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PROCESS CONTROL AND INFORMATION SYSTEMS

Dynamic real-time optimization increases ethylene plant profits

This project resulted in a $12.5 million/yr profit increase and one-month payback

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Considering process dynamics in optimization pays off. These new technologies quickly push a plant to the optimum real-time economic performance, while integrating a plant's dynamic behavior and constraints to maintain fast optimization capability even during process transients. Borealis Polymers Oy reports annual benefits of $12.5 million for its 300,000 tpy ethylene plant in Finland and a new production record. Project payback period was one month.

Closed-loop optimization of processing plants has been practiced with varying degrees of success and failure since the 1970s. The optimization projects of the ethylene plant in this article in the 1980s serve as typical examples.[1] A profit function was generally maximized subject to plant process constraints and product quality specifications. Traditionally, steady-state process models spanning all plant process units have been used.

Consistent matching of such a model to real process measurements and quality data has, however, proved exceedingly difficult. This is due to many reasons, but in particular to the fact that all real plants are frequently in transient states. Consequently, matching can be performed only occasionally, thus undermining the concept of steady-state optimization. A new closed-loop dynamic real-time optimization (DRTO) technology was developed to address this shortcoming associated with the traditional approach.

This article describes the closed-loop dynamic real-time optimization application, which optimizes Borealis Polymers Oy's Porvoo, Finland, ethylene plant profit. The natural solution to incorporate the process dynamics in the model proved successful. The full nonlinear optimization execution of the DRTO application took only one to two minutes and, therefore, qualifies as real-time optimization. The models used span the plant dynamics for over a 10-hr future horizon and were able to swiftly match the observed process conditions and constraints of the plant even while in transient states.

The DRTO has been online continuously since the beginning of 2004 and the experience has been outstanding. Borealis reports annual benefits on the order of $12.5 million due to a substantial increase in production rates. The DRTO project's calculated payback period was less than a month.

Process description. Ethylene plants are highly complex process plants, specializing in olefins production. The essential production process consists of hot and cold sections and is supported by an extensive infrastructure for utilities, logistics, etc. In the hot section, an olefin-rich product mix is produced from varying feedstocks in 10 different cracking heaters, and in the cold section the product mix is separated into high-purity sales products in several multistage processes. The Borealis Polymers Oy ethylene plant, with a fairly typical structure (Fig. 1), has quadrupled its capacity from the original 1970 design to about 300,000 tpy ethylene due to continuous revamping and process development efforts. Several fresh feedstocks ranging from naphtha to liquid petroleum gas (LPG) and recycle feedstocks are fed to different cracking heaters through complex header systems with high degrees of operational flexibility. Recycle feedstocks include ethane, propane, C_4 mixtures and heavier liquid raffinate recycles. The normal decoking cycle is on average roughly once per month. Decoking, logistics and market fluctuations lead to frequent

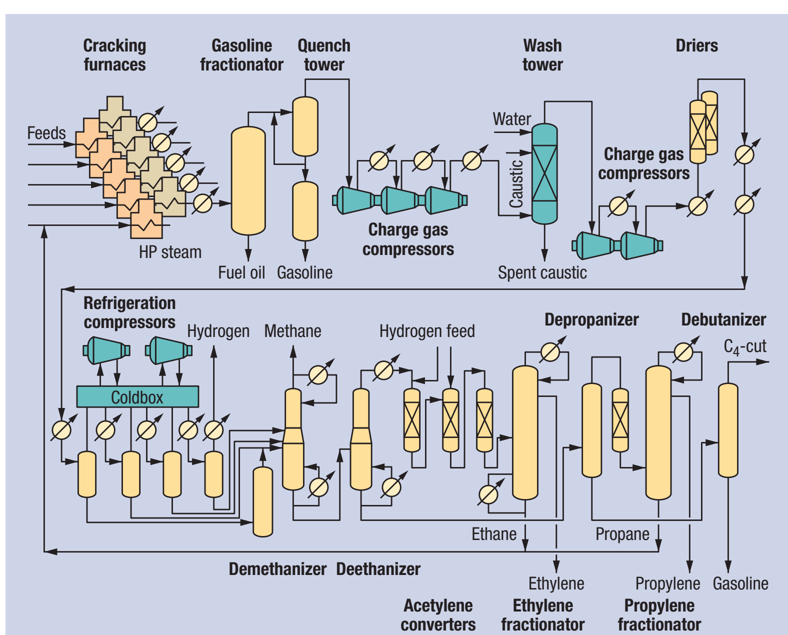

FIG. 1 The ethylene plant capacity has quadrupled from the original 1970 design.

FIG. 2 First page of the panel operator's DRTO six-page display.

substantial production schedule and dynamic process changes. In the 1980s, traditional real-time, closed-loop optimizer (RTO) technology was used at the plant,[2] but later proved impractical to cope with plant operation moving to a more dynamic regime. A 1997 project replaced dated advanced controls with modern multivariable predictive control (MPC),[3] and in 1999 severity and conversion controls were installed.

Optimization project. The real-time optimization project started in 2003 and formed a part of the Borealis program for enhancing plant operations and the economic result. During the project study phase, proposals were requested from different providers. The selected technology was calculated to give the best profitability, lowest investment cost and fastest pay-back time. The chosen technology provides a DRTO of the ethylene plant profit, which was identified as a key success factor. Further advantages included a faster schedule, better robustness and more advantageous total project economy.

The annual project benefit was originally estimated to be at least $1.5 million with the understanding that the total potential could extend to several times more than this. This conservative estimate was based on an optimization of product yields and available feedstocks, corresponding to 3% propylene and 0.5% ethylene increases. The original calculated payback period was about half a year and the internal rate of return (interest) in excess of 200%.

Experience with DRTO technology. The installed DRTO technology proved capable to optimize the plant's profit even while in transient states. The DRTO also integrated familiar features from MPC and rigorous process models with nonlinear optimization in a dynamic framework. The models for the application span plant dynamics for over a 10-hour future horizon, with a resolution of typically one minute.

Using dynamic models in lieu of traditional steady-state models realized several substantial advantages:

- The natural inertia of the plant, caused by various process delays, is accurately taken into account and the models are consistently propagated even during severe transients.
- At each real-time optimization cycle, the model matching the measurements can be immediately completed, without waiting (often hours) for approximate steady states.
- No additional data reconciliation function was needed and illusory measurement errors attributable to plant dynamics were avoided.
- The DRTO dynamic problem definition tools enabled configuring embedded MPC tasks within the optimization, providing economic optimization subject to fulfilling also the control requirements.
- Because of consistent dynamic problem formulation, external MPC and other module results and functions could be exploited, avoiding multiple computations.
- Real-time optimization was achieved, due to efficient optimization and distribution of the computing load.

It is well known that the dynamic optimization problem solution is computationally highly intensive and the matrices exceptionally sparse. The DRTO nonlinear optimization is geared to provide high efficiency in solving simultaneous control and optimization problems exploiting multiple processes and network nodes. Reaching the solution proved so fast that the plantwide dynamic optimization application could be completed in a minute or two. Experience showed, that the models were able to adapt to real process conditions and plant constraints within two minutes on average and four minutes worst case both in transient and steady states.

DRTO technology has proven both accurate and reliable. Plant operation security has been safeguarded by employing the same proven setpoint downloading checking as the MPC used for years. No errors or failures have been logged during more than two years online.

Plant optimization strategy. The objective function of the ethylene plant DRTO is the gross margin (before fixed costs), which in this article is called profit for brevity. This essentially consists of product revenue, feedstock costs, utility costs and other processing costs and revenues, for instance decoking costs and conditional penalty costs.

The DRTO cannot obviously choose a feedstock from the marketplace online nor change the selected feedstock to a heater. These are operational decisions resulting from long-term analysis of offline optimization, decoking sequence and logistics considerations carried through by operators. However, the DRTO immediately recognizes the dynamic effects of operator action, adapts to the resulting new process states, and takes quick optimizing action. All heater operation moves influence the run length. The effect of these on the profit has also been considered for the different heaters and feedstocks. This proved to be very important for optimum adjustment of heater loading throughout the whole run length. Naturally product price fluctuations affect substantially the optimum choice of production strategy. The first page of the panel operator DRTO six-page display set is shown in Fig. 2.

Normally, the maximum profit for an ethylene plant is closely associated with a maximum throughput strategy, but this is limited by the constraints of all process bottlenecks. Each constraint has its characteristic behavior and can be influenced by local process variables. The constraints can also be affected by the independent DRTO variables, such as feedstock feedrates, cracking conditions, etc. So far at least one, and most often two or three, simultaneous process bottlenecks have constrained the plant operation and profitability.

There are altogether 212 direct DRTO constraints, forming the operating window for the plant optimization. Different constraints become predominant for different prices and production strategies while others cease to be active. Typical constraints are: heater run length, cracked gas compressor (CGC) capacity, CGC effluent pressure, propylene refrigerant compressor (C_3 RC) capacity, deethanizer flooding, main product volumes, etc.

In addition, DRTO has 12 variables, which are actually controlled to targets while optimizing the profit. The most important ones are

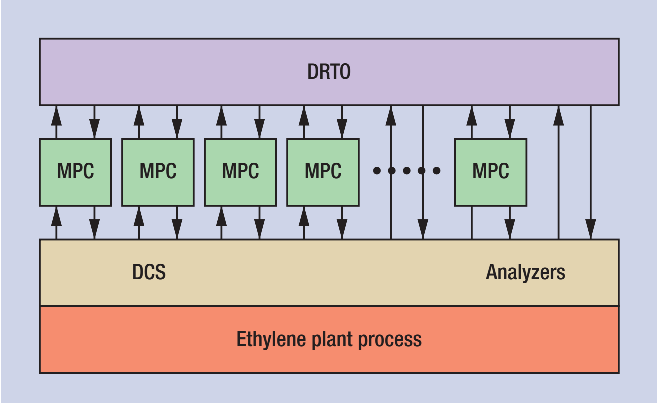

FIG. 3 Data flow upward to the DTRO is significantly higher than the number of moves generated by the DTRO.

level controls related to inventories of plant recycle streams. These embedded DRTO controls are essential for plant dynamics and to close material balances.

Before starting the optimization project, MPCs were in use at most plant process units in the hot and cold sections.[3] These unit-MPC applications were already originally designed to support and work together with a plant optimizer (Fig. 3).

The unit-MPC applications autonomously maximize unit process performance, subject to maintaining product specifications, by adjusting the local process variable set they are equipped with. They also have online-adjustable handles to ease control variables for accommodating maximum production volume while process constraints are active. For instance, in distillation, the flooding limit approach will automatically ease product impurity soft targets to maintain high feed rates. The MPC results are also exploited in the DRTO to decide whether the unit-MPC applications are independently capable to ride the constraints. The DRTO has to take action to ride the constraints if a unit MPC is found insufficient to alone handle the case.

Together there are 185 manipulated variables in the ethylene plant. The great majority of the adjustments made by the DRTO are directed toward unit-MPC parameters. The autonomous unit-MPC applications throughout the plant evaluate the effects of optimization moves at each control interval and download the corresponding required setpoints to the DCS controllers when the effects have propagated to each process unit.

For dynamically optimizing the ethylene plant, cracking heater operations proved to be of high importance. These determine almost the entire product component rates and have a major impact on all downstream units. More than 10 different feedstocks of different price, composition and cracking properties to all 10 heaters with different geometry had to be considered.

Each cracking heater has a unit-MPC that manipulates feed rate, steam rate, fuel gas pressure and effluent temperature for each pass. A heater has 14 manipulated variables to download to the DCS on average. The heater MPC features severity control, total feed rate control, coil outlet temperature balancing, feed rate pass balancing and steam-to-hydrocarbon ratio balancing. It also manages 84 constraints per heater on average. The cracking heater yield, load and run length are influenced in particular by the feed rate, severity, steam-to-hydrocarbon ratio, coil outlet pressure and feed composition. These were found to be the most important heater variables for plant optimization. The rest of the variables were handled by the heater MPC applications and the optimization cycle time could therefore be significantly improved. The supplementary external feed header stream moves, e.g. in the LPG, are directly downloaded to the DCS from the DRTO. All heater effluent streams are combined in the gasoline fractionation section, yielding a cracked gas stream. This stream and the main external feedstocks are fed to the CGC unit. The CGC suction pressure control through the CGC MPC is of particular importance, due to its upstream influence on all heaters. Heaters are also sensitive to effects caused by recycle streams.

Process modeling. All changes in process state are observed by a corresponding process model, which verifies the states and calculates required parameters.

The process models are detailed first-principles models. For instance, the cracking heater model is based on a gas phase tubular reactor model with molar and thermal expansion continuously changing the conditions along the axial direction of the tubes.

The cracking reactor is nonisotherm, where heat is transferred over the thermal resistances to the fluid bulk, driven by the firebox radiation, to sustain the endothermic reactions and the fluid temperature. Long-term changes in heat resistance are caused by continuous coke buildup on the tube inside, which influences the tube metal temperature, the coil pressure drop and temperature profile and, hence, the effluent composition. The number and identity of components in the model is selectable; in this project 38 product components and 35 feed components were defined. Some heavier minor components were lumped. The reaction kinetics and equilibrium expressions are provided with online adjustable parameters.

The online thermodynamic package has proven both very robust and accurate in fluid property and *V/L* equilibrium calculations. In the process model, a set of nonlinear differential equations for mass, energy and momentum conservation in the reactor is solved. The observed state is enforced by matching the resulting model conditions to the measurements and analyzer results using the model adjustment parameters. Each process model typically produces more than 500 supporting calculated variables, such as effluent composition, severity, conversions, run length, profit contribution, tube metal temperatures, pressures, coke thickness, etc. The models are interlinked with other process models, like for instance the five-stage CGC model. The conditioned severity and conversion results are used both in the DRTO and in the severity and conversion control MPC for each heater.

A dedicated recipe database was installed to support the multitude of different heater-feedstock combination cases. The decoking case also has its own special recipe. Each process model adapts automatically to the feedstock type selected for each heater.

For gas feedstocks the heaters have online feed gas chromatographs because the composition of each feedstock varies dynamically. Consequently, the automatic online bumpless transfer of cracking control modes from different conversions to severity and vice versa proved necessary. All heaters also have online effluent gas chromatographs analyzing five effluent components, used in matching the cracking reactor models. In addition, the plant has access to a separate yield calculation software package, which independently estimates also the non-analyzed effluent components with good reliability. The process model can use these estimates as it uses the analyzed components. The most important reason for this type of model matching is that the dynamically changing liquid feedstock composition is not known in real time. The necessary analyzer fault detection, result conditioning and adaptation of the process models, designed to avoid unnecessary bumps in operation, have been used for several years in the severity control modules and have now proved highly reliable also for the DRTO.

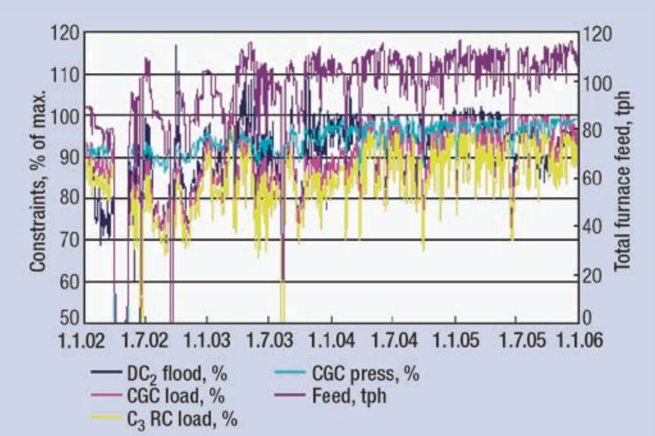

FIG. 4 A sample of the plant operating data for four years including data before, during and after the project. Notice the throughput increase in '03 and the consistent constraint riding since '04. The dynamic nature of the plant is clearly visible.

Project realization. The DRTO project results are a fine example of benefits emerging from seamless cooperation between user and provider. The results speak for themselves.[4]

The entire plant optimization project was completed in 16 months. Borealis strongly emphasized the importance of personnel participation. Besides the actual project group, another user support group was formed to study and comment on the operating tool design and to plan and support the personnel training and organizing of the commissioning. Advance screening of application functioning using engineering judgment and plant know-how also proved to be an important principle. The actual project execution started from the kick-off meeting and ended at the final acceptance as planned.

Two project phases were included in the schedule to support accruing maximum benefits as early as possible during the project. Each commissioning phase was preceded by a commissioning plan and ended by the user taking over the responsibility after a joint review period. Project meetings were once a month and at least weekly during the commissioning.

The first project phase consisted of maximizing production rates by the end of November 2003, or six months after the beginning of the project. Since the DRTO technology supports incremental application building, this proved to be an important additional benefit. It was found that by combining plant personnel know-how and optimized constraint handling, the DRTO was able to increase production rates enough to pay back the total project costs before even completing the first phase.

The second phase, with full economic optimization, was scheduled for the end of June 2004, or about 12 months after project start. In this final commissioning, the optimal tradeoffs of different constraints proved the most important, since many of these are sensitive to the interaction between fluid composition and process conditions. A remarkable breakthrough was discovered during commissioning, when it was found that the observed flooding in the de-ethanizer in particular was sensitive to process conditions and could also be affected by DRTO optimization in addition to the local MPC variables. Consequently, due to the DRTO "buying" a slack to the constraint, more feed was put through the de-ethanizer and more profit was made by accurately riding the constraint.

Other important constraints also becoming active and ridden proved to be the propylene refrigerant compressor capacity, the CGC capacity and effluent pressure. Similar observations were made with other constraints too. For instance, increased profitability over the initial estimates was obtained by DRTO quickly adapting to frequent process operation change swings and by the resulting agile heater feed allocation and severity adjustments.

This is reflected in Fig. 4, where riding different constraint combinations at different times can be seen, in addition to the general plant throughput increase pattern. Heater run lengths varied from 25 to 60 days and consequently, a long verification time for the coking rate models to match the observations was required. Generally speaking, there has been no need to change the coking parameters since the initial tuning period.

In the final acceptance tests, the influence of different price scenarios was also studied and found to agree well with offline optimization benchmark results.

Results. The DRTO for the Borealis Porvoo ethylene plant has been in continuous use since the beginning of 2004. The user comments are briefly that the DRTO is always on when the plant is on, which underlines the importance the personnel are feeling about this tool. This is also reflected in the high actual service factor figures: the DRTO was online for 99.7% of the time in 2005, and the total average service factor for all installed independent (optimized) variables was 96.2%. Another measure of actual performance is that in 2005 the DRTO was operating against 1.8 simultaneous process constraints on average. There has also been a clear positive effect on plant reliability figures, reaching full 100% during 2004 and 2005. There has been no need for DRTO software maintenance and only one small application scope change was performed in 2005 to exploit a new opportunity found in the CGC condensate stripper. The main DRTO maintenance activities required from the sole part-time control engineer in the plant have been the normal monitoring of operator DRTO use and surveillance of the instrumentation and analyzers.

The optimization has proved very agile to follow the dynamics and has operated the plant in a way that has been easy to justify and closely equivalent with the offline optimization results.

The original economic and functional targets set for the project were surpassed during and after the project by an ample margin. The total result was that ethylene production rate increased from 840 tpd to more than 920 tpd on average or about 9.5%. Due to the corresponding increase in the total feed rate, propylene and other product rates also increased a few percent. In 2004 a new yearly ethylene production record of 327,000 tons was set. The improvement to the previous record of 298,000 tpy was 29,000 tons or almost 10%. The economic benefit during 2004 amounted to $12.5 million. This was mostly due to increased production and the smart feed allocation of feedstocks with different prices and cracking characteristics. It surpassed the original asserted benefit estimate several times. The corresponding calculated payback period for the investment was, therefore, less than one month.

The Borealis Group technology board awarded the project group an innovation award for outstanding performance in ethylene plant optimization in 2005.

Outlook. Ethylene plants are complex processing facilities and optimizing their performance taking into account the plant dynamics is largely an unexploited opportunity. The DRTO proved able to consider the plant dynamics and to optimize plant operations within a minute or two, or essentially in real time. The entire project was carried through in 16 months and the DRTO has been in continuous use since the beginning of 2004. Due to the increase in production rate the economic benefit was $12.5 million/yr.

It is anticipated that such an example may inspire the industry and practitioners to more closely examine opportunities to be exploited in like cases. The technology used is currently in service in three other plants: a refinery unit, a chemical plant and another petrochemical unit, which emphasizes the general applicability of the concept. **HP**

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