Consistent product quality control (up for delivering the following benefits: and has proven its payback. Applying predictive control (MPC) technology has become commonplace in these plants and hence refrigeration capacity. 

Ethylene is the largest volume commodity chemical produced globally, at close to 120 million tonnes annually, and is the core building block for most organic chemicals. Modern mega crackers (ethylene plants) produce in excess of 1.2 million tonnes/year. Feedstock for ethylene plants range from ethane and ethane/propane mix (E/P) to heavy naphtha and vacuum gas oils. Liquid crackers are designed with raw material flexibility in mind to take advantage of feedstock economic opportunities that arise from time to time, while gas crackers utilise ethane.

Ethylene produced is used in the production of polymers and other ethylene derivatives such as ethylene oxide and glycol. Depending on the plant location, the ethylene produced could be supplied to an ethylene product pipeline grid, a dedicated derivatives unit, refrigerated storage for export, or a combination of the these. Plants that produce for dedicated downstream units are more challenging to operate, as they have to continuously adjust to the demand swings of the downstream facilities by increasing their refrigerated storage and hence refrigeration capacity.

Implementation of advanced process control (APC), with multi-variable predictive control (MPC) technology has become commonplace in these plants and has proven its payback. Applying APC solutions is generally responsible for delivering the following benefits: Consistent product quality control (up to specifications) allows the plant to increase the content of less valuable components into more valuable products. For example, additional ethane can be included in the ethylene product from the C₂ splitter and propane into propylene product in the C₃ splitter. By reducing quality standard deviation, the target can be brought closer to specifications, hence increasing the impurity contents. This also lowers refrigeration duty demand in the condensers, and hence energy cost. Consistent product quality is also important for downstream units. Recovery of more valuable components from less valuable streams. Examples are recovering ethylene in tail gas, ethylene in recycle ethane and propylene in recycle propane.

Maximise ethylene and propylene gains in the acetylene and MAPD converters.

Severity control improves furnace yields and run-length. Minimising furnace excess oxygen in the furnace controller reduces fuel gas consumption.

Addition with constraint control maintains the plant at limits.

The application of closed-loop optimisation above the advanced control layer delivers the following benefits:

—Production maximisation to plant physical constraints; this is especially challenging when plant constraints are in the back-end separation area. —Yield optimisation to increase the more valuable products while maintaining furnace run-length —Energy optimisation to minimise energy consumption while trading off against product recoveries and capacity —Stabilising overall plant operation, especially during disturbances (such as feed quality changes, recycle stream fluctuation from outside the battery limits and ambient conditions), and transient operations (such as co-crack feed changes, furnace decokes and reactor and dryer switches).

Typical demonstrated benefits are a 4% to 6% increase in olefins production, depending on an individual plant’s situation, and up to 10% in energy reduction. Typical ROI for projects like this is less than one year. Optimisation benefits usually increase with increasing furnace severity. This comes at the expense of furnace run-length, however, as coke deposit in furnace tubes is accelerated with higher severity or cracking temperatures. Severity is therefore constrained in the optimiser to maintain an acceptable run-length. The severity constraint is manually adjusted in the optimiser over time to reflect a particular furnace days on line and mechanical conditions. This severity or tube metal temperature (TMT) profile over the run length of a furnace is an output of the furnace schedule.

This profile is typically the result of a manual scheduling process. Given a forecast of product demand and feedstock availability, the scheduler must determine the run-length for each furnace as well as the severity profile over this run length, while ensuring the satisfaction of all constraints including storage availability. The results from this schedule are rarely coordinated with the control systems.

Closed loop optimisation

The decision variables for an ethylene plant closed loop optimiser to achieve the previously described benefits are: Furnaces, Individual furnace feed, Individual furnace severity, Individual furnace steam/hydrocarbon ratio, Converters, Individual reactor loading, Reactor inlet temperature, Reactor CO addition (if applicable), Compressors (CGC, ethylene and propylene), Suction pressures, Columns, Pressures, Recoveries of economic ends (such as ethylene in tail gas and recycle ethane), Utilities, Boiler load allocation, Turbine load allocation.

To find an optimum for these operational targets a relationship is needed between these degrees of freedom and plant economics as well as constraints and an optimisation algorithm. These relationships are typically non-linear such as furnace cracking yields to furnace operating conditions.
**Figure 1** Equation-based optimisation approach

**Conventional technology**

In the 1980s, given the limitation of computing power and sophistication and robustness of mathematical solvers, regression based models for cracking kinetic relationships and simple recoveries for the separation models were used along with LP or SLP solvers. These systems were able to run at a high frequency, due to their simplicity, and deal with plant disturbances, but were unable to push to physical limits to achieve maximum potential benefits due to their simplified conventional or steady state models.

With the exponential improvements in computing power, and the need to model process non-linearities, these models were replaced in the 1990s with steady state rigorous, fundamental engineering based models. These models were written in open equation format and solved simultaneously with advanced SQP solvers that utilise sparse matrix techniques. The advantages of these systems were non-linear representations of the process and synergies between design models and online optimisation models. A user could utilise an existing simulation model to build the optimiser flowsheet and obtain consistent results between off-line usage and online closed-loop usage.

The advances in SQP solvers and sparse matrix techniques allowed the solution of a set of equations on the order of 200,000 in approximately 30 minutes. Also, the open equation formulation allowed for the same model use for simulation, parameter estimation and optimisation by simply changing variable assignment from fixed to optimise.

Steps for system execution are shown in Figure 1 (equation-based optimiser execution cycle). For a model-based solution, even with very high model fidelity, models cannot be continually used without periodic matching to the process. This update is required to account for mismatch that comes from changes in the plant over time. Equipment fouling and catalyst de-activation are examples of the sources of these mismatches.

The first step of the execution cycle is steady state detection. A steady state snapshot of the plant data is needed to update these models. This data is checked for gross errors and reconciled before the parameter estimation step takes place. Optimisation is then performed on the updated models to find a new optimum state. The plant is checked to ensure that it is still at the same steady state identified in step one for a valid solution. Finally, the new optimum is passed in small step sizes at every execution cycle to the advanced controllers below.

The characteristics of Ethylene plants, however, are very dynamic and the plant is seldom at steady state due to issues such as:

- Frequent feedstock (or co-crack) changes due to spot market opportunities create incentives as well as disturbances
- Frequent feedstock quality changes (naphtha or gas oil) ripple through the distillation train
- Frequent product demand changes (especially for integrated complexes to downstream polymer units that go through a lot of grade transitions) requiring rapid response to refrigerated storage diverted production and pressure fluctuations
- Frequent furnace decokes and reactor and dryer switches create havoc in the distillation back-end
- Rapid ambient condition changes can lead to opportunities for refrigeration-limited plants
- Additional process disturbances

This meant a much lower frequency of execution for these steady state optimisers, and hence, lower benefits. In addition, solving some constraint violations were beyond the control of a local controller and required adjustments to overall plant feed or severity by the optimiser.

The absence of the optimiser at these times, due to the plant being unsteady, resulted in the operator being required to take action and therefore undermining his confidence in the system. Additionally, these steady state optimisers, through a lack of understanding of dynamics, could not properly handle the phasing of the optimum results to the underlying dynamic controllers and had no direct way to control optimisation speed other than through limiting step sizes. This, however, for a very large system is a complicated task given that aggressive step sizes could lead to dynamic in-feasibilities and small ones will lead to a slower optimum implementation.

In summary, an on-line system should also be able to address, along with the usual issues, requirements like optimisation that incorporates plant dynamics in their models, and a structured way to deal directly with optimisation speed and how to incorporate the optimum solution into the MPC controllers to avoid conflicts between layers.

These challenges led to various creative approaches with steady state optimisers to better deal with the dynamic issues. One approach was to divide the plant into separate envelopes that had similar dynamics. This allowed for higher steady state detection frequency and, based on state results, either an update of the models or optimisation based on parameters from a previous model update. Others experimented with projecting steady state from a dynamic system and biasing the steady state models.

The advantages and disadvantages of this solution have been reviewed for some time. There is a growing consensus that a pragmatic, robust solution that can address the dynamics of the plant, and be easily maintained, shows significant advantages over the conventional optimisation approach [Freidman Y Z, Closed loop optimisation update – a step closer to fulfilling the dream; Control, January 2000].

**Dynamic technology**

The proprietary Profit Optimiser has been developed for closed loop real time optimisation (CLRTO). Instead of a steady state optimisation model this optimiser has a dynamic model of the process. The dynamic optimisation model is not developed from scratch. Instead, the starting point for the optimisation model is the MPC control model of the process.

A limited amount of additional modelling effort is required to tie together
various controller models using dynamic bridge models, source-clone models and combined constraint models. These models add the dynamics between all of the individual controllers and hence process equipment. Rigorous steady state models integrated with the dynamic system are used to update gains in both the APC and optimisation layers to account for non-linearities, providing an effective, hybrid solution.

Since MPC is a prerequisite for optimisation, use of MPC models as a base optimisation model represents a significant savings of engineering effort as the steady state models required are far less in scope than those of a complete closed-loop steady state optimiser.

As Profit Optimiser is based on a dynamic model of the process, steady state operation of the plant is not a prerequisite for optimisation. At any given time, the optimiser is aware of the various transients in the process and has a prediction for their consequences - that is, it predicts the steady state values for all variables.

At each execution of the optimiser, process feedback is obtained. This frequent process feedback compensates for model mismatch and forms the basis for data reconciliation. Profit Optimiser allows gains to be updated “on the fly”. For process areas where non-linearities are significant, such as the reaction system, frequent gain updating is performed using non-linear models such as cracking yield models.

In this framework, the optimisation problem is formulated as a semi positive definite quadratic programming (QP) problem for which solution is fast and guaranteed. QP formulation along with frequent gain updating in essence amounts to an SQP type of optimisation. The optimiser typically executes at the frequency of MPC controllers, usually once every minute.

In this approach, there is a tight integration between optimisation and controls. Not only does the optimiser pass down the optimum desired values to the controllers, it also passes down the optimisation speed factor to each of the controllers that takes into account the dynamic characteristics of each of the controllers corresponding to the plant characteristics. The overall speed of optimisation implementation is directly determined by the user, with one intuitive tuning parameter that is a function of plant settling time.

This overall speed, along with the patented cooperative technology, generates the individual speed passed on to the controllers with the optimum targets.

This dynamic optimiser moves the process towards the optimum along a minimum energy path that eliminates optimiser induced dynamic violations of controlled variables. In addition, the dynamic optimiser helps the process better cope with disturbances by applying corrective action to multiple controllers as and when needed in a coordinated fashion.

New approaches are being developed to improve furnace scheduling using multi-period, mixed integer optimisation technologies to solve the quality and logistics problems associated with furnace operations [Vasbiner R and Kelly J]. Ethylene furnace scheduling: ERTC Petrochemical Conference, Vienna, October 2004].

These scheduling solutions take into account the furnace mechanical constraints and maintenance schedule, cracking conditions, logistical constraints and utilise a coking and yield model correlation, or rigorous models such as the proprietary SPYRO. The resulting profile, then, becomes a constraint space for the online optimiser to manage within and the resulting schedule is implemented to maximise furnace capacity and stability. This can significantly improve the coordination between scheduling and control.

In summary, the Profit Optimiser is an effective technology for optimisation of dynamic processes such as ethylene plants. Since this optimiser is based on dynamic process models, steady state operation for the process is not a requirement for optimisation. The optimiser formulates a semi positive definite QP that can be solved efficiently with guaranteed optimum, typically at the same frequency as the controllers. The optimiser allows for on-the-fly gain updating, which in essence makes it a SQP optimiser.

Since the controller models provide the bulk of the optimisation model, initial development efforts, as well as long-term maintenance efforts, for the optimiser are greatly reduced. The online factor is also now the same as the underlying controllers.

**Implementations and results**

The first ethylene plant optimisation project using proprietary Profit optimisation technology was implemented by Honeywell in October 1997 at Petromont in Varennes, Quebec, Canada. The scope of the project was for APC and CLRTO, also utilising SPYRO yield models for the furnaces from Technip USA. This project was completed in November 1998 and exceeded the predefined objective of increasing plant ethylene production capacity.

This optimiser has been running utilisation rate over 90% since initial deployment.

Figure 2 shows typical debottlenecking benefit of olefins production after commissioning of Profit Optimiser. Note that sustained increase in olefins production is due to two effects: improved selectivity in olefins yield, and ability to increase feed throughput.

Numerous other projects in ethylene, and other processing plants, have been implemented since the initial application. These projects have demonstrated significant benefits while increasing optimisation robustness and reduced life cycle cost.

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