cross in the heat exchanger. Suspected to be malfunctioning.
11	PI-02	Bottom pressure indicator	- PV: 2,676,885	
12	TI-05	HEX-01 outlet temperature indicator	- PV: 776,284	- Readings' range suggests that the temperature is not controlled. - No signal from 24/03 – 26/03.
13	TI-06	HEX-02 outlet temperature indicator	- PV: 4,517	- Readings' range suggests that the temperature is not controlled. - No signal from 24/03 – 26/03.
14	TI-07	HEX-03 outlet temperature indicator	- PV: 4,517	- Readings' range suggests that the temperature is not controlled. - No signal from 24/03 – 26/03.
15	FC-06	Bottom product flow control	- PV: 3,099,017 - SP: 150,677 - OP: 1,602,733	- It's an MPC manipulated variable but there are no time trends for the target and limits.
16	FI-07	C-03 overhead flow indicator	- PV: 2,348,209	-
17	FI-08	C-04 overhead flow indicator	- PV: 16,788	- Oscillation is observed from April 10^{th} to 24^{th}.
18	FI-09	C-05 overhead flow indicator	- PV: 5,900	-

19	TI-09	C-03 overhead Temperature indicator	- PV: 2,348,209	-
20	TI-10	C-04 overhead Temperature indicator	- PV: 16,788	-
21	TI-11	C-05 overhead Temperature indicator	- PV: 5,900	-
22	FC-03	Reflux flow control	- PV: 2,713,459 - SP: 268,800 - OP: 2,702,693 - T: 125,356 - UCL: 5,141 - LCL: 5,150	- Target can violate the limits when the APC is in operating mode2.
23	LC-02	Accumulator level control	- PV: 2,720,971 - SP: 4,388 - OP: 2,721,870	- PV and OP are always oscillating. There are variations in the oscillation period and amplitude. - Oscillation period ranges from 15 to 40 minutes.
24	FI-05	Overhead product indicator	PV: 1,773,549	- Shows the same oscillation pattern as PC-01 and LC-02.
25	AI-01	Acetone analyzer	- PV: 23,924 - T: 106,394 - UCL: 37,115 - LCL: 37,027	- Changing target. - Very frequent target changes from April 1st to April 4th. - Targets can violate the limits while APC is in operating mode 1. - No signal from 10th May to 15th May. - High deviation from the target observed from February 9th to February 11th and from May 25th to May 28th

26	BR	Bottoms to feed ratio	- PV: 55,322 - T: 16,423 - UCL: 4,535 - LCL: 4,532	- Changing target. - Targets can violate the limits while APC is in operating mode 1.
27	APC status	APC operating modes		- Frequent mode changes. - Mode changes often accompany feed disturbances.

Table 3: Study data set

4. Data Analysis Steps

4.1 Data Assessment and Processing

By visual inspection of the data obtained from a process point of view, it becomes evident that some instruments are not reliable. The instrument's readings are suspect because they suggest the column is operating in violation of physical law constraints. An examples is the temperature cross found in the integration heat exchangers.

Reliability questions were confirmed by plant personnel as these instruments are not periodically maintained or checked for accuracy. This makes the instruments' reliability doubtable. As good instrumentation performance is essential for good control loop performance, the existing potentially inaccurate instruments could be contributing to poor control performance.

The visual inspection also showed that, the measured data has outliers that need to be removed before the data is used in the mathematical model, the correlation and causality analysis. Using Seeq tool, data was cleansed and outliers were removed.

4.2 Correlation Analysis

The correlation analysis is conducted using Seeq analysis tool on the entire data set to find the mutual relationships between each two signals, whittle down the number of variables and help to focus on the process variables correlated to the steam consumption. Seeq correlation tool allows for the shifting in time of signals with respect to the signal of interest. Thereby, making the studied signals more useful for interpretation and allowing for the discovery of meaningful process insights. The correlation analysis is conducted on the process variables in different operation periods.

4.2.1 First Correlation Period

First correlation period, shown in

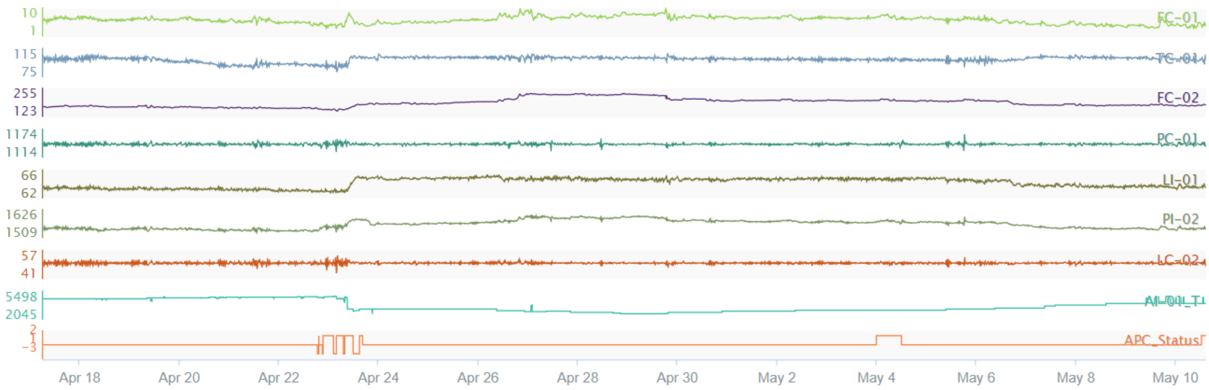

Figure 17, is from April 18th to May 10th, during this period, Feed flow FC-02 and acetone target change many times, steam disturbances are observed, oscillation takes place in TC-01, PC-01 and LC-02 and the APC is mostly operating in mode 1.

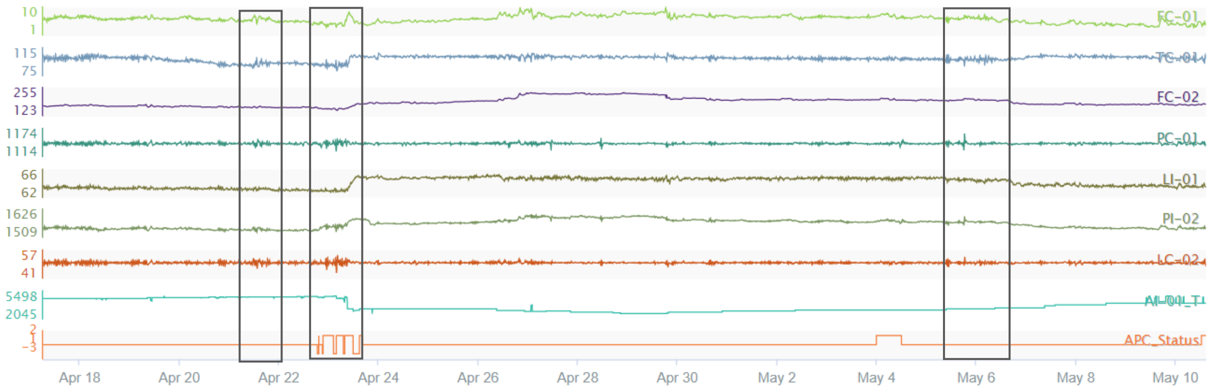

Figure 17: First correlation period

The correlation results (correlation coefficients between each pair of signals) are displayed in the heatmap correlation matrix shown in Figure 18. The matrix summarises the correlation coefficients between each pair of signals. Blue cells represent a positive (direct) correlation between the two signals. The bluer the cell, the higher the positive correlation. While red cells represent negative (indirect) correlation. Negative time shift means the reference signal proceeds the shifted signal. Positive time shift means the shifted signal proceeds the reference signal.

The correlation analysis during this period shows the following:

- Level inside the column is highly correlated to the feed flow FC-02 (correlation coefficient = 0.81). Column's Level is not controlled. As a result, the bottom pressure PI-02 changes with the feed changes (correlation coefficient to the feed = 0.93).

- The correlation coefficient between the pressure PC-01 and the feed FC-02 is 0.03. Column's top pressure PC-01 is well controlled.
- Temperature TC-01 correlation coefficient to the composition AI-01 is -0.51 which means temperature does not perfectly indicate composition.
- Reflux flow FC-03 correlation coefficient to the composition AI-01 is -0.97.
- Temperature TC-01 correlation coefficient to the steam flow FC-01 is 0.46 with a time shift of -0.72 which indicates a slow and poor acting control loop.
- Outlet temperatures from integration heat exchangers TI-05, TI-06 and TI-07 are positively correlated (correlation coefficient > 0.95) to the hot fluid flows to heat exchangers fI-07, FI-08 and FI-09.
- During this correlation period, temperature's set point and temperature's process value change but the set points and process values of PC-01 and LC-02 do not change. The result is high correlation coefficient between PC-01 and LC-02 (0.72) and low correlation coefficients between temperature process value TC-01 and the process values of PC-01 and LC-02 (> - 0.1) despite the simultaneous oscillation observed in their time trends. So this period is not the best choice to study the correlation between the three variables.
- The time shifts between TC-01, PC-01 and LC-02 can be useful to determine the start of oscillation and its propagation path. The time shift between TC-01 and PC-01 is - 0.1 hour and between TC-01 and LC-02 is - 0.2 hour, i.e., TC-01 proceeds PC-01 by 0.1 hour and proceeds LC-02 by 0.2 hour.
- Correlation coefficients and time shifts between steam consumption (FC-01) and other process variables are summarized in Table 4.

ID	Shifted signal	Correlation with FC-01	Time shift to FC-01 (hours)
1	AI-01	-0.66	9
2	FC-04	0.81	-9
3	TC-01	0.46	-0.72
4	FC-06	0.78	-1.63
5	FC-03	0.74	-0.05
6	FC-02	0.8	0.05
7	TI-02	0.52	8.08
8	PC-01	-0.1	-0.57
9	TI-03	-0.34	-9

10	LI-01	0.52	9
11	TI-04	0.52	8.08
12	PI-02	0.72	-0.08
13	TI-05	0.61	9
14	TI-06	0.49	9
15	TI-07	0.67	-3.05
16	LC-02	0.12	-0.17
17	FI-05	0.8	-0.25
18	FI-07	0.7	-3.17
19	FI-08	0.4	9
20	FI-09	0.65	9
21	BR	0.33	2.07

Table 4: Correlation results with respect to FC-01

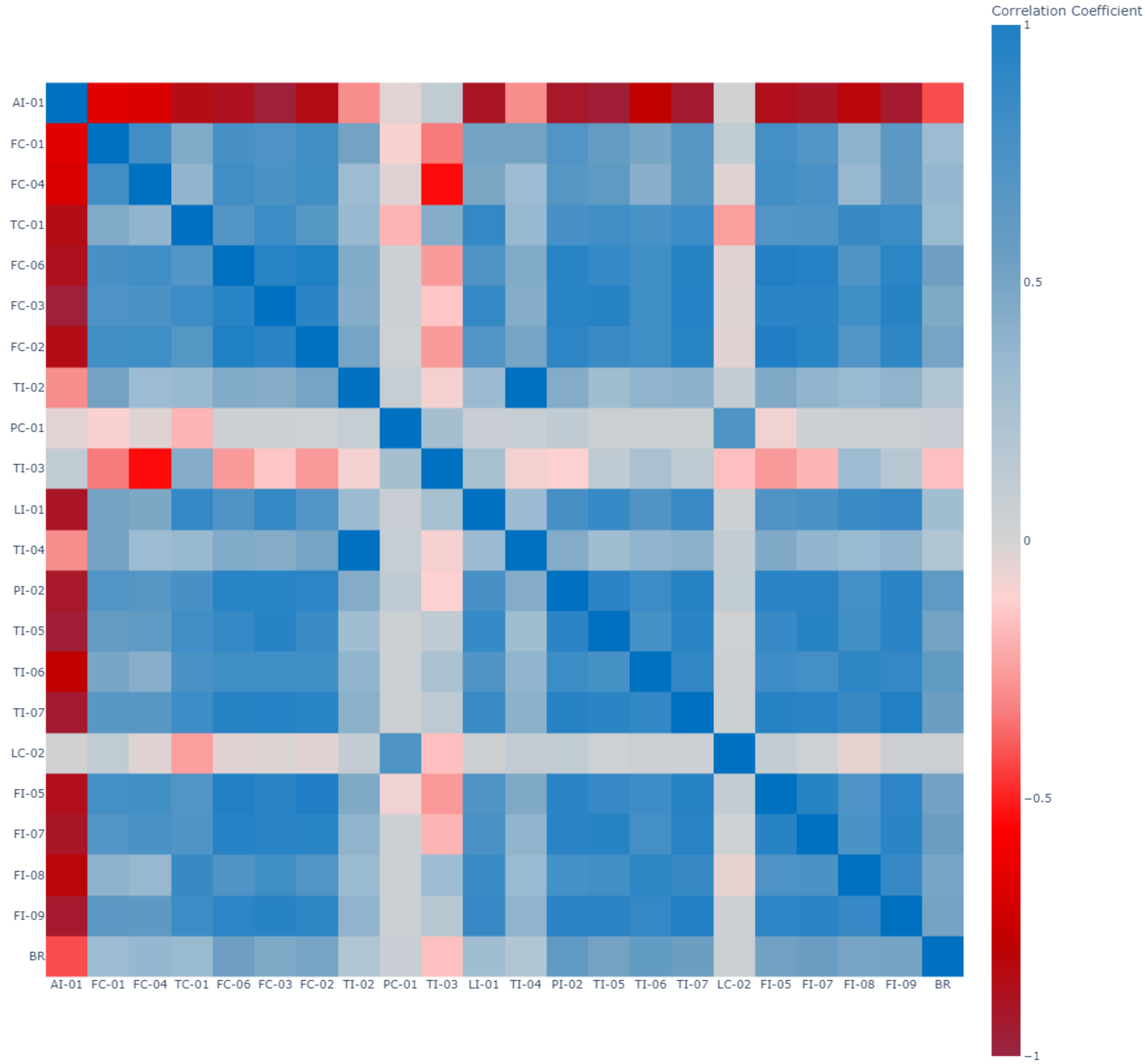

Figure 18: Correlation heatmap - April 18th to May 10th

4.2.2 Second correlation period

Second period, shown in Figure 19 from May 25th to May 28th, during this period, the APC is in operating mode 2, temperature controller's set point is constant which allows studying the correlation of TC-01 with PC-01 and LC-02, feed flow changes, oscillation in different process variables and high composition deviation from the target are observed. The correlation results (correlation coefficients between each pair of signals) are displayed in the heatmap correlation matrix shown in Figure *20*.

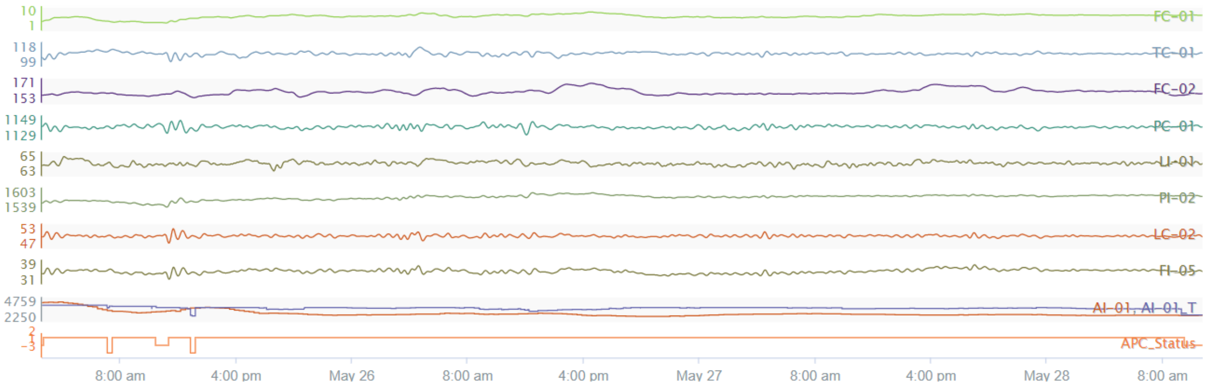

Figure 19: Second correlation period

The correlation analysis during this period shows the following:

- The correlation coefficient between the temperature TC-01 and PC-01 is -0.48 with a time shift of – 0.1 hour. The correlation coefficient between the temperature TC-01 and LC-02 is -0.69 with a time shift of – 0.2 hour.
- Temperature TC-01 correlation coefficient to the composition AI-01 is -0.27. Composition is not controlled by temperature when APC is in mode2.
- Reflux flow FC-03 correlation coefficient to the composition AI-01 is - 0.92. Reflux flow is an important parameter controlling the composition whether the APC is in mode 1 or 2.

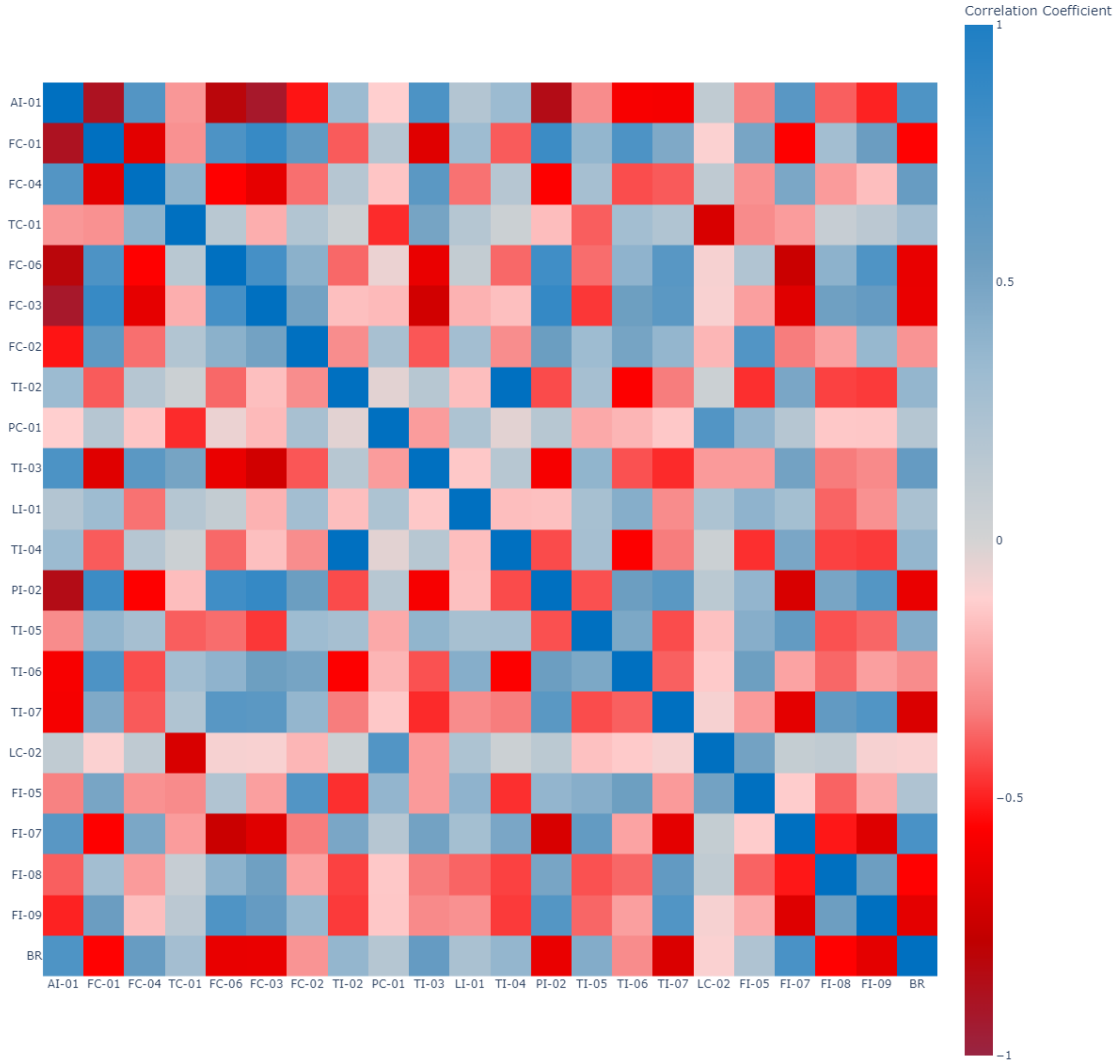

Figure 20: Correlation heatmap - May 25th to May 28th

4.2.3 Third correlation period

Third period, shown in Figure 21, is from March 10th to March 15th, during this period, the APC is in mode 1 and the composition target change frequently followed by changes in the set points of the temperature and reflux flow. The changes in controlled variable target and manipulated variables set points make this period suitable to study the correlation relationships between these variables. The correlation results (correlation coefficients between each pair of signals) are displayed in the heatmap correlation matrix shown in Figure 22.

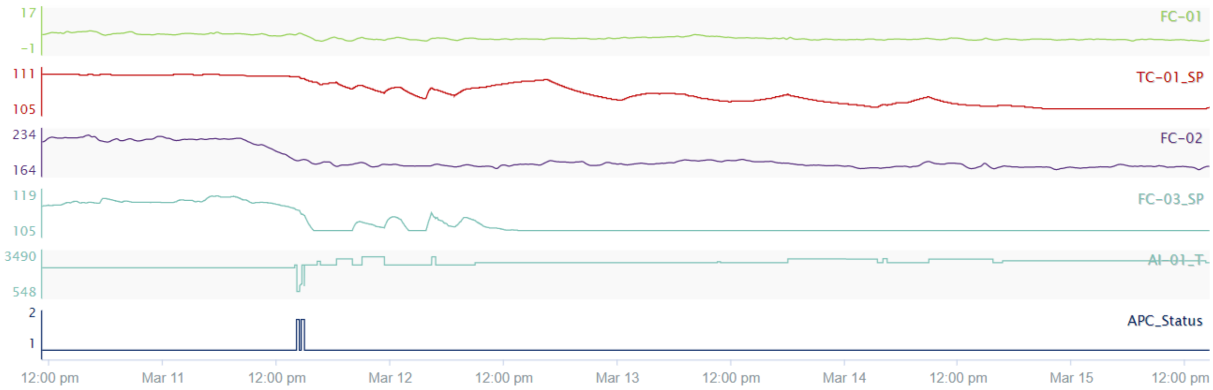

Figure 21: Third correlation period

The correlation analysis during this period shows the following:

- Temperature TC-01 correlation coefficient to the composition AI-01 is -0.65.
- Reflux flow FC-03 correlation coefficient to the composition AI-01 is - 0.79.

From the correlation maps we can see that, different results (correlation coefficients) are obtained from different analysis periods due to different operating conditions. Therefore, these correlation analyses are useful to study the effect of different operating conditions.

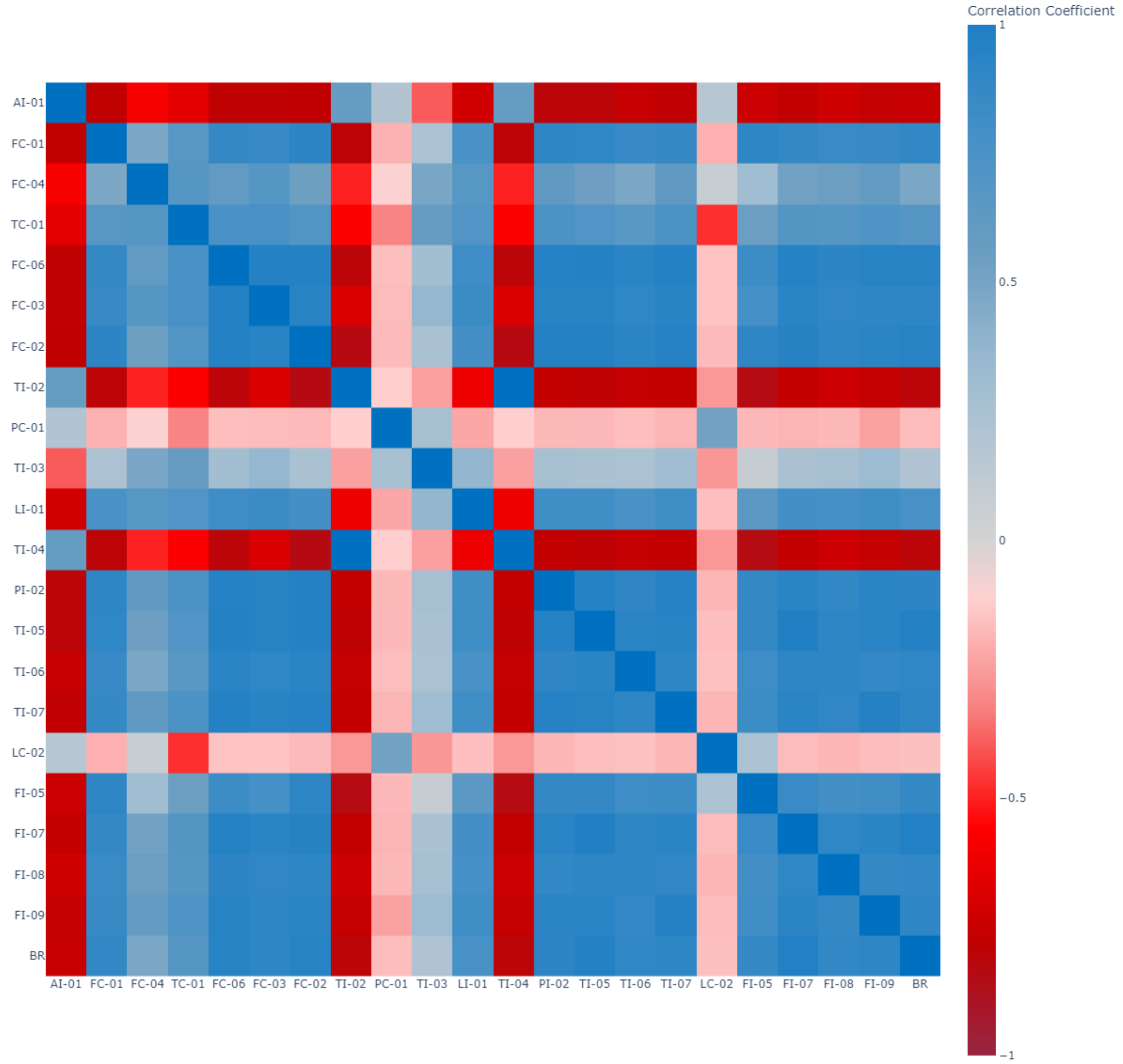

Figure 22: Correlation heatmap - March 10th to March 15th

4.3 Causality Analysis

To help find out the causes for steam oscillation Seeq causality analysis was also performed. The causality analysis conducted on the periods from May 25th to May 28th and from March 10th to March 15th. The results are displayed in the causal map shown in Figure 23 and Figure 24 respectively. The blue circles represent the cause signals while the red ones represent the effects.

Figure 23 shows the following:

- Steam consumption FC-01 changes are caused by LI-01, TI-05, FI-05, AI-01 and FC-04. FC-01 is the one of the causes behind changes in FC-03, TI-03, TC-01, PC-01, PI-02 and FI-08.
- TC-01 is caused by FC-02, AI-01, LI-01, FI-07, and PC-01. The correlation analysis on the same period showed that the oscillation in TC-01 proceeds PC-01.

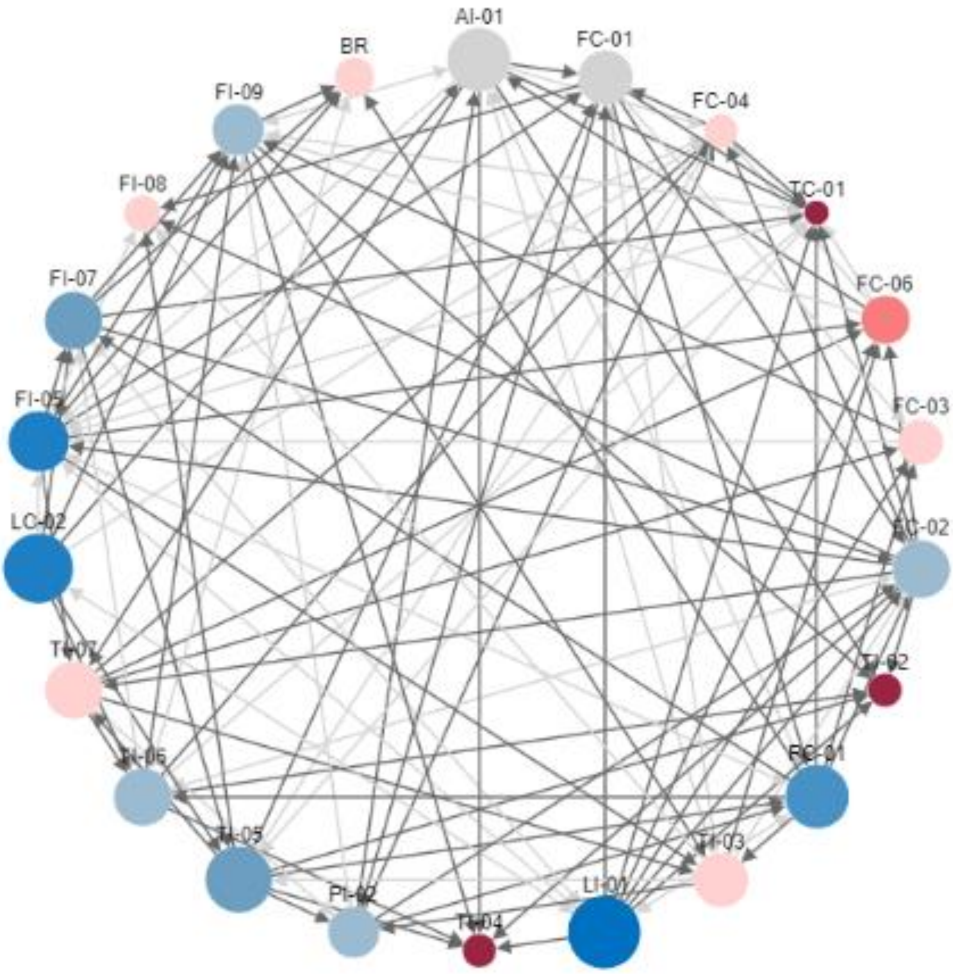

Figure 23: Causality map -May 25th to May 28th

Figure 24 shows the following:

- Steam consumption FC-01 changes are caused by FI-08. FC-01 is the one of the causes behind changes in TI-02, PC-01, TI-04, PI-02, TI-06, and TI-07. There are bi-directional relationships between FC-01 and FI-09, LC-02 and TI-05.
- TC-01 changes are caused by FC-01, FC-02 and TI-03. TC-01 is the one of the causes behind changes in AI-01, BR, TI-02, TI-06, LI-01 and FC-03. There are bi-directional relationships between TC-01 and FI-05, FI-07, FI-08, FI-09, LC-02, PI-02, PC-01 and TI-05.

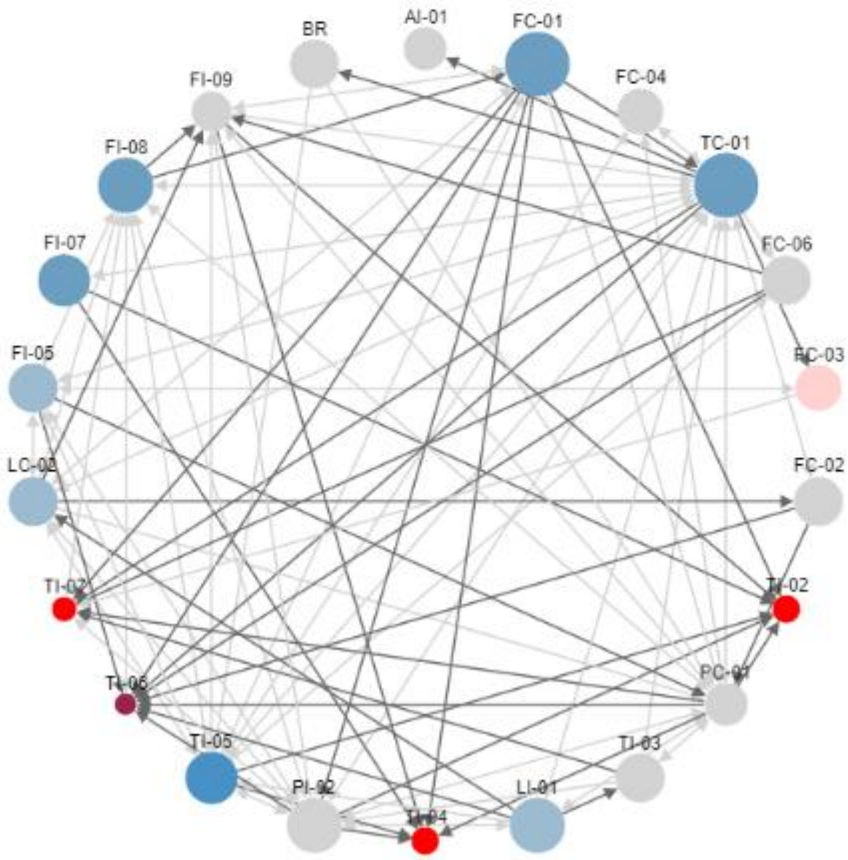

Figure 24: Causality map -March 10th to March 15th

As can be seen, the causal maps give different causal relationships between the same pairs of signals at different operating conditions. It also gives results that do not go along with the results obtained from the correlation analysis which makes this analysis not very helpful in defining the root cause of oscillation.

4.4 Column C-01 Mathematical Model

An important step to assess the operation of the distillation is to develop a simplified mathematical model using the available operation data. The goal is to check if the mass and energy balances are satisfied, calculate the amount of steam should be consumed and find out if the real consumption matches what the model suggests.

4.4.1 Mass and Energy Balance Calculations

Performing simple overall mass and energy balances is one method to detect instrument errors, plant problems and losses. As a model predictive control is employed to control the column C-01, it is important to check the satisfaction of the mass and energy balance. A dissatisfaction in the mass and energy balance will result in model errors and poor model predictive control performance.

Mass, component and energy balance calculations for column C-01 are performed using the first principles equations stated in section 2.2.1. The equations are modified to take into consideration the second feed to the column and the heat input by the integration heat exchangers.

The calculations are performed through the following steps:

I. **Mass balance calculations**: the calculations are performed on daily basis to avoid inlet – outlet mass difference caused by process lags. The inlet and outlet masses are totalized using the aggregation tool in Seeq which enables users to aggregate a signal over a period of time using totalizing function.

$$F = B + D \tag{1}$$

$$F = FC - 02 + FC - 04 \tag{6}$$

$$D = FC - 05 \tag{7}$$

$$B = FC - 06 \tag{8}$$

Where F is the total feed entering the column at time k, D is the distillate flow rate at time K and B is the bottom flow rate at time K at which the flow rates were sampled.

$$M_{in}^{daily} = \sum_{k=0}^{N} F[k]T \tag{9}$$

$$M_{out}^{daily} = \sum_{k=0}^{N} B[k]T + \sum_{k=0}^{N} D[k]T \qquad (10)$$

$$M_{diff}^{daily} = M_{in}^{daily} - M_{out}^{daily} \qquad (11)$$

Where M_{in}^{daily} is the daily mass inlet, M_{out}^{daily} is the daily mass outlet, N is the number of samples per day (8640), T is the sampling rate (10 seconds), M_{diff}^{daily} is the daily inlet-outlet mass difference.

II. **Components balance calculations:** the lab composition analysis reports for the two feeds and the distillate are available. Applying Equation (4) to all the components, the composition of the bottom stream can be obtained.

III. **Streams' enthalpy calculations**: The enthalpy calculations are performed on the streams entering and leaving the columns C-01 and on the overhead products of the columns C-03/04/05 using the bottom temperature (TT-04) of column C-01 as a reference temperature. Liquid state is the reference state. The enthalpy calculation equations and regression coefficients are obtained from Yaws handbook. (Yaws, 1999)

- The change in enthalpy of each individual component in the liquid phase is calculated using the following equation. T_{ref} is the reference temperature and T_s is the stream's temperature.

$$\Delta H_i^{liq} = A_i^{liq}T + \left(\frac{B_i^{liq}}{2}\right)T^2 + \left(\frac{C_i^{liq}}{3}\right)T^3 + \left(\frac{D_i^{liq}}{4}\right)T^4 \Bigg]_{T1}^{T_2} \qquad (12)$$

Where $A_i^{liq}, B_i^{liq}, C_i^{liq}, D_i^{liq}$ are regression coefficients for the component i in the liquid phase. $T1$ and $T2$ are the temperature limits.

- The change in enthalpy of each individual component in the gas phase is calculated using the following equation.

$$\Delta H_i^{gas} = A_i^{gas}T + \left(\frac{B_i^{gas}}{2}\right)T^2 + \left(\frac{C_i^{gas}}{3}\right)T^3 + \left(\frac{D_i^{gas}}{4}\right)T^4 + \left(\frac{E_i^{gas}}{5}\right)T^5 \Bigg]_{T1}^{T_2} \qquad (13)$$

Where $A_i^{gas}, B_i^{gas}, C_i^{gas}, D_i^{gas}, E_i^{gas}$ are regression coefficients for the component i in the gas phase. $T1$ and $T2$ are the temperature limits

- Heat of vaporization for each individual component is calculated as follows:

$$\Delta H_i^{vap} = A_i(1 - T_{b_i}/T_{c_i})^n \quad (14)$$

Where T_{b_i} is the component's boiling point, A_i, T_{c_i}, n are regression coefficients for the component i.

- The total enthalpy of each individual component is calculated as follows:

$$\Delta H_i = \Delta H_i^{liq} + \Delta H_i^{vap} + \Delta H_i^{gas} \quad (15)$$

The temperature limits used in the liquid phase are T_{ref} and T_{b_i} while the limits used in the gas phase are T_{b_i} and T_s where T_{ref} is the reference temperature, T_{b_i} is the boiling temperature and T_s is the stream's temperature

- The enthalpy of each streams is calculated in the following equation:

$$\Delta H_s = \sum_{i=1}^{n} X_i * \Delta H_i \quad (16)$$

Where X_i is the mass fraction of component *i* in the stream and *n* is the number of components. For simplification, the mixture is assumed to be ideal.

- The heat of vaporization for each stream is calculated as follows:

$$\Delta H_s^{vap} = \sum_{i=1}^{n} X_i * \Delta H_i^{vap} \quad (17)$$

- The enthalpies of the two feeds ($\Delta H_{FC-02,} \Delta H_{FC-04}$), overhead vapour of column C-01 ($\Delta H_{FC-03,} \Delta H_{FC-05}$) and overheads of columns C-03/4/5 ($\Delta H_{FI-07,} \Delta H_{FI-08,} \Delta H_{FI-09}$) are calculated. The bottom product state (temperature and pressure) is used as the reference state thus ΔH_{FC-06} is zero. Table 5*: Streams enthalpies and temperatures* shows the streams enthalpies and the corresponding temperatures.

ID	Stream's enthalpy	Stream's temperature
1	ΔH_{FC-02}	TI-02
2	ΔH_{FC-04}	TI-08
3	ΔH_{FC-03}	TI-03
4	ΔH_{FC-05}	TI-03
5	$\Delta H_{FI-07})_{gas}$	TI-09

6	$\Delta H_{FI-07})_{liq}$	TI-05
7	$\Delta H_{FI-08})_{gas}$	TI-10
8	$\Delta H_{FI-08})_{liq}$	TI-06
9	$\Delta H_{FI-09})_{gas}$	TI-11
10	$\Delta H_{FI-09})_{liq}$	TI-07

Table 5: Streams enthalpies and temperatures

IV. **Energy balance calculations**: overall energy balance is performed to roughly calculate the energy that needs to be provided by the reboiler R-01. The following steps are applied:

- The total heat output Q_{out} from column C-01 can be calculated from the following equation:

$$Q_{out} = \Delta H_{FC-05} + \Delta H_{FC-06} + Q_C \tag{18}$$

Where Q_C is the heat removed by the condenser and calculated as follows:

$$Q_C = \Delta H_{FC-03}^{vap} + \Delta H_{FC-05}^{vap} \tag{19}$$

- The total heat input Q_{in} to column C-01 can be calculated as follows :

$$Q_{in} = \Delta H_{FC-02} + \Delta H_{FC-04} + Q_R + Q_{H-01} + Q_{H-02} + Q_{H-03} \tag{20}$$

Where Q_R is the heat added by the reboiler R-01. Q_{H-01}, Q_{H-02}, Q_{H-03} are the heat added by the integration heat exchangers HEX-01/02/03 and calculated as follows:

$$Q_{H-01} = \Delta H_{FI-07})_{gas} - \Delta H_{FI-07})_{liq} \tag{21}$$

$$Q_{H-02} = \Delta H_{FI-08})_{gas} - \Delta H_{FI-08})_{liq} \tag{22}$$

$$Q_{H-02} = \Delta H_{FI-09})_{gas} - \Delta H_{FI-09})_{liq} \tag{23}$$

- For the energy balance to close, the total heat input to column should equal the total heat output from column. Thus reboiler's duty can be calculated from the following equation:

$$Q_R = (\Delta H_{FC-05} + \Delta H_{FC-06} + Q_C) - (\Delta H_{FC-02} + \Delta H_{FC-04} + Q_{H-01} + Q_{H-02} + Q_{H03}) \tag{24}$$

  - The theoretical amount of steam $\dot{m}_s$ supplied to the reboiler is calculated from Equation (17). Where Q_R is the reboiler's duty and Ls is the steam specific latent heat.

$$\dot{m}_s = \frac{Q_R}{Ls} \tag{17}$$

- The theoretical amount of steam is trended in Seeq workbench and is compared to the actual steam consumption as shown in Figure 25: Steam consumption model comparison to the real consumption

4.4.2 Mass and Energy Balance Results

The mass balance calculations show that the inlet and outlet mass flows for column C-01 are not balanced with an in-out mass difference M_{diff}^{daily} of nearly 0.05-3%. This mass difference suggests flowmeter inaccuracy.

Since the measured values are subject to measurement errors, process lags and delays, it is certain that the steam consumption model cannot be perfect or identical to the real steam consumption. A deviation from the actual consumption is expected. Yet, it is still beneficial to compare the model to the actual consumption and look for the periods with high deviation.

A high model deviation from the real consumption is only observed in the period from 9 to 11 March, shown in Figure 25. The model calculated the steam consumption to be 0, while the real consumption was in the normal range. The deviation could have many sources; Valve leakage, valve stuck open, controller misbehaving or model calculation error.

By checking the time series trends and the consumption model calculations, the following findings were reached:

- The controller was able to keep the temperature measured value stable at the set point value.
- The valve opening during this period correctly responds to controller output.
- the hot fluid outlet temperature from the integration heat exchangers TT-05/06/07 were recorded to be 0°C instead of the actual operating temperature which is always higher than 100°C. These reading errors resulted in an error in the calculations as follows; huge amount of heat input via the heat exchangers to the column was calculated so the steam's heat input calculated from the heat balance equation was 0.

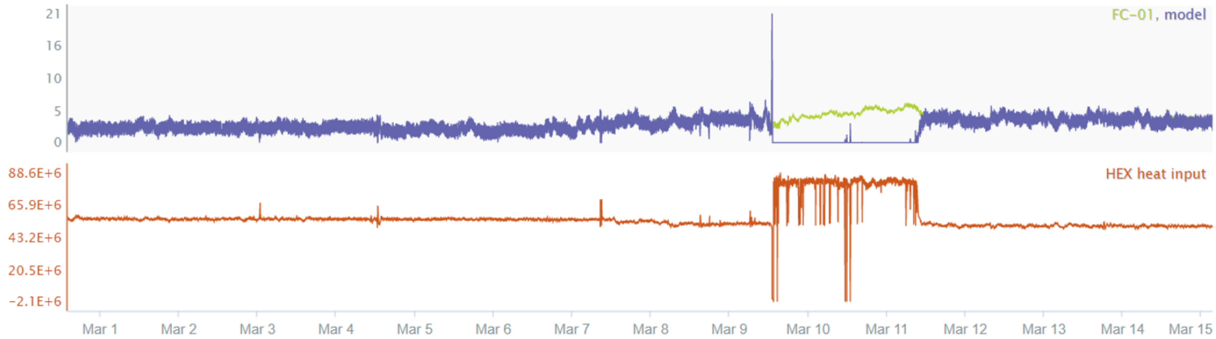

Figure 25: Steam consumption model comparison to the real consumption

4.5 Process Design Assessment

One of the common reasons resulting in operation problems is the poor process design. In this section we investigate the design problems affecting the operation.

4.5.1 Integration Heat Exchangers

As the column is thermally integrated with three other columns C-03/04/05, the temperature control loop performance is affected by the amount of heat provided to the column by the integration heat exchangers HEX-01/02/03. It was expected that the bottoms flow rate to the integration exchangers is somehow controlled to make better use of the changing heat available in the heating streams from other columns. However, the operators and process engineers in the plant mentioned that there are no automatic or manual control valves on the line going from the column C-01 to the heat exchangers. There is no control also on the lines of the hot fluids (overheads from columns C-03/04/05) supplied to the heat exchangers. As a result, controlling the temperature inside the column is problematic.

Controlling the column temperature becomes more problematic in case of total dependence on integration heat exchangers to provide heat to the column. The operators mentioned that in this situation they change the operating conditions in columns C-03/04/05 to increase their overhead flows. Accordingly, the heat provided increase.

Figure 26 shows the period from June 5th to June 7th during which the operators used the integration heat exchangers to provide heat instead of steam. The temperature time trend showed that the temperature from June 5th to June 7th was out of control and exceed the upper control limits.

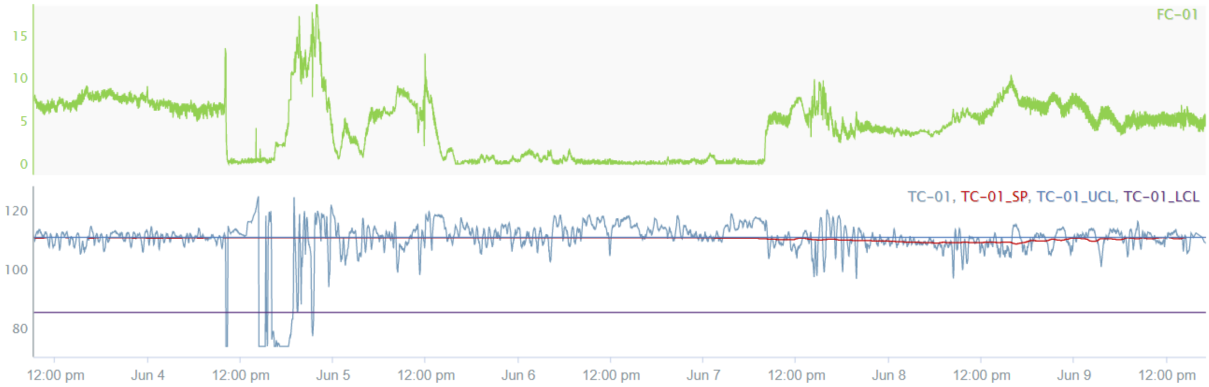

Figure 26: Providing heat through integration Heat exchangers only

4.5.2 Bottom pressure control

The correlation analysis results, Figure 18, showed that the column's top pressure is well controlled but the bottom pressure inside the column changes with the feed changes. An increase in the feed will result in increase in the bottom pressure PI-02 (at the control tray location). Column pressure at the control tray location is known to affect the boiling point of the mixture and the control temperature, making the temperature control troublesome. An increase in pressure will be interpreted by the temperature control as a rise in heavy components concentration and will be counteracted by increasing the boil-up (increasing steam) (Kister, 1990).

4.5.3 Control Tray Temperature

The control tray should be selected to give a good indication of composition, i.e., the correlation coefficient between the tray temperature and the composition should be higher than 0.9. The correlation analysis results displayed in Figure 18 and Figure 22 show a correlation of -0.51 and -0.65 between TC-01 and AI-01, during these analysis periods the APC is in mode 1. Both coefficients imply that, the control tray is not properly selected to indicate composition. The smaller correlation coefficient in the first correlation period could be due to the more frequent changes in the column's differential pressure which affects the mixture boiling point and decrease the ability of temperature to indicate composition.

In Figure 20, the correlation coefficient between TC-01 and AI-01 is -0.27. During this correlation period, the APC is in mode 2, the temperate controller's set point is constants and the APC uses only the reflux flow to control the composition. So the composition changes while the temperature does not.

4.6 Control performance assessment

In this step, the time series data is used to assess the current status of the control system of the column C-01 and to detect poor performing control loops. The assessment of the APC control and the base layer control performance is discussed in chapters 5, 6

5. APC Control Assessment

As mentioned in Chapter 3, an APC system with embedded MPC is applied as an advanced control level besides the base control level to control the distillation unit operation. Accordingly, the first step to be done in this chapter is to assess the APC performance. The performance monitoring and evaluation of the APC is carried out through visual inspection to detect abnormal control performance and using graphical measures of performance such as constrains analysis to detect the violation of constraints.

The assessment includes identifying the abnormal conditions and the correspondent inadequate performance, the percent of time the APC works in automatic mode, the percent of time the constraints are violated and the APC's response to step changes in the controlled variables targets.

The period to be studied is six months long and was chosen to start with the date when the APC was put into operation (January 1st, 2022) and end with the start date of this work (June 30th, 2022). For reasons of confidentiality, the values of the process variables and the time shown on the plots are modified values, not the original values obtained from the plant.

The control performance of the APC can be evaluated from the trends of manipulated and controlled variables during the period of study. Before looking at the time trends of the process variables, the operating status of the APC was investigated to know how often it is switched between automatic and manual modes.

In the time period of six months, the APC is frequently switched between different operating modes as shown in Figure 27. The operating modes description and the percent of time for each mode are shown in Table 6.

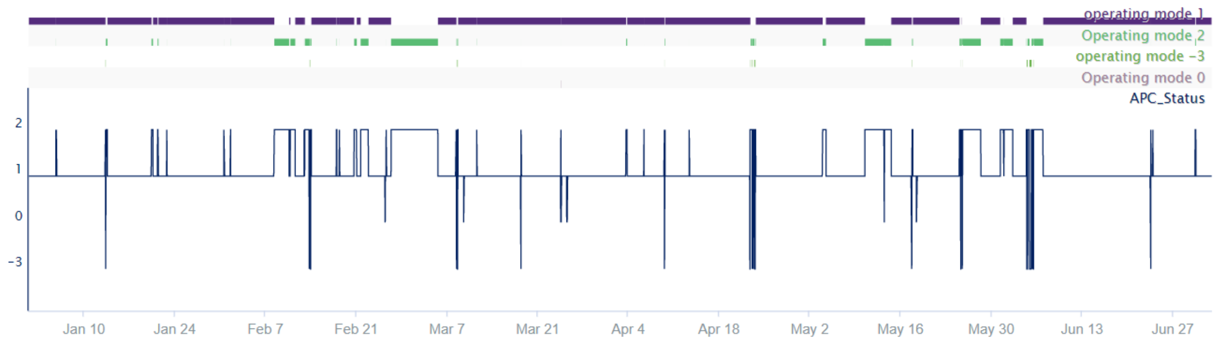

Figure 27: The APC status over a period of 6 months

APC operating mode	Automatic /manual	Duration Percentage %
Mode 1	Fully automatic (all APC inputs and outputs are in automatic mode)	85.127
Mode 2	Partially automatic (the optimizer continues to calculate the controlled variables targets, the APC selects the set point for reflux and bottom flow controller but keeps the temperature set point constant at the last calculated value when the APC was in operating mode 1.	14.294
Mode 0	APC switched off	0.011
Mode -3	Limits are closed (upper and lower limits for each variable are equal)	0.5684

Table 6: APC operating modes duration over the 6 months period in Figure 29

As steam flow is controlled by the temperature controller TC-01, no doubt, the operating characteristics of each mode, especially the controllability of the temperature controller's set point, have an impact on the steam consumption which is the main concern of the plant personnel. So, the characteristics of each operating mode and its impact on the steam consumption is discussed in section 5.2. In the following section we discuss the possible causes for switching between operating modes.

5.1 Possible Causes for Switching between Operating Modes

Switching the APC between different operating modes can be due to many reasons. Among the possible reasons could be process disturbances or operating changes that the operators had made causing plant dynamics to change and accordingly unsatisfactory APC performance in full automatic mode. The dissatisfaction can also be the result of plant dynamics that are significantly different from the model's expectations such that the model is not representing the plant well and its quality is degraded. Other reasons for unsatisfactory APC performance can be soft sensors errors, sticky valves, unreliable sensors, poor tuning and sluggish response of the base level PIDs. In this section the possible causes for poor performance are discussed.

5.1.1 Load Disturbances

One possible reason for APC instability is frequent disturbances. These disturbances can be operating condition changes, load variations, feed composition variations. The most pronounced and frequent measured disturbance is the feed flow rate. So the APC status and controlled variables' targets and process values are checked against load variation to see how these variations affect its operability.

Figure 28 displays the feed flow and the APC's status time trends. The feed flow to the column has changed several times. The changes are often accompanied by switching of the APC to operating mode 2 or -3.

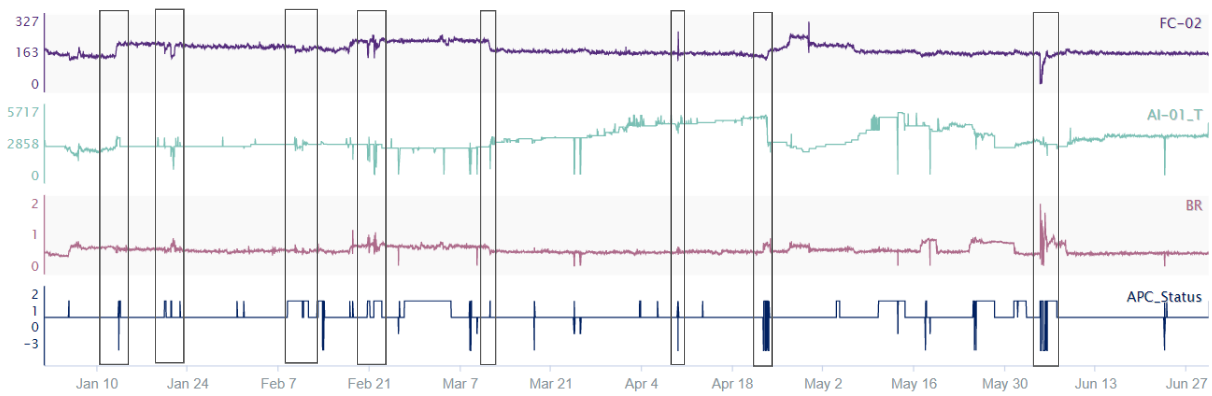

Figure 28: APC performance with load variations

The APC is switched to operating mode 2 or -3 due to feed flow disturbances observed that are accompanied by rapid changes in controlled variable target values. The model predictive control (MPC) feasibility is known to be lost with such rapid target changes. The loss of feasibility means the inability of the MPC to concurrently reach the new target and satisfy all process constraints. In this case, the optimization problem leads to non-admissible terminal region (unreachable set points) or leads to set points that causes constraint violations.

To come over the infeasibility, the controller switches to mode 2 or mode-3. The operating mode -3 prevents movement of the manipulated variables and the constraints are closed. The operating mode 2 tolerates deviation from the controlled variables' target, a relaxation of the reflux flow and bottom flow constraint but rejects the new temperature's set point and keeps it constant as it is not allowed to exceed constraints for safety reasons.

The MPC frequent switches to mode 2 and its inability to achieve the targets and keep constraints at the same time (feasibility loss) in case of disturbances means that the MPC model is not robust enough to deal with uncertainties and disturbances that impact the dynamics. As mentioned in chapter 2, Dealing with frequent disturbances is a main cause for upgrading the control system and using an advanced process control (APC) control. However, the advanced control system in the unit under study has a problem with robust tracking of target sets in presence of disturbances.

5.1.2 Operational Changes

The MPC allows the operators to determine the PID controllers' set points when operating in mode2. As a result, operators intentionally switch the APC to mode 2 in cases of large operating changes such as in case of start-up or total dependence on the integration heat exchangers to provide heat to the column (no steam is supplied to the column). Figure 29, shows switching to mode 2 because of start-up (area 1) and steam cut off (area 2).

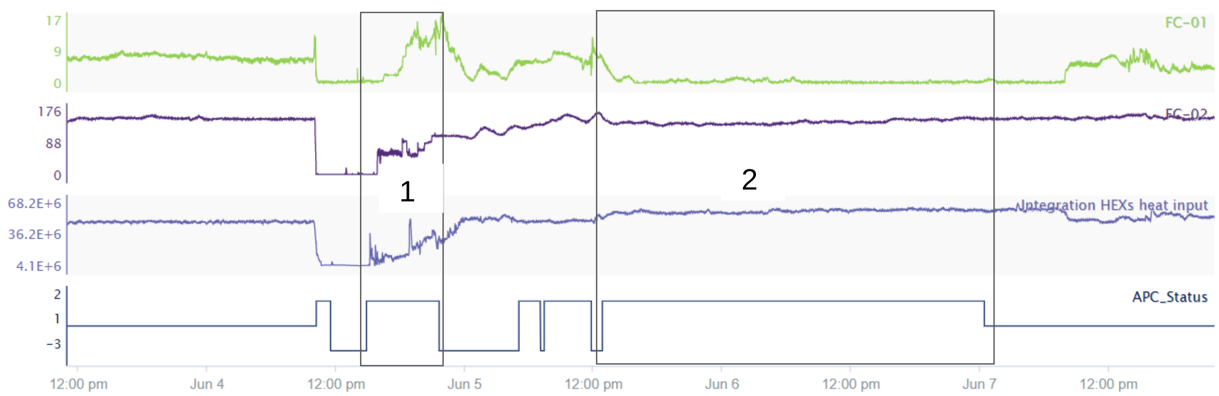

Figure 29: Unit start-up and steam cut off

5.1.3 Model errors

A good quality or plant-representative MPC model is essential to ensure good control performance. However, it is very common for models to have errors that lessen the model's quality and ability to work efficiently. Model errors can be the result of mass non-balanced systems which is the case in the system under study. The distillation column has a 0.05 to 3 percent mass difference between the input and output. It can also result from sensors' measurements errors which is likely a contributing factor here as well. One more cause that reduces the model accuracy is re-tuning the base level controllers but not compensating for it by re-tuning the APC. The plant personnel confirmed that they have re-tuned the temperature loop many times but did not mention re-tuning the APC.

5.1.4 Soft sensor errors

The APC uses soft sensors (inferential sensors) to predict the acetone composition in the overhead from other available and continuous measurements in the plant. Errors in soft sensor predictions will result in poor control performance so the predicted values are compared to the actual measured values and the deviation is calculated and plotted in Figure 30. From the figure it is observed that estimates of the composition are very close to the composition analyser results.

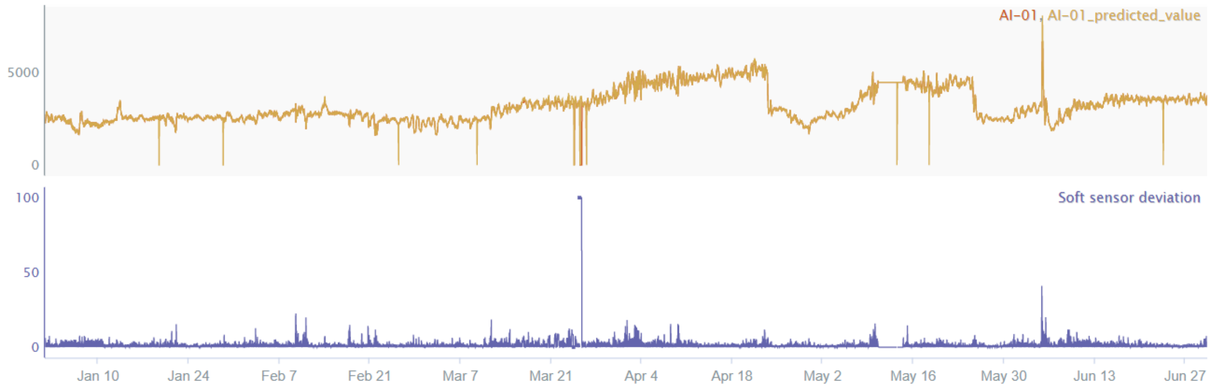

Figure 30: Soft sensor deviation

The percent of deviation of the soft sensors prediction and the percent of time of each deviation is presented in *Table 7*.

Soft sensor deviation %	Percent of time %
0 - 2	90.0
2 – 4	7.95
4 - 6	2.05

Table 7: Soft sensor deviation distribution

For only 90 percent of the study period, the sensor has a deviation between 0-2 percent. Such a result is expected as the sensors depends on different process measurements to calculate the predicted composition value and the process measurements, as mentioned before, are questionable.

5.1.5 Base layer control poor performance

Another important reason for poor APC control performance can be poor base control layer performance. The performance of this layer is discussed in chapter 6.

5.2 Impact of Different Operating Mode on Steam Consumption

As the main concern of the plant personnel is the disturbances and oscillation in steam consumption, the steam consumption in operating modes 1 and 2 and the impact of switching between modes on consumption are discussed in this section.

5.2.1 Operating Mode 1

When the APC is in mode 1, controller inputs and outputs are all in automatic mode. In this mode, the APC is tracking the controlled variables targets and its output determines the slave controllers' set points.

The process under study is characterized by changing the controlled variables targets and accordingly the slave controllers set points are changing as shown in Figure 31. The constraints of the manipulated variable are not allowed to deviate in this mode. These frequent changes make tracking targets for the constrained MPC problematic and can lead to feasibility loss.

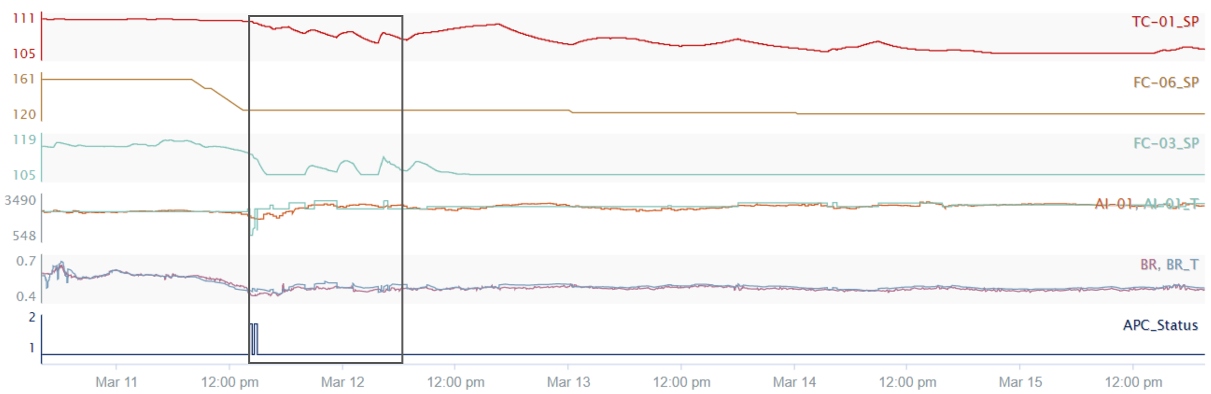

Figure 31: Controllers changing set points during APC Mode 1

5.2.1.1 Steam Consumption When Operating in Mode 1

Acetone composition target is calculated by the optimizer to achieve the optimum operating and economic conditions. In operating mode 1, the APC is controlling the target which is frequently changing. The changes in target value are highly likely to contribute to steam usage disturbances.

To investigate the effect of changing targets, the period from the 30th of March to the 6 of April was chosen for study. During this period, the composition target was initially only exhibiting small changes before a period of frequent notable changes in short succession followed by target stabilization as shown in Figure 32. The APC was in automatic mode (mode 1) except for four hours between April 3rd and 4th when it was in mode 2. Due to this variation in set point, this period is suitable to compare how steam consumption is affected by changing the composition target.

The figure shows that temperature, reflux flow and steam flow controllers' set points exhibit high oscillation with highly changing targets (area 2). Steam is less oscillating when the target is constant (area 3) or slightly changing (area 1).

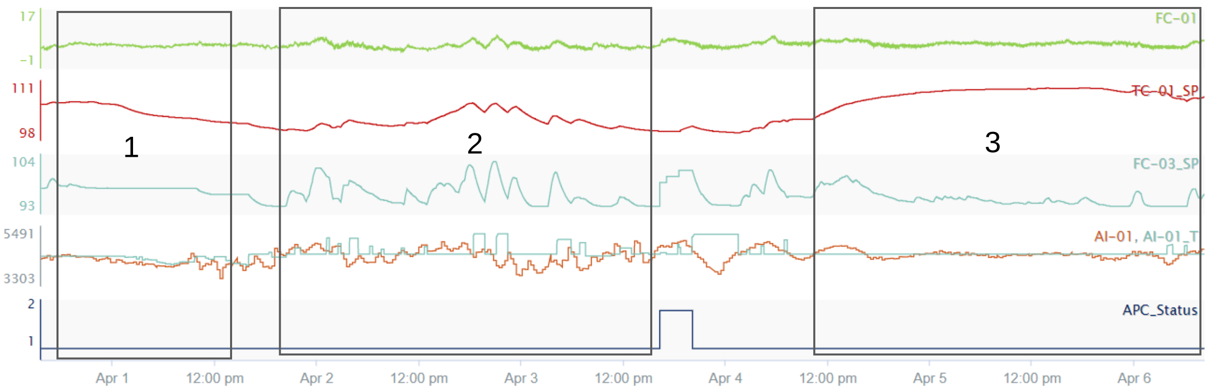

Figure 32: Steam consumption in operating mode 1

5.2.2 Operating Mode 2

When operating in mode 2, the APC determines new set points for the reflux and bottoms flow but keeps the temperature set point constant (except in the event of operator intervention). As temperature is the main manipulated variable that determines the acetone composition, keeping the temperature out of the APC control for a period of time will result in deviation from the composition target. The bottoms flow is the manipulated variable responsible for achieving the bottom ratio target. As it is kept under the APC control, the bottom ratio does not deviate. Also, this mode relaxes the constraints for the reflux flow and the bottom flow, allowing the APC to determine their set points out of the limits.

A closer look at the performance when the APC in mode 2 is presented in Figure 33. The figure shows a period of 15 days. During this period the APC switched to mode 2 several times. We can see that the bottoms ratio target is still achievable as the manipulated variables set point (FC-06_SP) changes over time , the deviation from target is observed in the composition (AI-01) as the temperature set point is kept constant and the APC is trying to achieve the composition target by manipulating the reflux flow only.

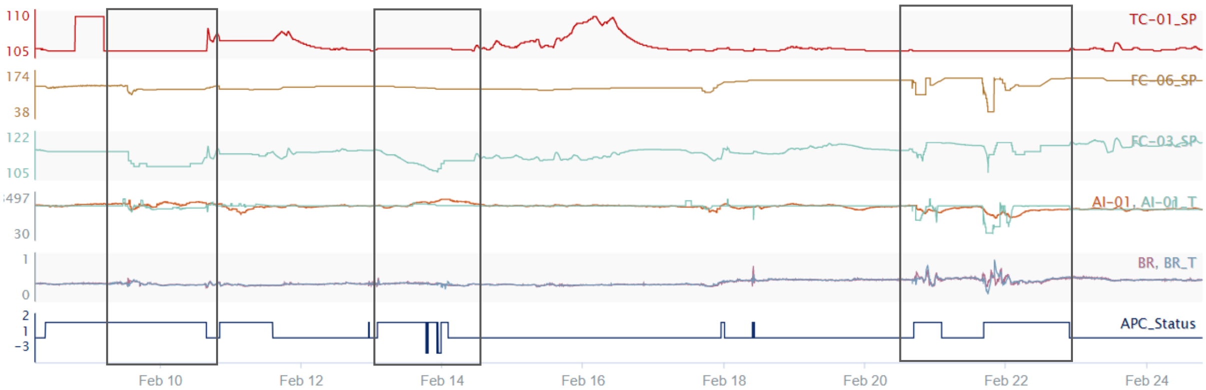

Figure 33: Operating mode 2 performance

As a result, the composition deviation obtained when operating in mode 2 is significantly higher than that obtained when operating in mode 1 as shown in Figure 34.

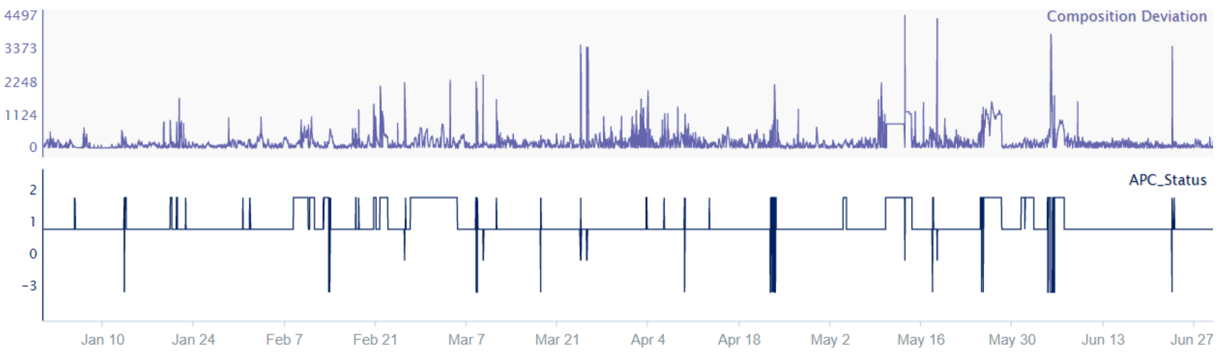

Figure 34: Composition deviation from target for various operating modes

The average deviation is calculated in each mode, the results are shown in Table 8.

	Operating mode 1	Operating mode 2
Deviation from composition target %	4.2	18

Table 8: Deviation from composition target

5.2.2.1 Steam Consumption When Operating in Mode 2

In this section we study the effect of mode 2 control characteristics on the steam consumption. The steam consumption is mainly determined by the control tray temperature TC-01, the reflux flow FC-03 and the composition target AI-01_T. So the focus of this study will be on these variables and their impact on steam.

The deviation from composition target can be positive or negative. Positive deviation means the temperature is less than the required temperature to achieve the target and less than required steam is supplied to the column. Negative deviation means the temperature is more than the required temperature to achieve the target and more than required steam is supplied to the column.

An example to explain the inefficient steam consumption when operating in mode 2 is shown in Figure 35. The figure shows the negative deviation of composition from the target while the APC was in operating mode 2. The deviation resulted as follows. On May 25th, the measured composition value started to deviate above target. To decrease the composition, the temperature set point should be increased. The temperature set point was already close to the upper limit so an increase means that the temperature set point would exceed the limit. The APC switched to mode 2, the temperature set point was kept constant and the reflux set point increased to enhance separation (area 1). Then, the composition of acetone started to decrease, the calculated temperature target shows that the temperature set point should have been decreased. Instead it was kept constant at a value notably higher than the optimal temperature target. The measured composition value decreased below the target, for the APC this deviation indicates a high operating temperature and the column needs to be cooled down. The APC increased the reflux flow to decrease the operating temperature and, at the same time, the temperature controller (TC-01) tried to keep the temperature close to the set point. TC-01 increased the steam supplied to the system (area 2). The excess steam supply continued for three days.

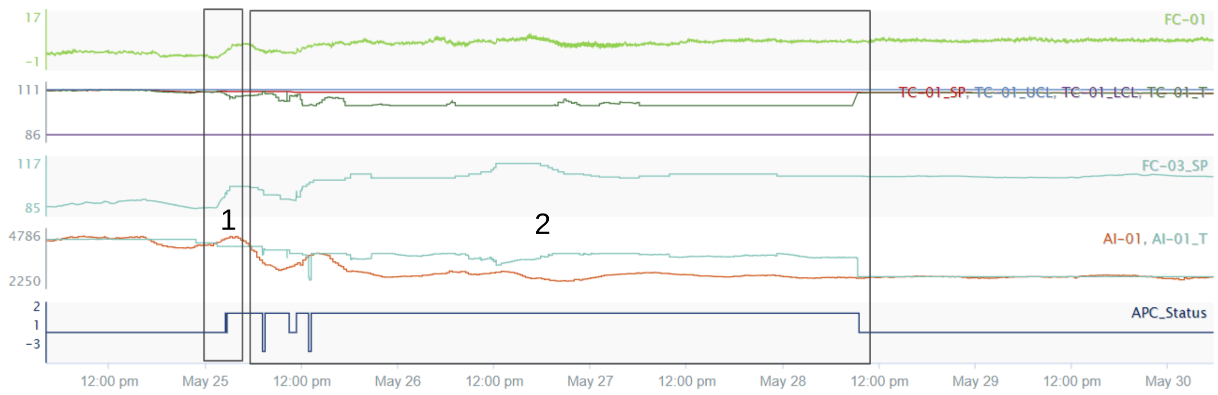

Figure 35: Steam consumption in operating mode 2

5.2.3 Effect of Mode Switches on Steam Consumption

As explained in previous sections, operating mode 1 generally shows little deviation between target and measured composition. However, mode 2 is generally accompanied by deviation from the target. In this section, the impact of switching between the two modes is explained.

A period emblematic of this problem is used to describe in detail how switching between operating modes affect the steam consumption and is presented in Figure 36. The figure shows a period of 4 days. In these days the APC was switched to operating mode 2 twice. As the bottoms ratio and bottoms flow are not affected by switching between modes, we focus in this example on the composition target, temperature and reflux flow.

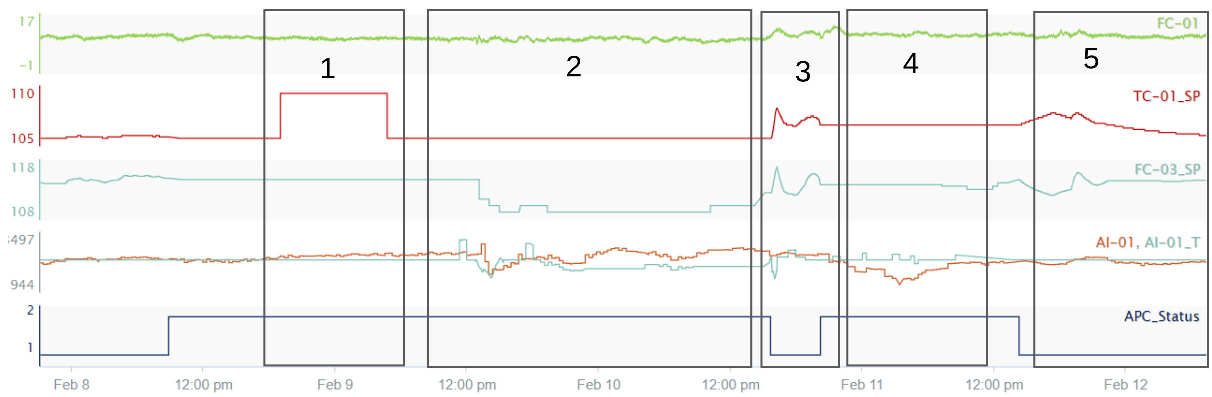

Figure 36: Effect of APC mode switches on steam consumption

In the first period, the operator was monitoring the composition measurement versus the target. When the measured value stared to deviate, the operators increased temperature set point then returned it back to its initial value (area 1). The deviation continued to increase especially after the optimizer changed the target value and the APC controller decreased the reflux flow due to low control temperature (area 2). When the APC was switched back to operating mode 1, it took the responsibility to achieve the composition target. To eliminate the deviation and get closer to the target, the APC increased the set points and accordingly increased to steam supplied to the system (area 3).

When the APC was switched to operating mode 2 for a second time, the composition started to decrease below target. The temperature set point was kept unchanged which led to high deviation in composition. This high negative deviation means more steam than required was supplied to the system during this period. After a few hours the APC decreased the reflux flow set point and the deviation was eliminated (area 4). When the APC was back to operating mode 1, it increased the temperature set point and steam consumption. Two steam consumption peaks can be observed in area 5.

From the figure we can conclude that the switch between different modes results in steam consumption fluctuation as the APC always needs to take action on temperature set point when it is switched to operating mode 1. This action affects the steam required for the system.

After investigating switching between the operating modes, its causes and its effect on steam consumption, the APC control performance is further assessed to evaluate its ability to keep variables within their control limits and achieve the targets when it is turned on and put in full automatic mode (mode 1).

5.3 Constraints Analysis

Controlled and manipulated variables constraints are the thresholds defining the boundary region for process operation. In well controlled operation the process value should not exceed the upper or lower constraints. When the constraints are violated, the process is said to abnormal or not well controlled.

One method to assess the control performance of this system is to plot the data of manipulated variables (the control tray temperature, the reflux flow rate and bottom flow rate) and controlled variable (acetone composition and bottom ratio) during a period in which the APC was in operating mode 1. The variables can then be checked against their boundary limits (LCL: lower control limit and UCL: upper control limit).

The data for upper and lower limits for the bottoms flow rate could not be found in the data set used in this study. The other four process variables are plotted in a Shewhart chart shown in Figure 37. The variables plotted in the period from January 21st to February 8th. During this period, the APC was largely in operating mode 1 with minimal switching to mode 2. The composition target exhibits a step change that allows for studying the constraint violation and associated controller response.

Figure 37: Shewhart chart for manipulated and controlled variables

To provide better visualization that allows for studying the constraints violation, the Figures Figure 38, Figure 39,Figure 40 and Figure 41 take a closer look at each variable during the same period from January 21st to February 8th. During this period, the APC was largely in operating mode 1 with minimal switching to mode 2. The composition target exhibits a step change that allows for studying the constraint violation and associated controller response.

Figure 38 shows the acetone composition data AI-01, AI-01 upper and lower control limits (AI-01_UCL and AI-01_LCL) and AI-01 target (AI-01_T). In the Figure, we can see that the AI-01 measured value exceeds the upper limit. The target value is set at the upper limit and the process value is oscillating around it. At several points in the study period, the target value exceeds the upper limit.

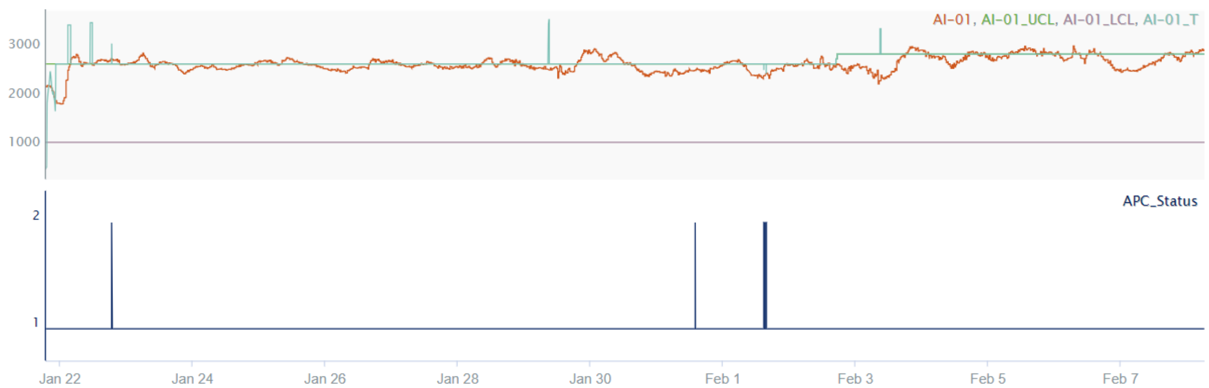

Figure 38: Shewhart chart for AI-01 –Jan

Figure 39 shows the bottoms ratio data (BR), upper and lower control limits (BR_LCL, BR_UCL) and control target (BR_T). It can be observed that the target value and the process value exceed the lower limit several times.

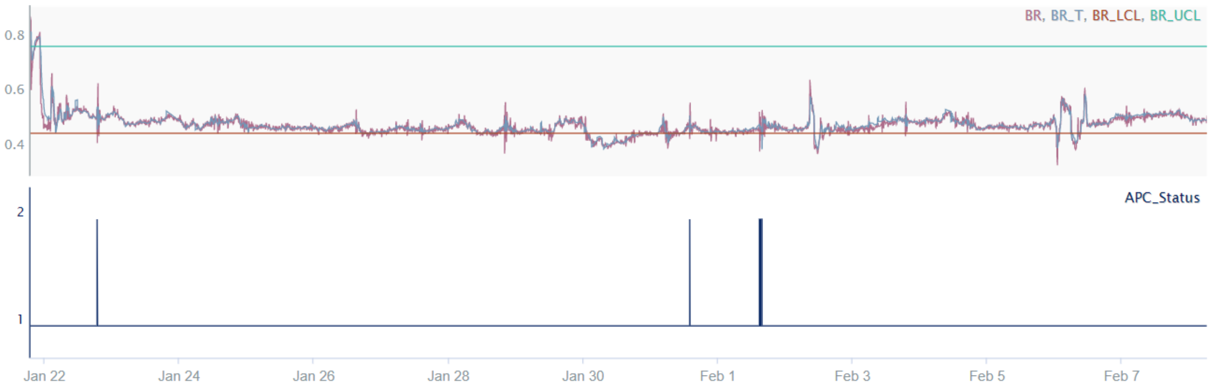

Figure 39: Shewhart chart for Bottom ratio

Figure 40 shows Shewhart chart in which the temperature data TC-01, TC-01 upper and lower control limits (TC-01_UCL and TC-01_LCL) and TC-01 target (TC-01_T) are plotted. The figure confirms the control tray temperature violation for both upper and lower limits.

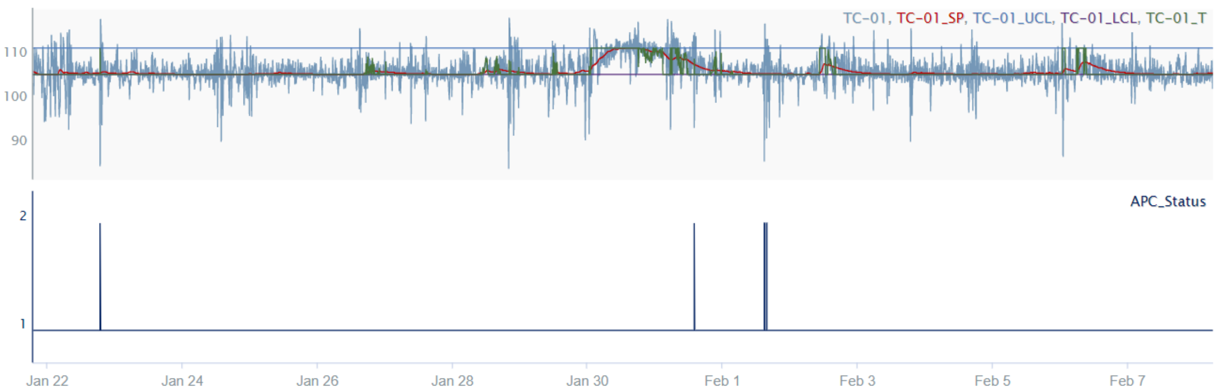

Figure 40: Shewhart chart for TC-01

The reflux flow data FC-03, FC-03 upper and lower control limits (FC-03_UCL and FC-03_LCL) and FC-03 target (FC-03_T) are plotted in Figure 41. In Figure 41, FC-03 measured value exceeds the limits while APC is in mode 1 in the area highlighted

Figure 41: Shewhart chart for FC-03

From the figures, it is obvious that process values of all variables exceed the constraints although the APC operates in mode 1. The next step is to calculate the percent of time for each constraint violation in the six months period of study. The results are presented in Table 9. From the results, we can conclude that the APC controller misbehaves and the process is not well controlled.

Constraint violation	Percent of time %
TC-01 > TC-01_UCL	11.72
TC-0 < 1TC-01_LCL	14.586
FC-03 > FC-03_UCL	0.074
FC-03 < FC-03_LCL	5.62
AI-01 > AI-01_UCL	40 .06
AI-01 < AI-01_LCL	0.167
BR > BR_UCL	1.3
BR < BR_LCL	32.62

Table 9: Percent of time for each variable's constraints violation

5.4 APC Response Assessment

A period where a step change in the composition target took place while the APC is operating to mode 1 is chosen to assess the response of the APC to changes.

In Figure 42 an oscillatory control and sluggish response is exhibited when the composition target changes. It takes the composition 24 hours to reach the new target.

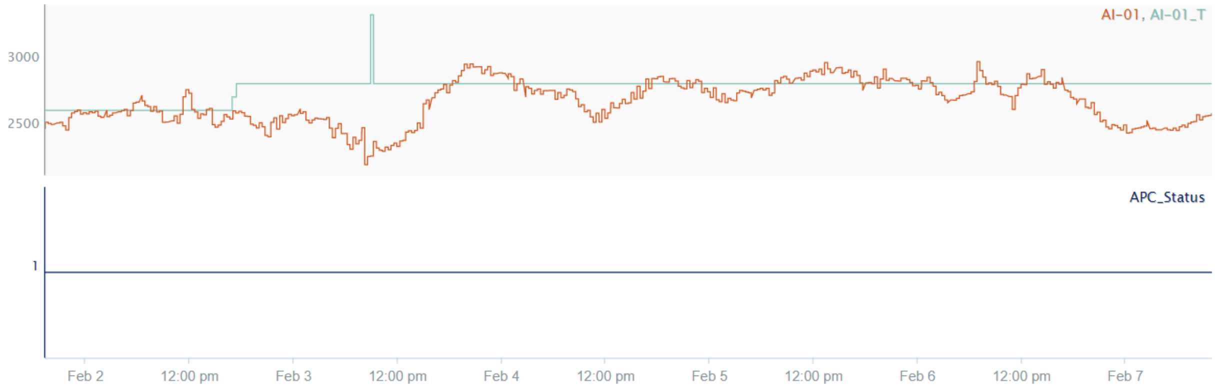

Figure 42: APC response to target step change

6. Base Control Layer Assessment

The advanced control level performance highly depends on the performance of the base layer. The APC cannot work properly if the base layer is poorly performing. The control loop basically consists of sensors, controllers and valves. Failure of any of these components impacts the performance of the whole loop which in return impacts the APC performance. Poor slave controller tuning, oscillation and sticky valves are major contributors to poor APC performance. As the temperature control loop is the highly oscillating loop and is suspected to be the highest contributor to the APC poor performance. In this section the assessment is performed for the temperature control loop and its cascaded flow loop. The assessment for each loop is conducted by manual graphical analysis.

6.1 Temperature Control Loop TC-01

Temperature control loop is suspected to be poorly performing due to the oscillation observed and due to the irregular data reporting frequency. Oscillation is one of the most common problems in control loops. Oscillation can be the result of many things such as high controller gain, oscillating external disturbances, interacting loops and non-linear valves. In this section each of these factors is investigated.

6.1.1 External Disturbance

The heat supplied to the column is provided by two sources; the reboiler's steam and the integration heat exchangers' hot fluids. One reason behind operating problems in temperature controller can be amount of heat supplied by the integration heat exchangers. Heat transfer is not controlled for all heat coming from the columns C-03/03/05 and supplied to column C-01. The operation is as follows; some of the C-01 bottoms flows to the integration heat exchangers. This flow rate is not controlled in the existing column design as verified by plant personnel. The pipe is not equipped with neither control valve nor manual valve. The heating fluid in these heat exchangers are the overhead vapours from columns C-03/04/05. As the feed flow to column C-01 is changing, these vapor amounts from the following columns also change with significant lag resulting in changing amounts of transferred heat. Besides the change in the amount of heat supplied, the HEX heat input trend shows oscillation which can contribute to temperature oscillation. Control tray temperature and heat input for the heat exchangers in question are plotted in Figure 43.

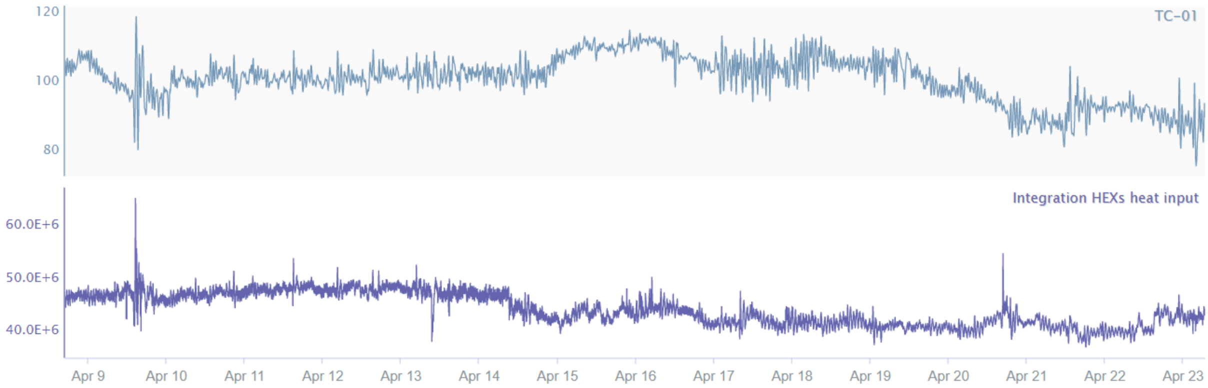

Figure 43: TC-01 external oscillation

6.1.2 Control loop tuning

Aggressive loop tuning is known to be a very common reason for loop oscillation. The plant operators have tried several times to re-tune the temperature control loop to come over the oscillation problem but were not successful. The oscillation is more likely to be caused by reasons other than aggressive tuning.

6.1.3 Interacting Loops

The distillation column is characterized by interacting control loops. Oscillation in one loop propagates to the other loops. In the data set we have, oscillatory behaviour is present in other control loops not only in the temperature control loop. An example is shown in Figure 19 where TC-01, PC-01 and LC-02 exhibit oscillation. The correlation analysis, Figure 20, showed that the oscillation in TC-01 proceeds PC-01 and LC-02. The causality analysis conducted on the same period, Figure 23, shows causal relationship between PC-01 and TC-01 where PC-01 is the cause signal. However the analysis conducted on the period from March 10th to 15th, Figure 24, shows bi-directional relationships between TC-01 and PC-01 and LC-02. Due to the different results obtained from analyses conducted on different periods, a root cause analysis is recommended to identify the root cause of oscillation.

6.1.4 Valve Stiction

The temperature controller is the master controller in the flow - temperature cascade control. Oscillation in the temperature control loop can be the result of FC-01 control valve leakage or stiction. So FC-01 control valve performance is assessed in section 6.2.

6.1.5 Data Reporting Frequency

Beside the operating temperature oscillation, there is also a change in the temperature data reporting frequency can be seen in Figure 44 and it can be heavily contributing to the poor performance.

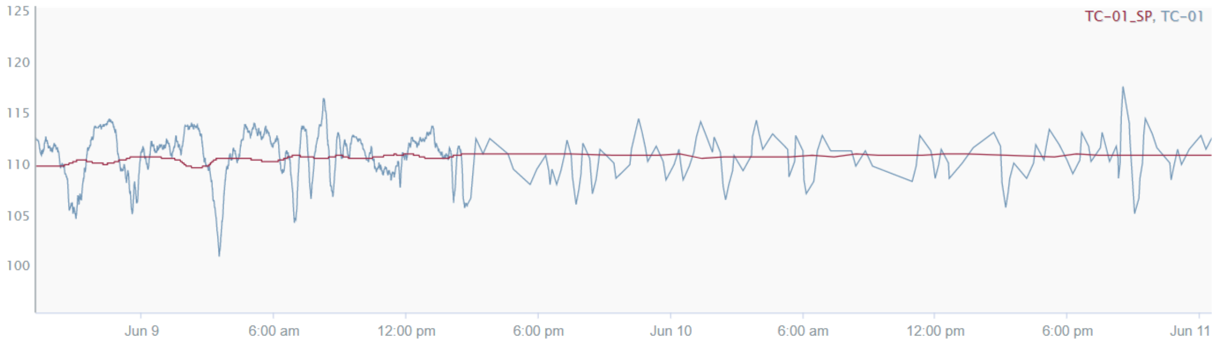

Figure 44: Temperature data reporting frequency

6.2 Flow Control Loop FC-01

The steam flow slave loop FC-01 is also oscillating. This oscillation can be external due to its interaction with the oscillating temperature control loop or can be internal due to loop components malfunction. The most common cause for flow loops oscillation is valve stiction. In this section, the internal oscillation due to a sticky valve is investigated.

6.2.1 FC-01 Control Valve Stiction Detection

Stiction is the resistance to motion, the valve is not able to move in case of small changes of controller output and moves suddenly when the controller output is high enough to move the sticky valve. The valve stiction can be detected by plotting the process variable against the controller output as shown in Figure 45. The plot reveals that the loop has non-linearity problem. The horizontal lines in the figure indicates that the steam flow is constant although the controller output changes. When the change is large enough to move the valve it moves to a new position.

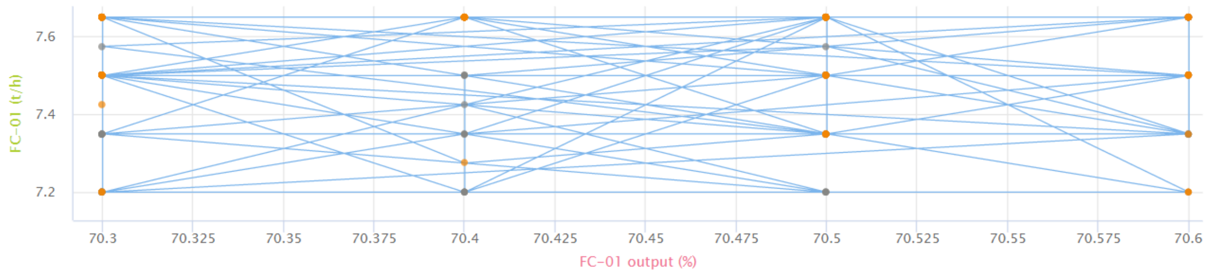

Figure 45: FC-01 control valve PV-OP plot

The valve performance can also be visually assessed from the time series trends of the steam flow rate and the flow controller output. The valve performance is assessed during the unit shutdown when the controller output is 0, followed by the unit start-up when the controller output starts to increase. The time trends are plotted in Figure 46.

The figure shows that there is a flow in FC-01.PV even though the controller says 'fully closed'. This can be explained as a valve stiction such that the valve is stuck at a position that is not fully closed even though it is reported as closed. This can also be explained as backlash or steam leakage either through the valve or from the pipeline. However, Stiction is more likely though because of the oscillation in the right hand side. When the operators start to open the valve the steam flow did not change. There is a resistance to motion in the beginning till the operators increase the opening past a breakoff amount and the valve moved to a new position.

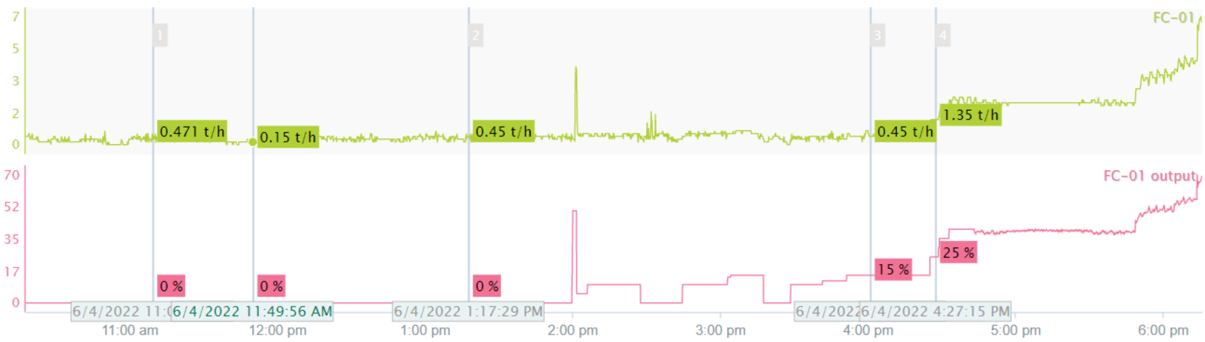

Figure 46: steam flow and flow controller output time series trends

Figure 47 shows another valve problem as we can see that steam flow rate does not correspond to the valve position. For example, point 4 has a valve opening less than the valve opening in point 5, however more steam is supplied. Some plausible reasons for that could be change in steam supply pressure, valve tuning, or flow transmitter noise.

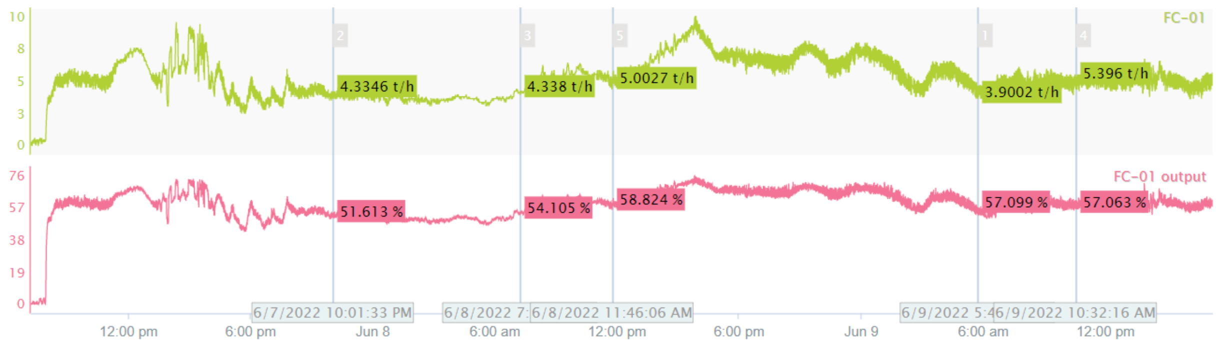

Figure 47: Steam flow correspondence to vale opening change

From the assessment of the temperature and steam flow loops we can conclude that they are poorly performing and this poor performance can be heavily contributing to the poor APC performance.

7. Conclusion

This thesis work was intended to study operation of the distillation column controlled by advanced process control (APC), find the problems and suggest areas of improvement. The main problem reported by the plant personnel was the distillation column's steam consumption disturbances and oscillation. The cause adopted by them is the temperature control loop oscillation, they thought the problem is with the loop tuning and retuned the loop several times but wasn't success.

To achieve the aim of this work, the historical data of the column operation was studied to assess the current status of the column, the control system performance and identify the possible causes of steam oscillation and disturbances. The system assessment started with conducting correlation and causality analyses in different operating periods and the results were compared to find the impact of different operating conditions on the relationships between signals. Then, mass and energy balance calculations were performed and the mass balance was found to be dissatisfied. The mass balance dissatisfaction could result in model predictive control poor performance. The steam consumption was modeled and compared to the actual consumption, only one period with high model deviation was found and the main reason was faulty sensors readings.

The APC performance was assessed. It was found that, during the six months period of study, the system worked in full automatic mode (mode 1) for only 85 percent of the time. The rest of the time it switched to partial automatic mode (mode2) in which the temperature set point wasn't part of the APC control and the reflux constraints were relaxed or it switched to mode -3 when the constraints were closed either by the APC or manually by the operators to prevent the process from moving to new set points.

The steam consumption in each mode and the effects of mode switching on steam consumption were evaluated. It was found that when operating in mode 1 the composition target was achieved by manipulating the set points of the reflux flow and the control tray temperature which resulted in a satisfactory composition control. However, due to several composition target changes followed by temperature set point changes, the steam flow to the column exhibited disturbances. When operating in mode 2, the temperature controller was no longer controlled by the MPC. The MPC was achieving the composition target by manipulating only the reflux flow which resulted in deviations from the target and inefficient steam consumption.

The switches between the different modes were found to be accompanied by feed flow changes and sudden changes in controlled variables' targets. The most possible explanation is that, the target changes caused the optimization equation to be infeasible within the identified constraints. Thus, the APC switched to a different mode to relax the tolerable constraints and to reject the new temperature set point and prevent the temperature from exceeding its constraints. The APC frequent switches to mode 2 makes the MPC feasibility and the model robustness to deal with uncertainties and disturbances questionable.

As constraints are the operating limits for each variable that should not be violated when APC is in automatic mode, the ability of the APC to keep constraints can be used to evaluate its performance. Therefore, a constraints analysis was performed, and it showed that all the controlled variables' and manipulated variables' process values exceeded the constraints but for different percents of time which indicates the APC misbehaviour. The response to step changes in the controlled variables' target was also evaluated to be sluggish and oscillatory.

The possible causes of APC poor performance were investigated. The investigation showed that load disturbances followed by rapid target changes and loss of feasibility were highly contributing to poor performance. Model errors were found to be likely contributors to the poor performance in the system under study. Model errors can be the result of mass non-balanced systems, measurements errors or retuning the base control loop controllers without compensating for it by retuning the APC. Other possible causes such as soft sensors and poor base control layer performance were also investigated. Soft sensors prediction showed deviation of more than 2 percent from the measured process value for 10 percent of the time. Oscillation in the temperature control loop was detected and the loop was suspected to be poorly performing. Therefore, the factors that might cause oscillation were investigated. External disturbances caused by heat provided via integration heat exchangers and stiction in the steam flow control valve were found to be the most possible causes of temperature control loop oscillation.

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