Polimeri Europa Olefin Plant Maximizes Benefits from Advanced Solutions

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Abstract

The implementation of advanced solutions at the Polimeri Europa ethylene plant in Porto Marghera, Italy has realized significant, tangible economic benefits and received a high degree of acceptability from the plant personnel. The project's success is a direct consequence of continuous dialogue and interactions between plant personnel and the project team. As an extension to the modernization and re-instrumentation project at this plant site, twenty-five unit multivariable predictive controllers (MVPCs), a rigorous nonlinear online optimizer, fifteen model-based severity controllers and a plant-wide production controller have been commissioned.

In addition to an overall technical description, this paper highlights several novel features of the advanced control and optimization applications. Of key interest in the production controller design are the concept of emulating the back-end unit MVPC moves, the dynamic distribution of multiple furnace throughput targets, and the manipulation of refrigeration pressures and chill box temperatures to alleviate compressor speed limitations and recovery section heat exchanger constraints. Discussion on the real-time optimizer includes its dynamic flowsheeting capability (to reflect equipment status), the selection of model rigor for major equipment, MVPC shadowing to predict certain constraints, and the use of the offline plant model to provide MVPC gains. The value of model-based severity controllers to control pyrolysis cracking and to infer key constraints for use in the MVPC is explained. Specific detail is provided on the design of the MAPD reactor MVPC strategy, which maximizes propylene recovery against a nonlinear and near-zero impurity constraint.

Introduction

The Polimeri Europa Ethylene plant in Porto Marghera, Italy was originally constructed in the early 1970's and has been revamped several times to increase production capacity to its current level of 400 KTA of polymer grade ethylene. In 1999, the plant underwent a revamp during which there was a partial upgrade of field instrumentation. New gas chromatographs were added and analog and electronic single-loop controllers were substituted by a Distributed Control System (DCS). As an extension to this modernization effort, Polimeri Europa decided to implement advanced automation technologies that would position them to achieve both economic and operating benefits over the long-term and justify their capital investment.

Process Overview

The Porto Marghera ethylene plant has fifteen cracking furnaces. Fourteen of the furnaces are ABB Lummus-designed SRT-IV type furnaces with four radiant coils and two transfer line exchangers (TLEs). One furnace is a circular type heater designed by Montedipe; it has eight radiant coils and three TLEs. All fifteen furnaces can treat virgin naphtha whereas five of the furnaces can also treat ethane as a single feed or in split cracking with the virgin naphtha. With the exception of the eight-coil furnace, all furnaces have both wall and hearth burners. Hydrogen and methane recovered from the product purification processes are the main furnace fuel sources. The TLEs at the radiant coil exits cool the hydrocarbon effluents and utilize the heat to generate high-pressure steam, which drives major compressor turbines in the recovery section.

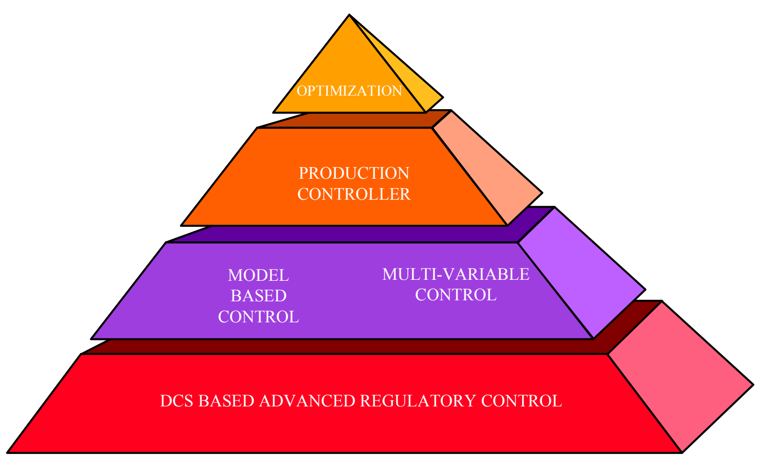
High-level heat is also recovered by a circulating quench oil system in the gasoline fractionator which separates out the pyrolysis fuel oil from the furnace effluents. The hot overhead vapors from the gasoline fractionator are then sent to the quench tower where additional heat is recovered by direct counter-current contact with recirculating water. The hot recirculating water from the quench tower supplies low level heat to the process, including the feedstock preheaters, and recovery tower reboilers. The quench tower overhead hydrocarbon vapors are compressed in five stages by steam-driven centrifugal compressors with interstage cooling. Under design rated throughput conditions, the plant operates two compressor trains in parallel each with its own set of suction drums. The compressed, cracked-gas is dried and then progressively chilled in an integrated heat exchange system (chilling-train) which involves exchange with various levels of propylene and ethylene refrigerants as well as cold process fluids. The cracked gas streams then pass through a series of binary and multi-component distillation columns where the desired products are separated. Two hydrogenation reactor systems convert acetylene and MAPD in the cracked gas primarily into ethylene and propylene respectively.

The propylene and ethylene refrigeration systems are closed multi-stage systems and provide refrigeration at different levels to heat exchangers in the product recovery train. Depending on the plant throughput capacity, two propylene refrigeration compressor trains can be operated in parallel.

Technology Overview

Polimeri Europa chose a multi-layered, hierarchical approach (Figure 1) to implementing the advanced solution technologies. Each layer interacts and overlaps functionally with the layers above and below it. At the highest level, a plant-wide real-time optimizer (RTO) determines optimal operating scenarios using a combination of mathematical optimization techniques and a library of high fidelity, steady state process models of the ethylene plant, including a rigorous pyrolysis cracking model. This optimum takes into account feedstock and utility costs, product values and plant-wide constraints. Beneath the RTO, a combination of model-based inferential, multivariable predictive , and DCS-based, advanced regulatory control strategies maintain the plant operation as close as possible to the targets established by the optimizer or by the operator, while smoothly responding to dynamic upsets and operational bottlenecks. Included in the scope of the advanced process control (APC) layers are fifteen model-based severity controllers, a plant-wide production controller and twenty-five unit-specific MVPCs.

Figure 1: Multi-Layered Implementation of Advanced Solutions



Classical advanced control strategies are implemented for the cracking furnaces, reactors and cold section recovery units on the newly installed DCS. These strategies combine traditional features of cascade, feedforward and feedback control in an integrated manner to minimize impact of high-frequency disturbances on plant operation. The control system strategies are designed to be robust so as to maintain high service factors while providing proper speed of response. In many cases, these lower level DCS strategies act as fallback modes for the higher level MVPCs, ensuring proper redundancy in a smooth and bumpless manner.

At the core of Polimeri Europa ethylene plant model is the PYPS pyrolysis cracking technology from ABB Lummus. This model provides prediction of yields throughout the operating envelope and provides the foundation for optimization and model-based control. The model evaluates the performance of cracking furnaces based upon feed properties, physical configuration, operating conditions, and degree of fouling. Performance results include yields, heat balance, and remaining run length. Some significant features relevant to Polimeri Europa's requirements are listed below:

- The pyrolysis-cracking model is available in both the open-equation and closed forms.
- The model was easily tuned to a variety of feedstocks, furnace operating conditions and furnace effluent analysis provided by Polimeri Europa and exhibits acceptable accuracy with respect to yield predictions and trends of key parameters.
- Reduced kinetic representation of liquid feedstocks instead of a fully kinetic formulation significantly reduces the complexity of the model, allowing easy integration into the online advanced applications without compromising performance.

Key technologies selected for the advanced applications are summarized below; their implementation details are discussed in subsequent sections:

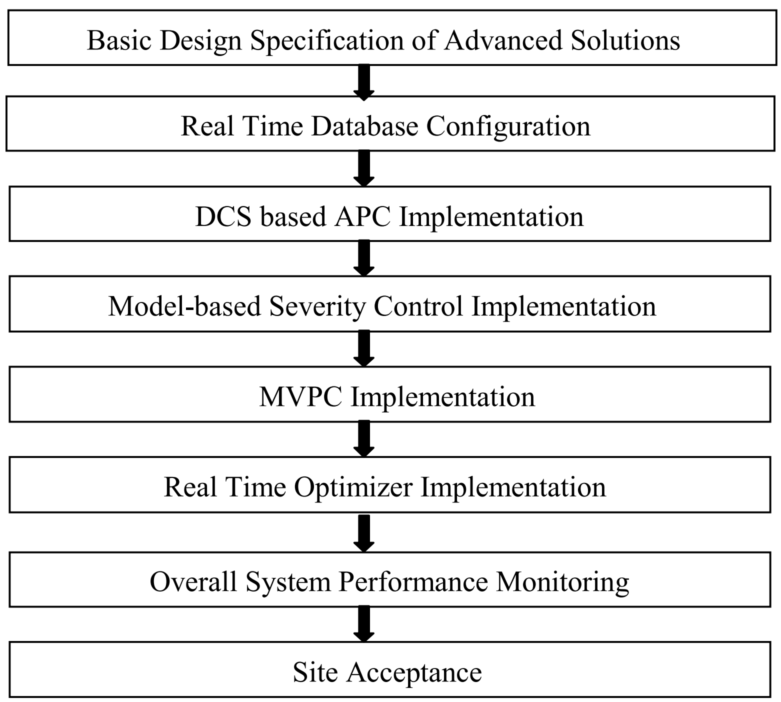
- Model-based severity control uses the online, closed form of the ABB Lummus PYPS pyrolysis-cracking model.
- Multivariable control applications are based on the STAR® technology from Dot Products. These controllers are adaptive in nature in that they re-synthesize their internal model dynamics on each cycle based on process feedback. The controllers incorporate an economic optimizer that runs at each control cycle and use a quadratic program (QP) to simultaneously solve the control and economic objective functions. If degrees of freedom are available, the optimizer uses the costs along with the gain information to find the most profitable operation that satisfies the set point ranges.
- The plant-wide RTO is based on the NOVA® optimization and modeling system from Dot Products. NOVA is designed for the solution of general constrained nonlinear programming problems. It uses advanced successive quadratic programming (SQP) methods and sparse matrix technology to simultaneously solve the ethylene plant process flow sheet for a given objective function and a set of equipment constraints. The open-equation form of the PYPS furnace model is integrated into the plant-wideNOVA model.

Project Execution Methodology

The project team comprised of six engineers assigned to work full time on the project implementation. This included three experienced engineers from the vendors and three from Polimeri Europa who were very familiar with the plant layout and ethylene technology. In addition, two to three Polimeri Europa personnel provided support for system integration, DCS and real time database configuration, and user interface development. The first milestone task for the project team was the development of the basic design specification for all the advanced applications. This specification was to provide an overview of the design, enumerate input data requirements for all the applications and describe in detail the implementation approach. Continuous dialogue and interactions between the project team was a critical factor in the development of a design basis that received a high degree of acceptability from the plant personnel and ensured efficient project execution. As shown in Figure 2, the advanced solutions were installed and implemented at the plant site in phases, starting from tuning of regulatory controllers, implementation of DCS based strategies, model-based severity control, multivariable control and finally the online optimizer that formed the top layer of the automation hierarchy.

The execution of the project in phases had obvious advantages. Each application in the automation hierarchy was installed and tested completely independent of the layer above it. This helped the project team to gradually introduce new concepts to the plant personnel and train them on the use of these applications. By keeping its project team members closely involved with each phase of the design and configuration of these advanced applications, Polimeri Europa was able to reduce project costs while ensuring they received customized products that they could quickly put online and maximize their return on investments. Key features of the various applications implemented at Polimeri Europa as part of the automation upgrade are discussed in subsequent sections of this paper.

Figure 2 : Project Execution Milestones

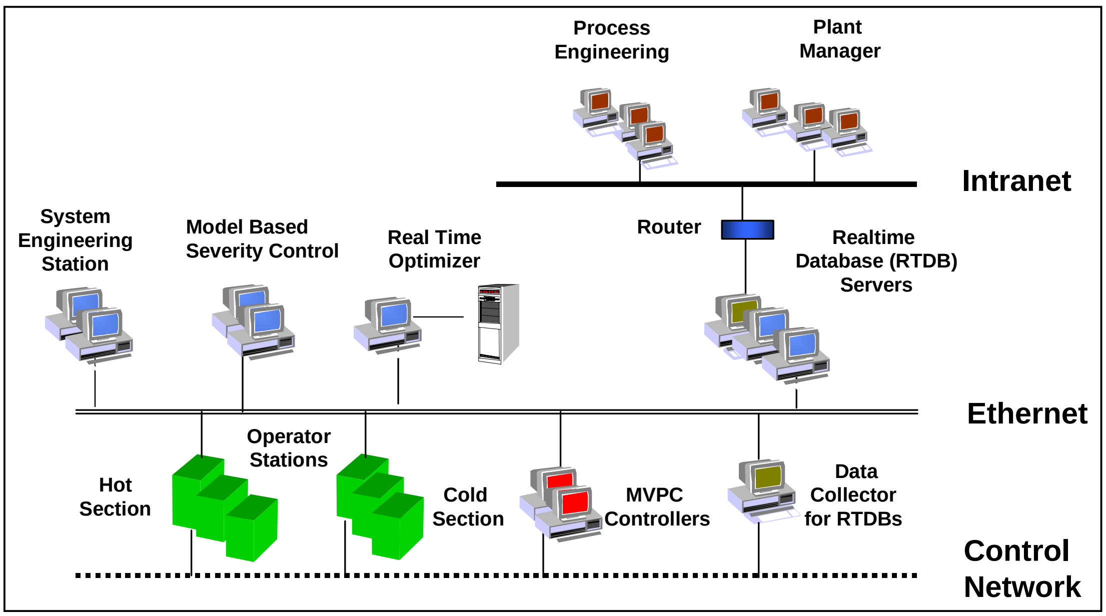


System Architecture

Due to the large scope of the project (25 MVPCs, 15 model-based severity controllers and the plant-wide RTO), the application hardware platform was carefully designed to satisfy both high data throughput requirements and short refresh time specifications. With the exception of the optimizer, all applications were installed on standalone work stations that operate under the Windows NT platform. The optimizer executes on a COMPAQ XP1000 computer (it is also, in fact, supported to run on a Windows NT platform). The optimizer user interface and data management system, from which the optimization cycle is sequenced and controlled, resides on a Windows NT workstation

Standard data communication interfaces were developed for these application using either OPC or vendor specific API protocol. Data for all the applications are distributed across three real time databases (Honeywell Uniformance ® PHD) or retrieved directly from the DCS (Honeywell TDC3000 ®). User interface displays running on the application computers are directly accessible in the control room via the operator stations. (This last feature was an essential requirement in obtaining the operator's acceptance for using these applications on a continuous basis as the primary tools for operating the plant. It also enabled operator training in-situ without exposing the application hardware to the control room environment.). The system architecture showing the relevant application nodes are shown in Figure 3.

Figure 3: System Architecture



Application Design Details

Individual MVPCs were installed and commissioned across the following process units at the Polimeri Europa ethylene plant:

- Pyrolysis cracking furnaces (15)
- Primary fractionator
- Quench water tower
- Condense stripper
- Demethanizer
- Deethanizer
- Acetylene hydrogenation reactor
- Ethylene fractionator
- MAPD converter
- Depropanizer
- Debutanizer

In addition the project scope included a plant-wide MVPC based Production Controller that was designed to coordinate the distribution of feed targets to each of the furnace MVPCs and to also relieve constraints across the ethylene plant process units. The severity control applications based on the online, closed form of the PYPS furnace model were implemented to run either in standalone mode or cascaded to the optimizer. The economic optimizer was the final application to be commissioned, as it required all the underlying control applications to be running closed loop in the online environment.

Given the extensive project scope and the interactions between the different applications, a key challenge facing the project team was the design of a user-friendly interface for each of these applications. These interfaces had to satisfy the requirements of both the project engineers as well as the operational staff. Since the operators were being exposed to several different new technologies at the same, it was critical that the user interface did not overwhelm them with information. Also proper handshaking between the various applications was crucial to keep them running smoothly without compromising the safety of the plant operations. With that in mind, the project team designed the interfaces and supporting programs with the following features:

- All data critical to quickly evaluate the applications performance or to diagnose problems are prominently displayed in the user interface.
- Inputs having multiple data sources (upstream controller, economic optimizer, or operator) are color coded for easy operator identification. This allows the operators to quickly identify their interactions with the various applications and is instrumental in ensuring the safe use of the overall system.
- Watchdog programs with built-in alarms verify proper handshaking between applications.
- Master switch logic allows the operators to quickly switch between the process control automation layers (basic regulatory control ➔ DCS based advanced control ➔ MVPC). The master switch logic automatically switches the modes of relevant DCS based control blocks to be consistent with the selected master switch position (for instance, Manual ➔ Auto ➔ Cascade) thereby relieving the operator from this time consuming task.
- Special programs perform tracking functions to ensure smooth and bumpless transfer of targets and setpoints across the different applications.

Furnace Section Controls

In keeping with the multi-layered approach selected for this project, the furnace control strategies are implemented at both the DCS and MVPC levels. Since the MVPCs are configured to run only once every minute, strategies that require faster control action are designed to operate at the DCS level. Specifically these controllers maintain:

- average COT of the feed coils by simultaneous manipulation of wall burner fuel gas pressures.
- steam-to-oil ratio by manipulating the flow of dilution steam to each coil.
- excess oxygen by manipulating the draft pressure.

Feedforward compensation is provided for the effects of feed or dilution steam flow changes and changes in fuel gas flow or quality on the average COT

The MVPC maintains total furnace throughput at the specified target while ensuring that individual coils remain balanced in both temperature and flow with respect to the corresponding average values for the furnace. Since safe and stable operation of the furnaces overrides all other objectives, the control strategy is designed to account for multiple process disturbances and honor all critical furnace constraints including valve saturation limits. A simplified control matrix that shows the interactions and gain relationships between the MVs and CVs including constraints, is shown in Figure 4. (The actual matrix size is 44 CVs and 15 MVs.)

In this design, the total throughput target is set by the operator or by the online economic optimizer when it is operating in closed loop mode. Fuel gas pressure biases are manipulated to simultaneously balance the individual coil hydrocarbon flows and COTs. Furnace loading and valve saturation constraints are relieved through manipulation of the average COT set point or individual coil flow controller set points. On these furnaces, floor burner pressures are typically base loaded to supply around 30 % of the total heat duty but can be manipulated (within a small range) by the controller to alleviate wall burner constraints.

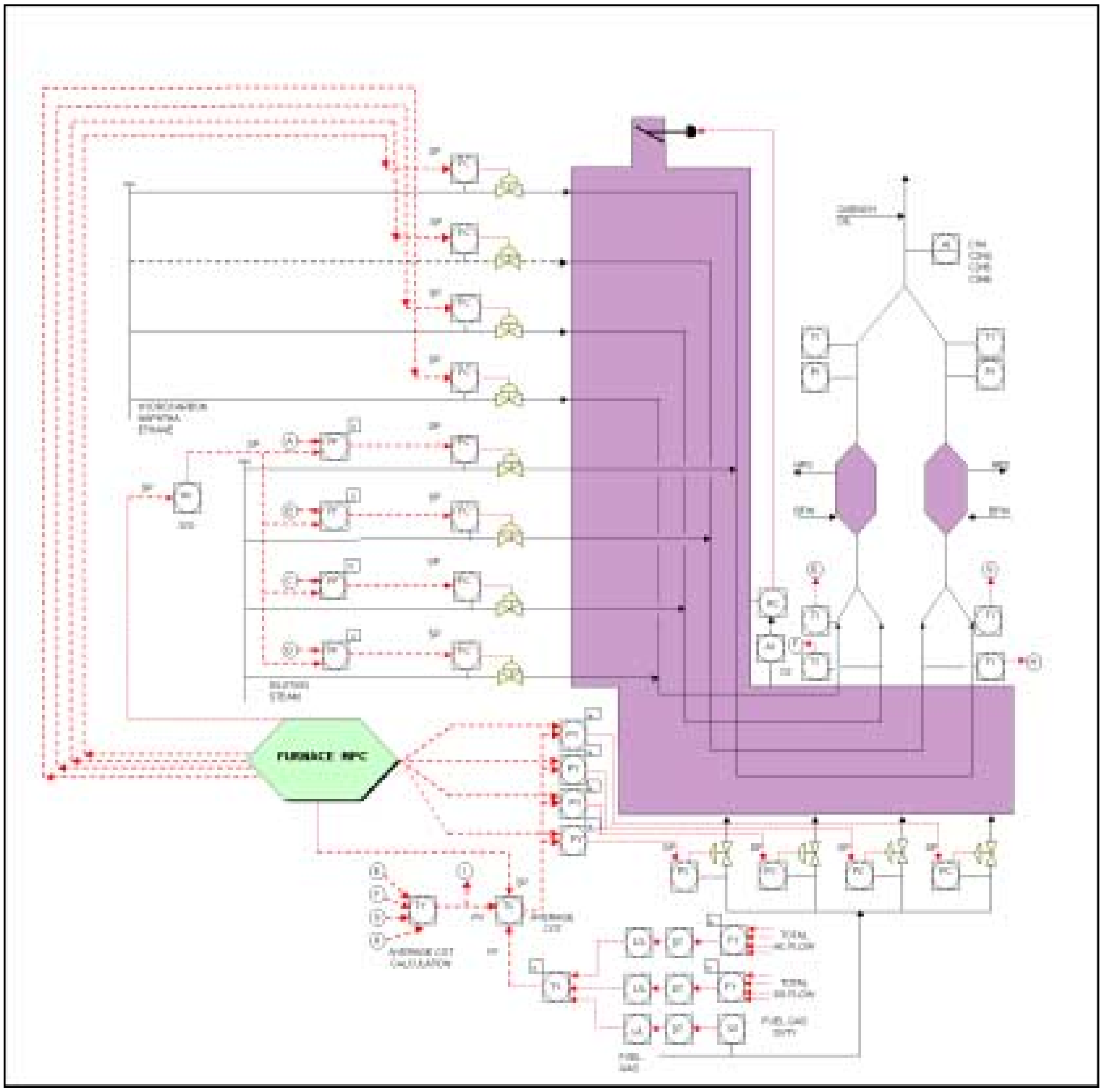
Figure 4 : Furnace Control Matrix

Controlled Variables (CVs)	Manipulated Variables (MVs)						
	Average COT	Coil 1 ~ N FEED FLOW		Coil 1 ~ N S/O		Coil 1 ~ N Fuel Gas Bias	
Total Throughput		X	X				
Average COT	X						
S/O				X	X		
COT Balance		X	X			X	X
Flow Balance		X	X				
XOT	X	X	X				
TMT	X	X	X				
CIP		X	X				
TLE OT	X	X	X				
Fuel Gas Pressure	X	X	X			X	X
Damper Position	X	X	X				
Excess O2	X	X	X				
CO in Flue Gas	X	X	X				
Coil Feed Flow Valve		X	X				
Coil Steam Valve		X	X	X	X		
Coil Fuel Gas Valve	X	X	X			X	X

Backup strategies with a subset of the MVPC functions are also implemented within the DCS layer to ensure proper redundancy. Master switch logic designed at the DCS level enable switching between the two strategies in a bumpless manner. Figure 5 shows how the MVPC strategies are layered on top of the active DCS based strategies.

The furnace MVPC strategies at Polimeri Europa have been online now for more than 18 months. During the course of this period, the controllers have been fine-tuned to achieve high online service factors. During major furnace disturbances, the controllers are designed to shed automatically to the control at the DCS layer based on user-specified maximum deviation limits of critical process variables. Moreover, the furnace operation is typically switched to the DCS layer when quick changes to the plant throughput are required to counteract emergency situations or during some scheduled events such as switching of compressors. Discounting these downtimes, the controllers are observed to have greater than 95% service factors.

Figure 5 : MVPC based Furnace Controls



Due to their inherent ability to decouple interacting variables, the MVPC strategies enable the furnaces to operate at multiple constraints simultaneously. The improved balancing of the individual coil hydrocarbon flows while maintaining constant total throughput is clearly highlighted in Figure 6. During the introduction of a co-crack feed stream (C5s) into the furnace, the total throughput strategy quickly reacts to this step disturbance by reducing the individual coil naphtha flows even as they smoothly converge towards each other.

The freedom to simultaneously manipulate the individual coil flows as well as the fuel gas pressure biases, allows the controller to balance the individual COTs (Figure 6) even in the presence of inherent interactions between burners. Moreover, as shown in Figure 8, when fuel firing is constrained due to valve saturation, the controller can manipulate the fuel gas pressure setpoint biases or (and) reduce the coil feed flows to bring the saturated valves back in control range. The same strategy also applies when any of the individual coil fuel gas wall burner pressures approach their hard constraint value (Figure 9). Without these strategies in place, the total furnace throughput has to be reduced to allow the average COT controller in the DCS the flexibility to increase the firing to maintain the COT. Evaluation of the long-term performance of the controller indicates a dramatic but consistent reduction in the average deviation of individual coil outlet temperatures from the target COT setpoint (Figure 10).

Figure 6 : Throughput Control and Coil Flow Balancing

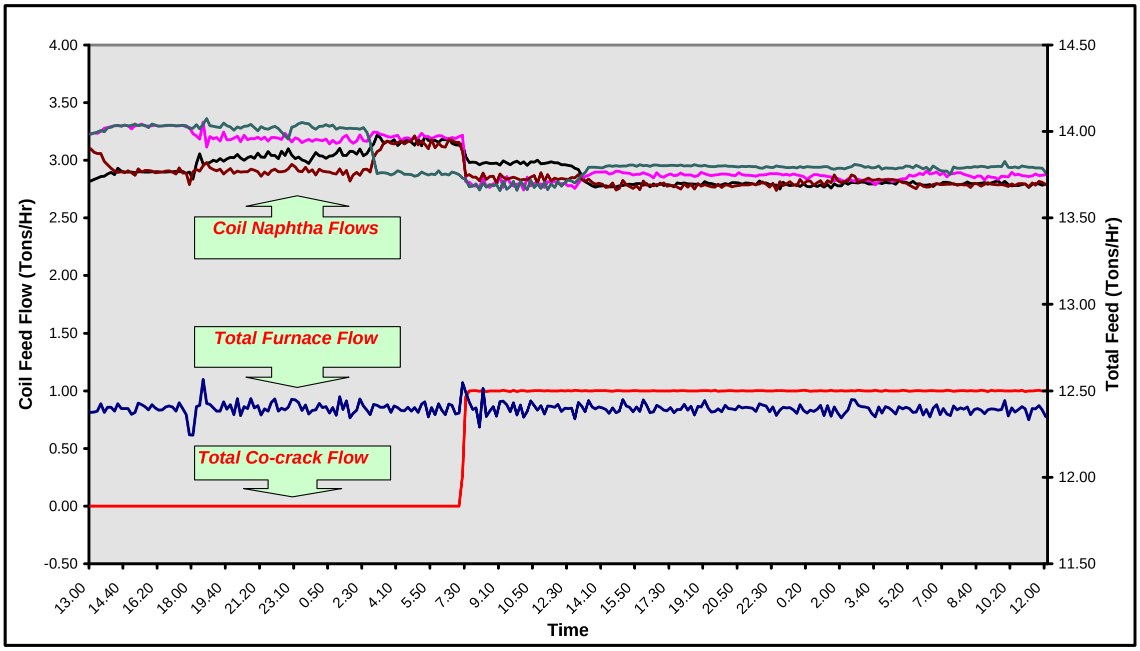


Figure 7 : Coil Outlet Temperature Balancing

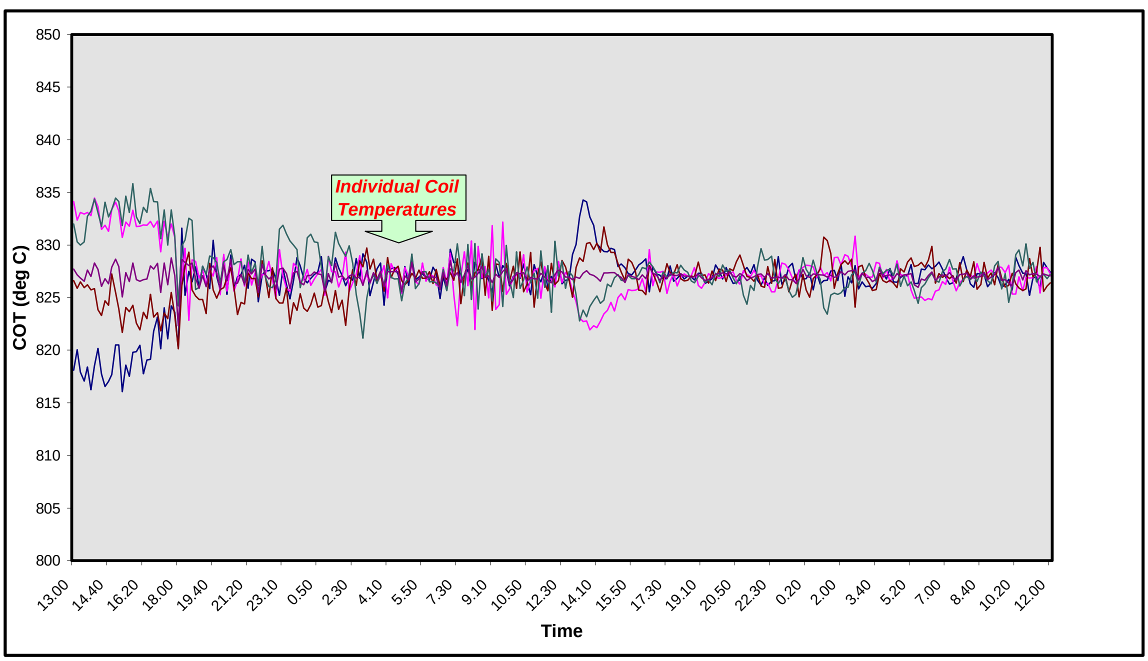


Figure 8 : Fuel Gas Valve Constraint Alleviation

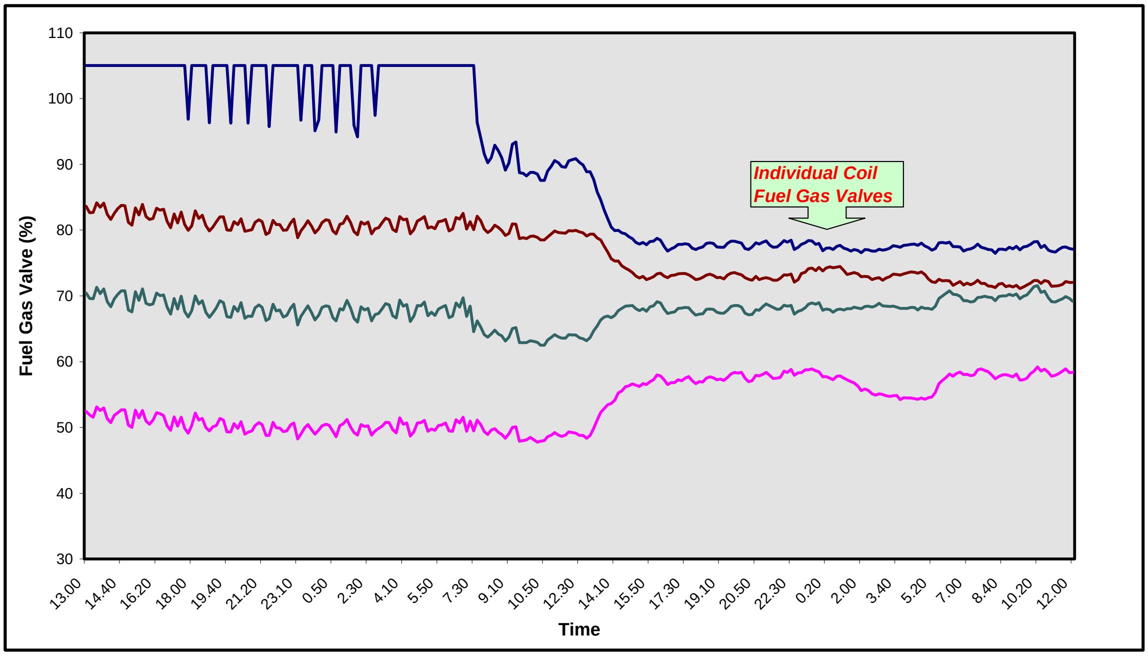


Figure 9 : Fuel Gas Pressure Constraint Alleviation

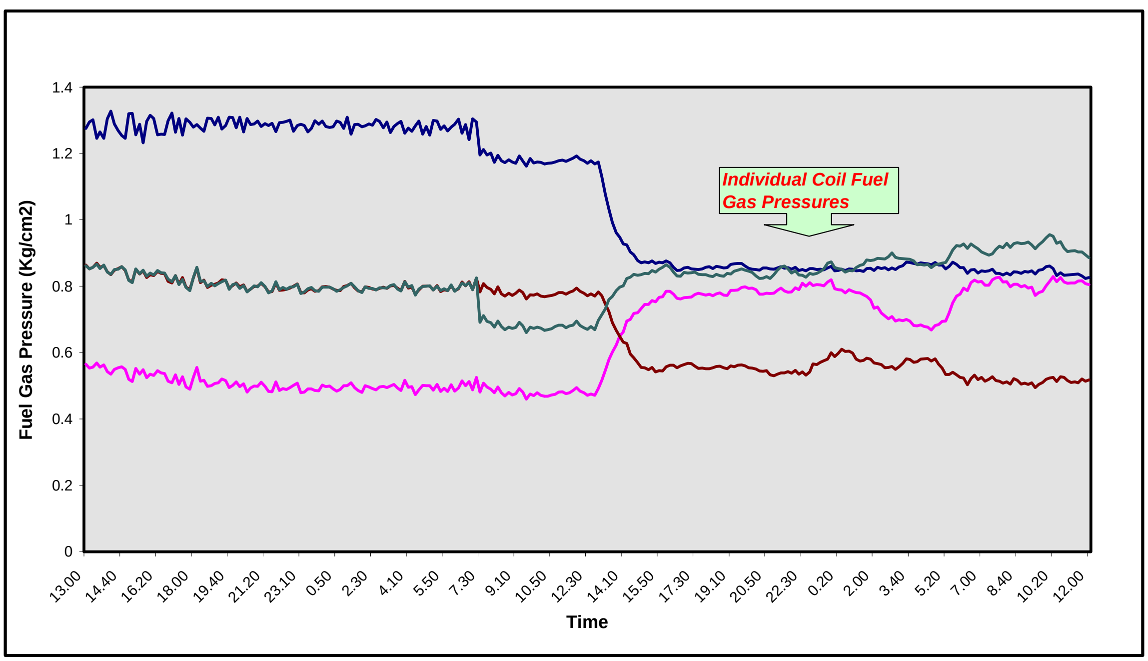
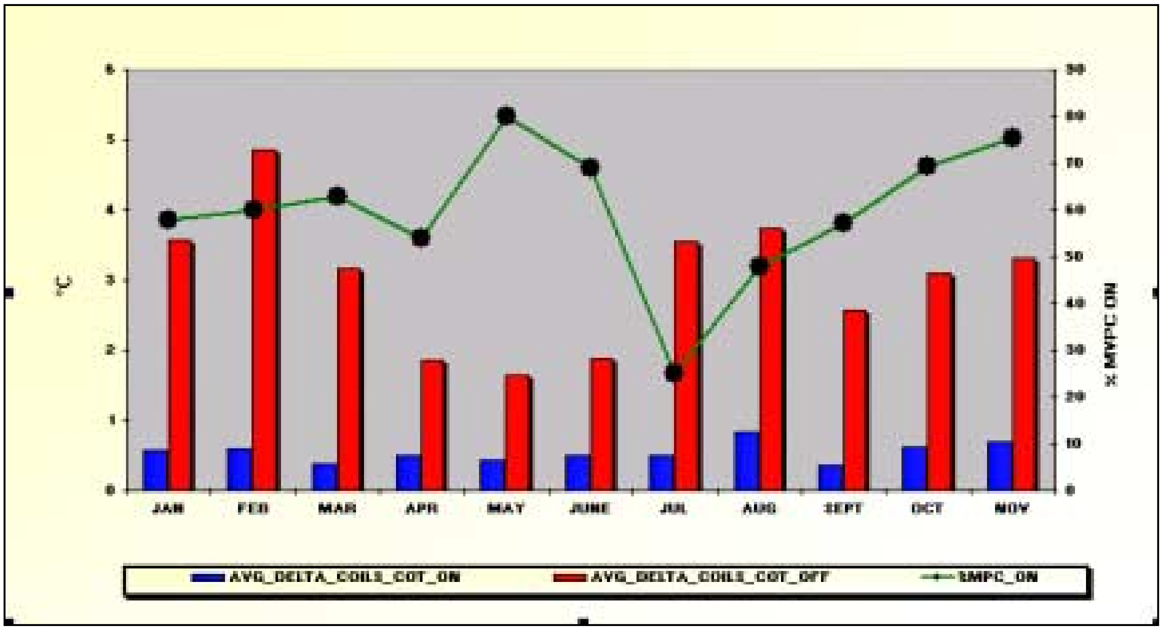


Figure 10 : Long-Term Performance Trends for COT Balancing

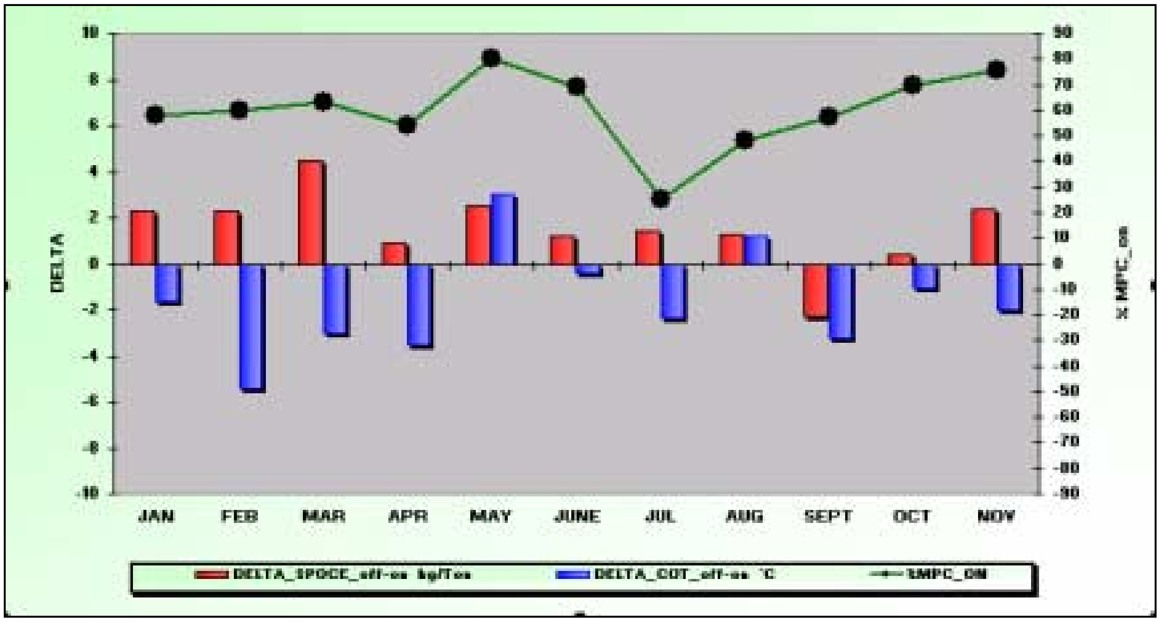


Note that the online service factors (%MVPC ON) shown in this plot include the time when the MVPC application is turned off during major plant upsets, emergency situations, furnace startup/shutdown operations, or during scheduled events such as compressor switches. During the month of July, there was significant downtime of the advanced applications due to plant troubles.

The APC has significantly improved the specific ratio (fuel energy consumption per unit processed feed) for the furnace operations as seen in Figure 11. The delta_spoce variable in this plot represents the change in fuel energy consumption per unit processed feed when the MVPC is in closed loop operation when compared to the period when it is not. Similarly, the delta_COT variable in the plot represents the difference in the average COT achieved across the furnaces when the MVPC is in closed loop operation when compared to the period when it is not. This shows that during the year 2002, when the MVPC was in closed loop operation on average, the furnaces were operated at higher COTs and yet consumed less fuel per unit quantity of feed processed. The only exception to this was during the month of September when the reverse was actually true.

A key intangible benefit obtained from the MVPC applications has been the significant reduction in operator intervention with the control system thereby enabling plant personnel to devote more time to instrumentation maintenance and other value added activities. It has also led to more consistent operation of the furnaces from shift to shift.

Figure 11 : Long-Term Performance Trends for Fuel Gas Savings



Recovery Section Controls

On cold section recovery units and distillation towers, the main objectives of the advanced automation strategies are to maintain product quality, minimize energy consumption (subject to constraint limits), and improve stability and performance of the operation during disturbances. Again, in keeping with the multi-layered approach adopted for this project, the recovery section control strategies are implemented at both the DCS and MVPC levels.

DCS based strategies on both the reactors consist of analyzer feedback controllers cascaded to ratio controllers that manipulate hydrogen flows to the reactor beds. These primarily act as fallback modes to the MVPC strategies and provide the required redundancy by satisfying a subset of functions that are critical for the operation of the unit.

On distillation towers, DCS based classical feedback reflux ratio control, reflux to feed ratio control, and tray temperature to reboiler flow cascades are implemented. Feed forward strategies that compensate for disturbances to feed flows, reboil medium temperatures are also provided.

Control objectives that required fast disturbance rejection capabilities are designed to operate at the DCS layer as it allows very fast execution cycles. Most other strategies are implemented from the MVPC layer from

Some key features of the MVPC strategies implemented on the hydrogenation reactors and the ethylene fractionator are discussed in subsequent sections of this paper.

1. Hydrogenation Reactor Controls

The primary control objective for the MVPC strategies on both the acetylene and MAPD hydrogenation reactors is to maintain the composition of the impurities (acetylene and MAPD respectively) in the reactor effluent streams close to but lower than the hard product specifications. The temperature rise across the reactor bed is a critical constraint to the control of the exit impurity concentration. This is because it is an indication of increased activity and total conversion but reduced selectivity towards the desired reaction. In order to minimize loss of desirable product it is critical that the reactor beds are not over-hydrogenated while maintaining impurity concentration in the reactor effluent at near-zero levels.

On the acetylene converter, the MVPC controller is tuned so that it reacts aggressively when the unreacted acetylene composition exceeds the hard product limit but reacts slowly when the acetylene composition increases from zero or near-zero levels. In the near-zero region, the process response is fairly nonlinear and if the hydrogen to acetylene molar ratio is decreased too rapidly it can cause a sudden breakthrough of impurities in the reactor effluent resulting in an off-spec product.

On the MAPD converter, the primary objectives are to prevent over-hydrogenation and to maintain the outlet MAPD concentration at a small non-zero value below the hard product specification. Given that the MAPD outlet concentration can change from zero to more than ten times the maximum allowed limit (typically < 2 ppm) in a very short period of time, it is inherently difficult to maintain the near-zero composition. For this reason, the primary controlled variable is treated more as a constraint rather than a true CV. Instead, a secondary "shadow" variable corresponding to the propane increase (delta propane) across the reactor bed is used as a CV to indicate the extent of over-hydrogenation. The delta propane directly corresponds to the conversion of propylene to propane.

The MVPC strategy is set up to minimize the delta propane towards a low limit (typically 0.4 mole%) representing the desired level of excess hydrogen. If the delta propane decreases below the low target limit or the MAPD concentration increases above the high limit, it is an indication that insufficient excess hydrogen is being maintained with respect to the reactor feed. In both situations, the controller increases the hydrogen to feed ratio so that the required minimum level of excess hydrogen is maintained. Since the controller execution frequency (4 cycles/minute) is substantially faster than the frequency of the online delta propane analyzer (update every 45 minutes), therefore between measurement updates the controller relies completely on its internal predictions to establish the future moves of the manipulated variables.

Figure 12 highlights the difficulties of controlling a process against a nonlinear and near-zero impurity constraint. The first half of the reactor operation shown in the figure represents the time when the reactor was under MVPC control and the later half when it was under direct operator control. The reactor performance during each of these periods is compared below.

At the start of the period under MVPC control, the reactor bed is seen to be over-hydrogenated as is indicated by the zero MAPD outlet composition and the high value of delta propane (0.8 mole% compared to the desired target of 0.4 mole%). As expected, the MVPC decreases the hydrogen-to-feed ratio causing the delta propane across the bed to decrease (towards its low target) and the propylene content in the propylene product to increase. This behavior continues until a disturbance in the depropanizer causes a sudden increase in the amount of unsaturated butylenes entering the reactor bed. Consumption of hydrogen due to its reaction with the butylenes reduces the amount of hydrogen available for MAPD hydrogenation. This results in a sudden break-through of MAPD (around 5 ppm) out of the reactor. The controller responds by increasing the hydrogen ratio to reduce the MAPD back down to zero. The measured value of propylene continues to increase for a period of time, a reflection of the lags and inventory holdups in the various drums and separation columns downstream of the reactor (propylene analyzer is located on the C3 rerun tower product stream).

On the next occurrence of the MAPD breakthrough, the operator takes control of the reactor and makes a very large change to the hydrogen ratio (to limit the MAPD break-through to less than two ppm). However, because the reactor continues to be over-hydrogenated for the next five hours a noticeable amount of propylene product is lost. The operator eventually reduces the hydrogen ratio (back to the level that the MVPC had taken it to) in order to minimize the over-hydrogenation.

Based on analysis of this data the project team plans to add a feed forward strategy to compensate for changes in feed C4s content, even though this disturbance has not been observed to occur frequently. This would improve the reactor bed control with respect to disturbance rejection while maintaining the objective of maximizing propylene recovery.

Figure 12: Reactor Outlet MAPD Constraint Control

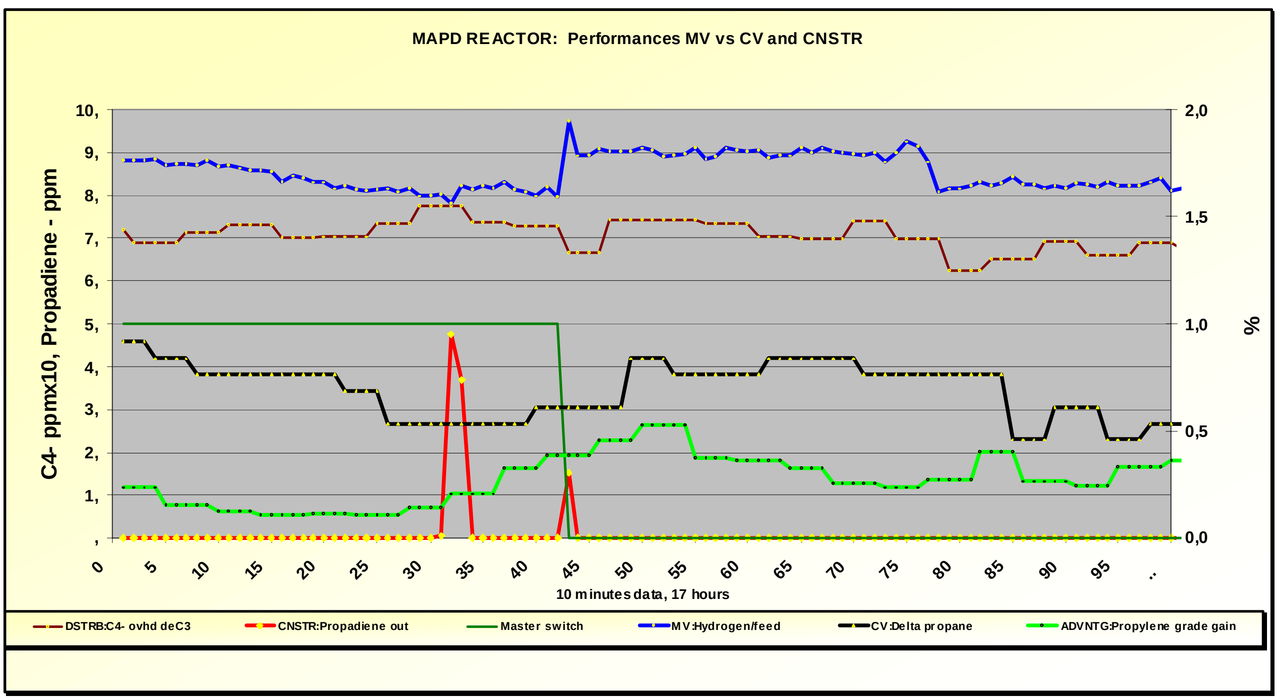
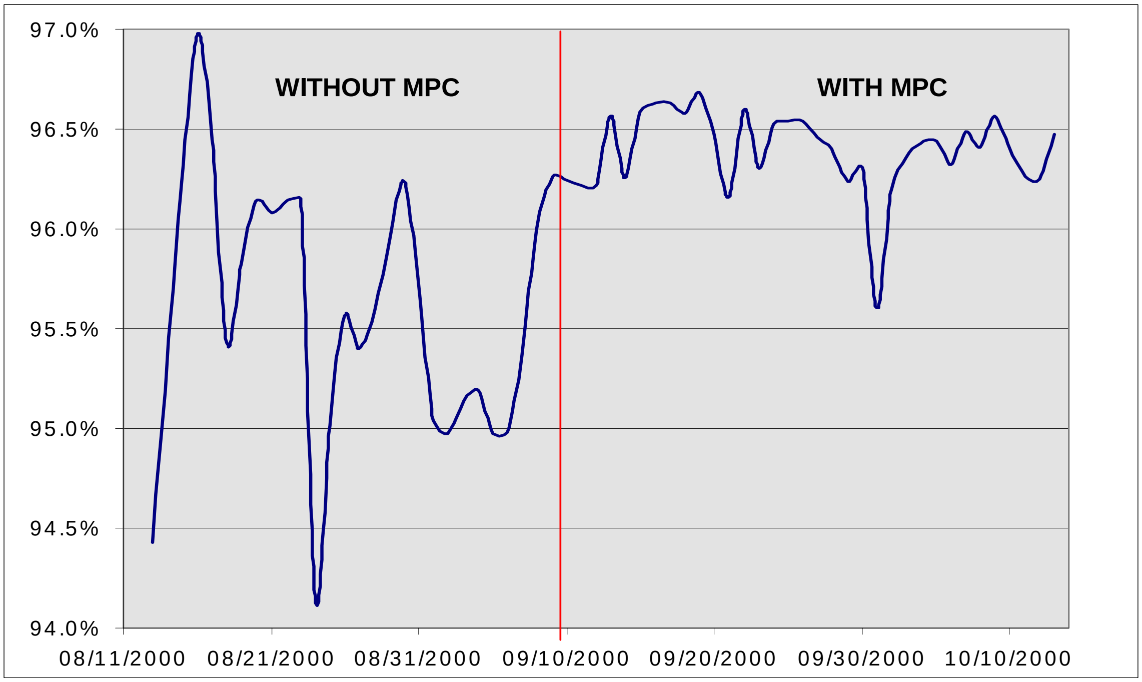


Figure 13 shows the long-term benefits of using the MVPC strategy with respect to maximizing propylene product recovery. Evaluation of data over a two-month period shows, that on average, the absolute increase in the propylene product was about 0.8 % with a corresponding price benefit.

Figure 13 : Percent Propylene in MAPD Reactor Outlet



2. Ethylene Fractionator

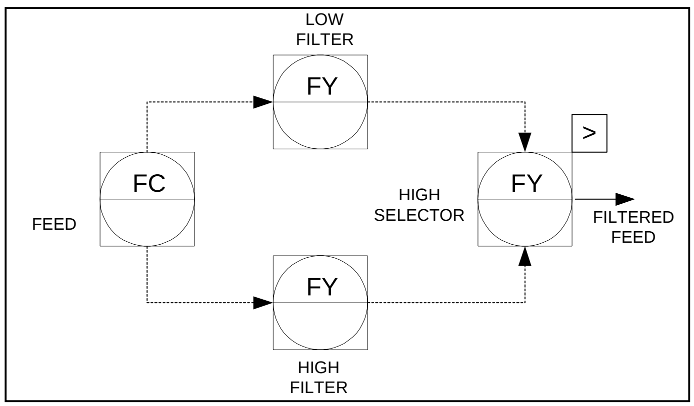
The ethylene fractionator at the Polimeri Europa plant was built with almost 30% over- design capacity with respect to the actual product rates. At current production rates, the column seldom approaches constraint conditions. However, this column offers significant challenges for control due to its tight integration with not only the refrigeration system but also the upstream process units. Major disturbances to the column include large feed flow changes during furnace switches and changes to the two side-reboil liquid draws that are used to condense the deethanizer overhead and charge gas to the chill box. These disturbances are observed to impact the control tray temperature, which in turn interacts with both the overhead and bottom composition.

The following DCS based strategies are layered below the MVPC layer:

- a reflux-to-feed ratio control that manipulates the column internal reflux in response to feed or set point changes
- a tray temperature control that cascades to a propylene vapor reboiler flow controller
- column pressure control that cascades to a propylene refrigerant condenser level controller
- a total side draw flow control to keep the total return vapor to the column constant
- feed forward control to compensate for the effect of feed changes on the control tray temperature
- feed forward control to compensate for the effect of total side reboil liquid draw on the control tray temperature

The reflux-to-feed ratio controller is designed to use a cross-limiting filter strategy (Figure 14) to compensate for transient composition swings during large changes to the column feed. This strategy ensures that during the transient state caused by the feed disturbance, the column is never under-refluxed (when the feed to the column decreases) or significantly over-refluxed (when the feed to the column increases). Since undershoots below the hard target are more acceptable than overshoots above it and the control objective is to push the product impurity level towards the hard high target, the feed signal to the reflux-to-feed ratio controller is filtered differently depending on whether the column feed increases or decreases. For instance, if the feed to the column decreases, the high-selector selects the high-filtered feed signal because it lags the low-filtered feed signal and therefore has a higher value during the transient state. Conversely, if the feed to the column increases, the high-selector selects the low filtered feed signal because it leads the high-filtered feed signal and therefore has a higher value during the transient state . Therefore during the transient period caused by a feed disturbance, the internal reflux to the column always lags the feed, more when the feed decreases and less when the feed increases. The net result is that the column is never under-refluxed during a feed disturbance.

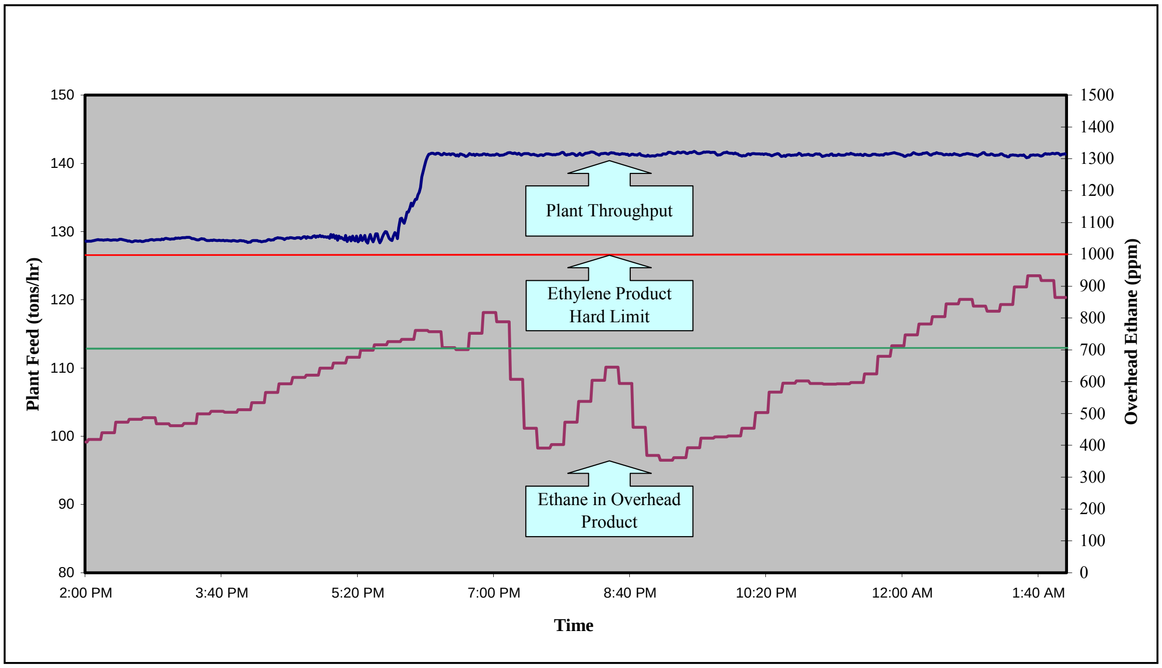
Figure 14 : Cross-Limiting Filter Strategy



For the multivariable controller layered on top of the DCS strategy, the primary control objective is to maintain the overhead draw ethane concentration at the product specification limits. The secondary objective is to drive the ethylene loss in the bottom stream to its set target. This target is either set by the operator or obtained from the online economic optimizer when it is in service. Since the column is over-sized for the current operation, a reflux to distillate ratio minimum constraint is imposed to ensure that there is sufficient liquid loading to facilitate proper separation. The set points of the reflux to feed ratio controller and the tray temperature controller are the primary manipulated variables. The controller is configured with cross model gains between the tray temperature and the bottom and overhead compositions so that these interactions can be properly decoupled. The column pressure variable is prioritized such that it would be manipulated for the overhead ethane composition only if the reflux to feed ratio control set point is at its hard limit.

Figure 15 shows the performance of the ethylene fractionator MVPC strategy during a major feed change (10%) due to a furnace being switched online. During this period the operator had set the target range for the overhead product impurity (ethane) to be between 700 ppm to 850 ppm. The hard limit for the impurities in the product is 1000 ppm. The MVPC controller initially reduces the reflux-to-feed ratio in order to drive the ethane specification towards the desired range. During the transient state caused by the increase in feed to the column, the DCS based reflux-to-feed ratio strategy causes the column internal reflux to increase quickly (with only a small lag relative to the feed). This causes the ethane content in the product to decrease below the desired range during the transition. When the column feed stabilizes, the ethane content is once again gradually pushed back into the desired range by the MVPC controller. The net result is that during the major feed disturbance, the ethane content in the product does not violate the hard limit.

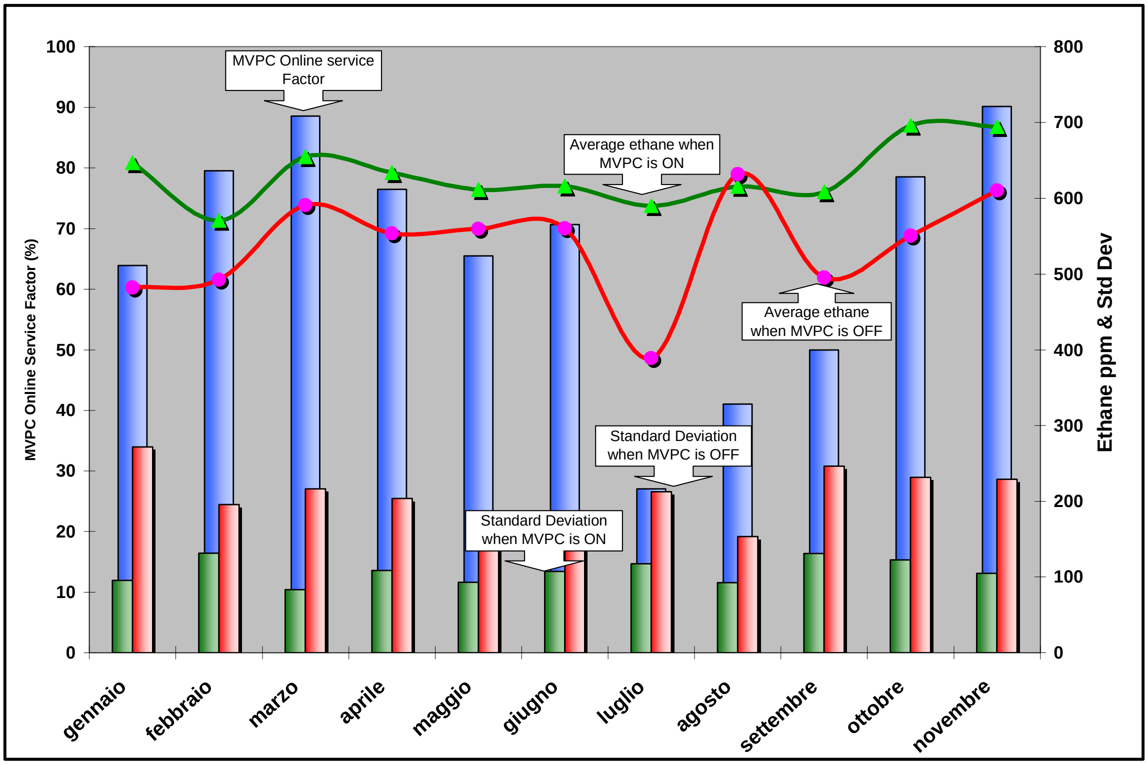
Figure 15 : Effect of Feed Disturbance on Overhead Composition Control



As shown in Figure 16, the addition of the advanced control layers results in significant overall improvement in the ethylene fractionator performance. Specifically, with the MVPC in service the standard deviation of the ethane in the ethylene product is reduced by more than 50%. On average, this allows the column to be operated almost 100 ppm closer to the hard product specification.

The reduction in over-purification of the ethylene product translates into energy savings in the propylene refrigeration system due to reduction in reflux and therefore condensing requirements. The economic benefits of this advanced control strategy have been evaluated by process simulation.

Figure 16 : Ethylene Fractionator APC Performance



Production Controller

The production control MVPC implemented at Polimeri Europa is a single, large controller consisting of ninety-one controlled variables (including constraints) and forty-one manipulated variables. It includes both hot section and cold section variables. It monitors and interacts with individual process unit MVPC's whose objective is to locally maintain their operation as close as possible to the desired targets. The primary objectives of the production control strategy are as follows:

- Control the plant throughput or production rate at the desired optimal value.
- Maintain optimal distribution of feedstock to each furnace.
- Relieve individual furnace constraints without reducing overall throughput or production rate
- Set suction pressure targets for the cracked gas compressor.
- Relieve plant-wide constraints with minimal throughput or production loss.
- Manipulate suction pressures of refrigeration compressors and/or chilling train operating temperatures to alleviate heat exchanger and recovery unit constraints.

The production controller is operated either in standalone mode or is cascaded to the plant-wide economic optimizer. In standalone mode, the operator sets the targets to the production controller. When cascaded to the optimizer, custom download programs transfer optimal set points calculated by the optimizer to the high or low CV targets of the production controller. Special tracking logic ensures bumpless transfer of targets when the production controller is switched from standalone mode to cascade. In general, variables whose targets are set by the optimizer are configured in the production controller as "pass-through" variables. In other words, they are configured both as CVs and MVs with a gain of one between them. In the absence of constraints, the CV targets are passed unaltered to the DCS through manipulation of the corresponding MVs. If related constraints become active, priorities and tuning factors on the CVs and MVs determine the sequence and extent to which different degrees of freedom are used to alleviate the constraints.

Due to the size of the control matrix and in order to simplify the commissioning activities, the controller was implemented in two phases. The first phase involved activating only the hot section part of the control matrix that included the furnace variables and the cracked gas compressor. This gave the project team the opportunity to fully focus on tuning the production controller with respect to just the furnaces and the compressor. As shown in Figure 17, the hot section control matrix does not explicitly contain the furnace constraints. The actual control of the furnace operation and alleviation of local constraints is completely managed by the furnace unit multivariable controllers. These controllers execute at higher frequencies (once per minute) than the large production controller (once every three minutes) and can therefore respond much faster to constraint violations and quickly move the furnace operation to a safe region until the constraints are relieved. The function of the production controller is limited to maintaining the desired (optimal or user entered) distribution of feedstock to the furnaces and redistributing the targets when local constraints become active in any of the furnaces. Furnaces whose unit MVPCs are monitored and managed in this manner by the production controller are categorized as interactive furnaces. Furnaces whose unit MVPCs are operated standalone and furnaces that are controlled from the DCS are categorized as non-interactive furnaces. A user-specified minimum number of furnaces have to be in interactive mode for the production controller to be running closed loop.

<u>Figure 17 : Production Control Matrix</u>

	Hot Section MVs				Cold Section MVs				Emulated MVs	
Controlled Variables (CVs)	Furnace 1 ~ N throughputs		CGC pressure	CGC balancing	Refrig pressure	Refrig balancing	Chilling Train Pressures	Chilling Train Temperatures	Backend Columns	
Total plant throughput target	X	X								
Furnace 1 throughput	X	X								
Furnace 1 offset	X	X								
Furnace 1 imbalance	X	X								
Furnace 2 throughput										
Furnace 2 offset										
Furnace 2 imbalance	X	X								
Furnace N throughput										
Furnace N offset										
Furnace N imbalance	X	X								
CGC suction pressure			X							
CGC load balancing ratio				X						
CGC constraints	X	X	X	X						
Refrig suction pressure			X		X					
Refrig load balancing ratio				X		X				
Refrig constraints	X	X	X	X	X	X	X	X		
Recovery section constraints	X	X			X		X	X		
Backend emulated CVs	X	X			X		X	X	X	X
& constraints	X	X							X	X
	X	X							X	X
	X	X							X	X

Each furnace has three controlled variables in the production control matrix: throughput, offset and imbalance. When in operation, the optimizer sets the individual CV throughput targets of interactive furnaces as well as the total plant throughput CV target. If the optimizer is not in operation, the operator is only required to specify a change in the total plant throughput target from its current value. Based on this required change, targets of unconstrained, interactive furnaces in the production controller are automatically calculated in proportion to their remaining capacities. If the furnaces are unconstrained, these targets are sent down unaltered to the CV throughput targets of the individual furnace MVPCs. Furnaces that reach local constraints, will have non-zero offsets corresponding to the difference between the desired targets and the actual flows achieved. In order to eliminate these offsets, the production controller decreases the targets of these constrained furnaces The throughput flows of other unconstrained interactive furnaces are simultaneously increased above their targets so that the total plant throughput is maintained. The imbalance CV on each furnace forces this redistribution of flow targets to occur proportionally across the unconstrained interactive furnaces.

Another novel feature implemented in this controller is its ability to continue in closed loop operation even during startup or shutdown of non-interactive furnaces. During this situation, the total plant throughput target is ramped up or down at the same rate as that of the non-interactive which is going through the transient operation. Therefore, no redistribution of interactive furnace targets occurs during this period. This was a requirement specified to the project team during the design phase.

Under name-plate throughput conditions, Polimeri Europa operates two cracked compressor trains in parallel each with its own set of suction drums. The compressors are driven by steam turbines. When both compressors are in operation, one of the steam turbines is operated with a fixed governor position whereas the governor position on the other turbine is manipulated to control the suction drum master pressure controller. Under these situations the compressors do not limit the plant throughput capacity and minimum flow recycles have to be activated to prevent them from going into surge. A DCS based load balancing controller that adds a bias to the governor position of one of the two compressors has been designed to distribute the loading between the two trains. When the production controller is in operation, the operator (or the plant-wide economic optimizer) sets the desired target for the charge gas compressor first stage suction drum pressure as well as the desired load distribution between the two compressor trains. If the compressor system is unconstrained, these targets are directly passed through as setpoints to the DCS control blocks. When compressor constraints become active, the production controller first manipulates the suction pressure before making changes to the plant throughput.

During the second phase of implementation, the cold section part of the control matrix that included the chilling train, propylene and ethylene refrigeration compressors and key backend columns was activated. Depending on the plant throughput capacity, two propylene refrigeration compressor trains can be operated in parallel. DCS based suction drum pressure and load balancing strategies have been implemented for this compressor. When the production controller is in operation, the operator (or the plant-wide economic optimizer) sets the desired targets for the ethylene and propylene refrigerant compressor suction drum pressures as well as the desired load distribution between the two propylene compressor trains. Several of the production controller gains relating the refrigerant drum suction pressures to local constraints or constraints in refrigerant users (such as recovery section heat exchangers) are obtained through the offline simulation of the RTO model.

The production controller also monitors the constraints on key backend columns. The actual control of these columns is actually managed independently by the MVPCs implemented on these units. To follow their dynamic operation, the production controller shadows or emulates the moves of all the manipulated variables of these downstream unit controllers. This is achieved by imbedding all of the control, constraint, and manipulated variables of these downstream units into the production controller matrix. These variables are tuned and prioritized in such a way that the manipulated outputs from the production controller in effect track the moves made by the corresponding MVs in the downstream unit controllers. This allows the production controller to predict when these downstream units can get constrained (flooding, off-specification products etc.) and whether there is a loss of degrees of freedom in the downstream MVs. In this situation, the production controller has the option to manipulate conditions in the chilling train or the suction pressure of the refrigeration compressors.

Good tuning and proper prioritization of all the control objectives is of critical importance in this controller because of the various interactions between the different units, the lags between them, and the significant differences in the dynamics between the hot and cold sections. The project team continues to closely monitor this controller and fine tune it as necessary to achieve the desired performance

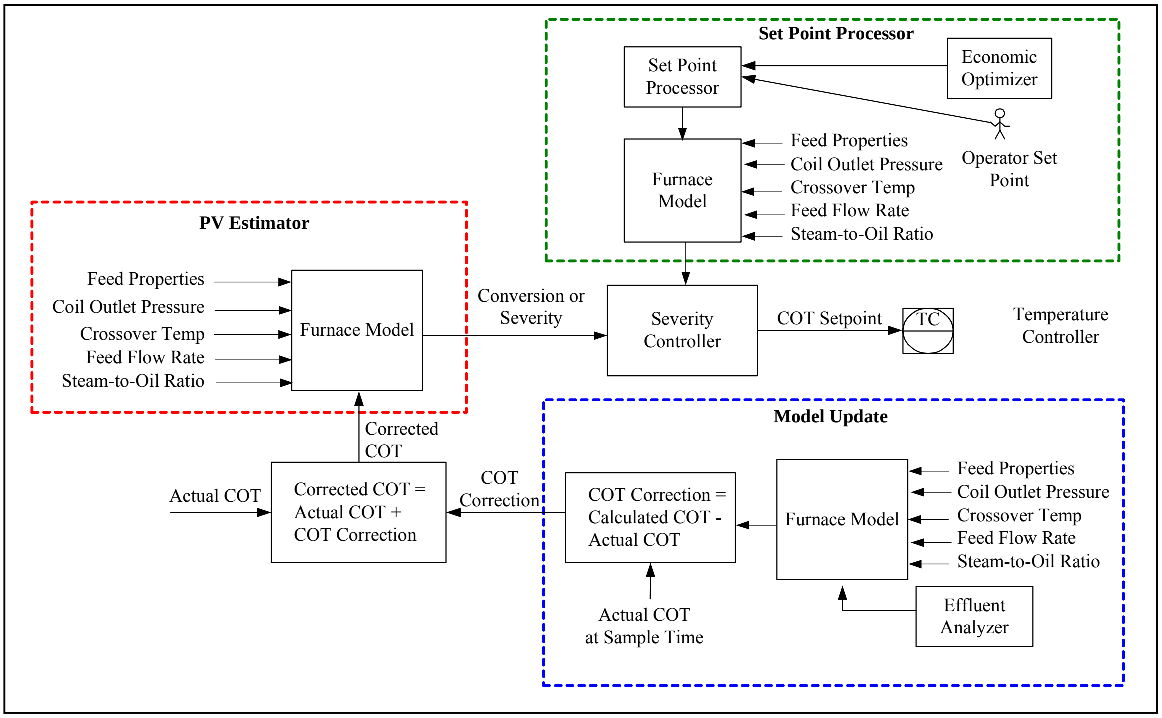
Model-Based Severity Control

The primary objective of the severity control applications installed at the Polimeri Europa plant is to maintain each individual furnace at a target severity, conversion, or ethylene yield set point based on current process conditions and feed properties. This strategy is achieved by the on-line implementation of the PYPS furnace model from within the severity control package. The furnace model calculates a coil outlet temperature (COT) required to maintain the target severity set point. The calculated COT is sent as a set point to the individual furnace MVPCs. The target severity can be entered either locally by the operator or directly downloaded by the economic optimizer. The online furnace model is updated whenever an effluent on-line analysis is available. The model therefore eliminates the measurement lags in slow analyzers and provides a continuous value of severity for the purpose of real time monitoring of this key variable. In addition, the severity control package provides important information to other upper level layers of the automation hierarchy as well as key furnace performance indicators that provide useful feedback to the plant personnel. Figure 18 provides an overview of how the furnace model is used from within the severity control package.

Key results calculated by the severity control package are summarized below:

- Estimates of coil outlet temperatures (COT's) required to achieve cracking severities or conversions
- On-line prediction of furnace effluent yield components
- Fouling of radiant coils and transfer line exchangers (TLE) that impact the cracking yields
- Predictions of furnace runlengths that can be used to validate cracking severity targets versus required decoking schedules.
- Calibration constants based on analyzer feedback that could be used by the online optimizer.
- Predictions of key unmeasured variables such as tube metal temperatures and radiant coil pressure drops that are used as constraints in the furnace MVPC strategies.
- Nonlinear process gain information related to manipulated variables around the furnaces that could be used by the furnace MVPCs.

Figure 18 : Severity Control Overview



The benefits of running closed loop severity control are most evident when the feedstock composition varies constantly. The naphtha feedstock quality is expressed in terms of a quality index that is a function of the naphtha properties. Higher the quality index, better is the naphtha quality.

As seen in the two plots in Figures 19 and 20, the quality index appears to cycle with a period of around one day. Figure 19 indicates that if the furnace COT was kept fixed over a period of four days during which the naphtha quality went through the observed cycles, the severity ratio C3-/C2- would fluctuate with a standard deviation of +/- 0.0136 around the mean. Figure 20 indicates that if the COT is continuously manipulated as the naphtha quality changes, the standard deviation of the severity ratio decreases almost 60%. This obviously translates to higher ethylene yields out of the furnaces and improved runlengths over the long-term. A related benefit is improved stability of the effluent compositions that facilitates tighter control of downstream units.

Figure 19 : Furnace Severity Variation without Severity Control

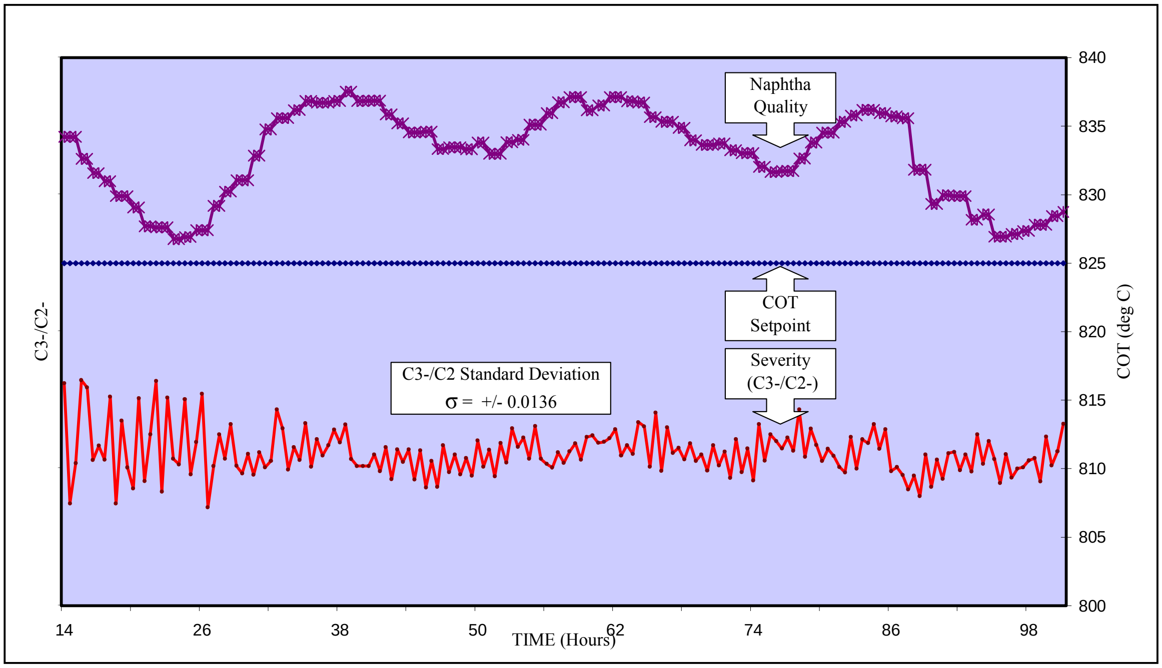
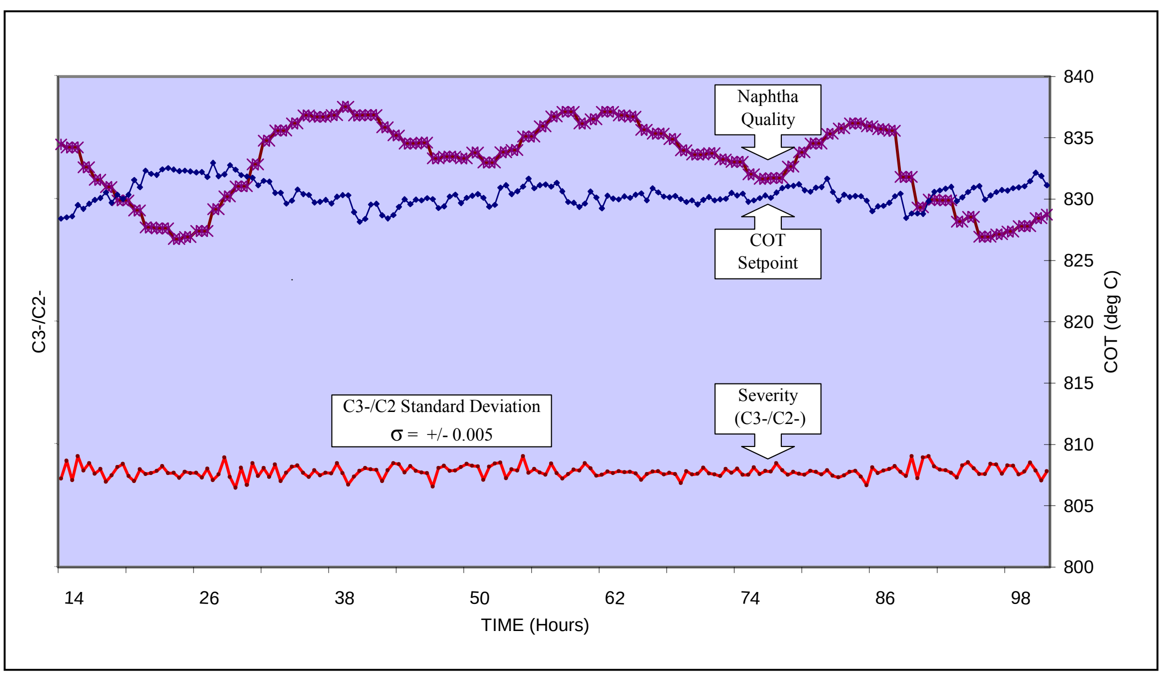


Figure 20 : Furnace Severity Variation with Closed-Loop Severity Control



Plant-Wide Optimizer

The primary goal of the plant-wide optimizer is to determine the best way to operate the Polimeri Europa ethylene plant given:

- Available feedstock
- Production targets
- Economics (feedstock and utility costs and tiered product pricing)
- Plant constraints
- An objective function – maximum profit or maximum ethylene or maximum propylene

Key design issues and some novel implementation features that were incorporated in the optimization system are discussed below:

Dynamic Flowsheeting Capability

The optimizer is designed to be available for service irrespective of which actual process equipment are in service. Some of the process line-up considerations for the Polimeri Europa ethylene plant that required dynamic flowsheeting capability in the optimization system include:

- Six (6) different furnace feedstock combinations to each individual furnace
- Line-up of reactor beds in acetylene and MAPD converter systems
- Various heat exchanger line-ups including quench water users, feed preheaters, refrigeration exchangers, and several steam and utility exchangers
- Various piping line-ups in utilities systems (quench, refrigeration, steam and fuel)
- Different line-ups for the compressor systems (charge gas, propylene refrigerant and ethylene refrigerant systems)
- Chilling train expander service status
- Operation without recycle furnace
- Downstream plants operational status – manifestation of such statuses covers recycle flow balancing affecting economics and the process optimum
- Varying control approaches – for example pressures floating (i.e. minimized) or on control with achievable set points

The combinatorial nature of the different process line-ups precludes configuring an explicitly different flowsheet covering every possibility. Instead these service statuses are reflected in the optimization model by setting a Scenario status. The scenario defines a set of model configuration changes (multiple connection/unit additions/deletions and specification changes) that are applied to a base model configuration. Active branches are defined within scenarios enabling a certain level of hierarchical configuration. As each scenario is independent of other scenarios there is a certain level of abstraction in the definition and configuration of many of the more than 50 defined scenarios for the project. For online optimization, the statuses of these scenarios are automatically set through calculated logic tags in the supervisory database; for offline studies the status of the scenarios could be defined by the user.

The ease of configuring the optimization model for the changing active plant configuration through the scenario concept demonstrates that even for very flexible plants, high availability factors are achievable for real time optimization utilizing process flowsheet models.

Selection of Model Rigor for Major Equipment

A real-time optimization application requires models that predict the true effect that changing the independent variables has on the objective function and all relevant plant operating constraints. Base model trends, or gradients, are of the greatest importance. In recent years this has led to the use of open equation chemical engineering models of equivalent fidelity to those found in process simulators used for process engineering design.

However, typically there are also requirements for short cycle run times and for high solution robustness. Generally, this requires compromising the model scope and complexity. Scope reflects not just the equipment that is modeled, but also the rigor of the model and selection of thermodynamic relationships and stream component slates. Large models increase both project implementation and cycle run times. Some high fidelity chemical engineering models, particularly those associated with thermodynamic equations of state, can introduce multiple roots and/or local optima into the problem presented to the mathematical solver—ultimately manifesting as a less robust system in the early stages of project commissioning. Additionally, the model needs to be flexible enough to support all reasonable equipment line-ups; sometimes this is most appropriately achieved by utilizing more abstract models for some plant sections.

In summary, model selection has to be a balanced decision based on the discipline of capturing the benefits associated with optimization and use of the system for offline case studies.

Reflecting this analysis, some aspects related to the model selection criteria used on this project are documented below:

- The pyrolysis furnace models – reflecting process licensor technology and including furnace runlength predictions – provide a reasonable trade-off between model ‘complexity’ and execution speed in the open environment.
- The Soave-Redlich-Kwong equation of state with binary interaction parameters is utilized for modeling the ‘stream-based’ sections of the model downstream of the furnaces. The component slate is varied for economy of model size; the maximum number of components in use for any one section of the model is about 35.
- Equilibrium stage column models are used for the major distillation columns in the recovery train. Column efficiencies were determined during design. Within the flowsheet, the columns downstream of the charge gas compressor system utilize a column model denoted as RTC (real-time column) that have special ‘fast’ robust thermodynamic calculations derived based on the method of ‘corresponding states’. This model is less nonlinear with respect to temperature and is suitable for columns that have reasonably consistent feed composition. (During configuration each of these column models were ‘calibrated’ using a parallel column model with more rigor.)
- Rigorous models for compressor stages are used. For each compressor stage, polytropic head and efficiency look-up tables are generated from the compressor performance curves. These tables are calibrated based on live plant measurements. Anti-surge recycle flows are calculated based on operating minimum flows.
- Simple UA models are used to predict heat exchanger performance and conserve energy balances. UA is generally made a function of hot and/or cold side flow. Prediction of level for some exchangers is implemented by decoupling the U and A terms based on a knowledge of exchanger geometry and level tappings.
- Valves are modeled primarily to reflect constraints. Flashes are used to model the hot or cold sides of heat exchangers as relevant; three-phase flashes are used for the charge gas compressor system. Many pressure drops are predicted as simple functions of volumetric flow.

- Simplified process models are included such that all streams are taken to their final use point (sales, recycle, fuel, etc). Balances related to sending streams to fuel or recycling to the furnaces for cracking are carefully defined to ensure that the predicted optimum is appropriate.

The model selection choices are made to ensure that the calculated optimum fully reflects the true economical situation and respects the plant constraints. The plant can be constrained either in the furnaces or recovery section area. The most important recovery section constraints that are predicted by the model relate to compressor constraints – and associated users for the refrigeration compressors.

Shadowing of MVPC Approach for Predicting Certain Constraints

The multivariable control system receives the identified optimum setpoints as targets to the CVs – with several of these CVs set up with pass-through to the relevant MVs. This ensures that the plant is operated in a safe and feasible manner and also provides incentive to mimic the underlying multivariable controllers within the process flowsheet model in certain instances.

Some plant constraints cannot be predicted practically using a 'fundamental' model. For such situations the process flowsheet model is developed from or made consistent with the approach implemented within the multivariable control for predicting these constraints. Some examples include:

- Charge gas compressor suction pressure control output (to woodward) is assumed to be a simple function of compressor suction pressure and mass flow.
- Fuel gas pressure constraints in the multivariable controller are reflected through fuel gas valve position constraints in the optimizer. The fuel gas valve position high limit reflects either maximum valve position or maximum fuel gas pressure assuming a simple relationship between fuel gas pressure and fuel gas valve position.
- The measured ratio of fuel gas hearth to total flow is the target ratio. If the fuel gas to wall reaches a valve constraint then this ratio can be increased within its high constraint.
- Whilst rigorous predictions using valve models are used to predict some furnace valve constraints (fuel gas, dilution steam), damper and quench fitting valve positions are predicted using simpler models consistent with those in the multivariable control.

In general the models referenced above in the process flowsheet are close to – but not exactly – linear. These custom models are implemented using high-order flexible scripting and integrated seamlessly with units defined using the available list of models in the modeling system.

Many of the constraint alleviation approaches incorporated into the model during the optimization step – such as the fuel gas hearth to total ratio mentioned above – are implemented through dynamic switching of the active equations to be solved using a feature denoted as 'complementarity'. Another such example is that during optimization the quench tower overhead and bottom temperatures are allowed to increase to keep the total quench water flow within constraints – consistent with the multivariable control implementation.

In addition to constraint prediction, the optimizer is designed to mimic the underlying control approach within the model in other relevant ways. For example the recycle furnace conversion and steam-to-oil ratio are determined as functions of recycle furnace flow (predicted by balance calculations when the process flowsheet is optimized).

Offline Steady State Gains Calculation For Control

In several instances, the optimizer models were simulated offline to generate process steady state gains used for configuring the multivariable controllers. This was especially true for process relationships that could not be conveniently obtained through plant step tests or discerned though analysis of historical data. Some examples include column pressure effect on the various column CVs, refrigeration drum pressures effect on relevant CVs (such product specifications and valves) and compressor speeds as functions of different loads (e.g. furnaces total throughput for the charge gas compressor).

Flexibility of Real-Time Optimizer

Some aspects related to the flexibility of the real-time optimizer were discussed above – including dynamic process flowsheet generation and model selection reflecting required objectives. Nevertheless, the open equation process flowsheet for the ethylene plant is a large nonlinear problem that is ultimately presented to the mathematical solver. In addition to technical decisions concerning modeling approach the following features are also implemented to improve the system robustness and availability:

- The full process flowsheet is apportioned into four sub-sections that are first solved separately. The first sub-section reflects the furnace area and the relevant results from this sub-section are used in solving the second sub-section, etc. Within each sub-section flowsheet certain units are initialized every time the optimizer cycle runs. This set-up achieves the objective of presenting a more solvable and better linearized full process flowsheet problem to the solver.
- A measurement model is used to bring all the process measurements into the flowsheet. This facilitates easy changes to which process measurements are emphasized during data reconciliation.
- L1-norm objective function is used for data reconciliation.
- Offline dynamic parameter estimation is incorporated (output from severity control and a separately executed rectified material balance).

In addition, a PC-based Optimizer Data Management Interface manages the optimizer cycle, providing the following functionality:

- Provides an optimizer executive which schedules the optimization cycle and its associated tasks
- Executes the tasks associated with the online executive
- Provides the optimization system user interface
- Provides a forms-based interface for configuring the details of the different tasks associated with the optimization cycle

A very flexible set of tools is provided for configuring the data collection and conditioning tasks and monitoring plant steady state. Thoughtful use of these tools has been an important factor in obtaining high availability and service factor for the real-time optimization system.

Polimeri Europa personnel are continuously monitoring the closed loop performance of the optimization system in order to quantify the long-term tangible benefits.

Summary

A large scale automation project has been implemented at the Polimeri Europa ethylene plant at Porto Marghera, Italy. The project objectives though very ambitious, were successfully met due to the strong commitment of the project team (including client and vendor personnel) to deliver a fully functional system that has already realized tangible economic and operational benefits. With continued operator training, and a strong push from the plant supervisors to keep the advanced control applications and the optimizer running in closed loop, Polimeri Europa expects to continue achieving favorable returns on their investment over the long-term.

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Biographies

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