ON-LINE OPTIMIZATION OF A PARAFFINS SEPARATION PLANT

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Abstract

This paper investigates the on-line optimization of an existing petrochemical plant. The process consists of a train of two distillation columns where a group of n-paraffins is separated into kerosene and lights. Because of process heat integration and recycles, a significant interaction between both columns is observed, which makes very complex the adjustments of process operation to the changes in feed conditions. Once developed the steady-state process and economic models for this scenario, the operability and the economic behavior of the process are evaluated by implementing on-line optimization over the dynamic simulation of the process. The dynamic model emulates the plant behavior when it is subject to disturbances with economical influence. This scenario is used to test two on-line optimization strategies: the conventional RTO methodology and RTE (Sequeira et al., 2002). Such validation scheme results very useful for an early identification of real process improvements and early problem identification. Furthermore, the validation procedure also shows the benefits offered by the latter strategy.

Keywords:

On line optimization, Real Time Optimization, Model-based on line Optimization, RTE.

Introduction

On line optimization using steady state models is still an attractive field of computer aided process engineering (Marlin and Hrymak, 1997; Perkins, 1998). The decrease in hardware and software costs has resulted in several implementations of this technology, showing quite attractive economical results (White, 1998). Research efforts in this area have been focussed on specific components of the system, to mention data acquisition and validation, gross error detection, data reconciliation and, of course, modeling and optimization. Besides, particular attention has been paid to the effects of uncertainty and noise over the final implementation. However, most of published studies are based on analysis of steady state models, with scarce attention to the possible consequences of the use of an RTO system in the process dynamics, which doubtless is significant.

This work focuses in the results obtained for an existing petrochemical plant when validating an on-line optimization system using dynamic simulation. It includes the use of a classical RTO system, and the RTE (Real Time Evolution) approach proposed in an early work. Firstly, the main ideas of RTO and RTE are briefly explained. Then, the process under study is presented along with the optimization analysis. After that, the validation methodology and the results obtained for a particular situation are presented. Finally, the results are discussed and conclusions are obtained.

RTO and RTE fundamentals

In a classical RTO, once the plant operation has reached steady state, plant data are collected and validated to avoid gross error in the process measurements, while the measurements themselves may be reconciled using material and energy balances to ensure consistency of the data set used for model updating. Once validated, the measurements are used to estimate the model parameters, thus ensuring that the model correctly represents the plant at the current operating point. Then, the optimum controllers' set points are calculated using the updated model, and are transferred to the control system after a check by the command conditioning subsystem.

Real Time Evolution has been introduced as an alternative to current RTO systems. The key idea is to obtain a continuous adjustment of set point values, according to current operating conditions and disturbance measurements (those which affect the optimum location) using a steady state model. Thus, rather than coping with an optimization problem, an improvement algorithm evaluates the neiboourhood of the current operating point to continuously identify the best direction to move. The steady state information is used by RTE for data reconciliation and model updating, while the core of the system is the recursive improvement, which does not need the process to be at steady state.

The Process

It consists in a train of two distillation columns where a group of paraphins mixture is separated from kerosene (Figure 1). The feed is a mix of hydrocarbons (paraphins containing significant quantities of aromatics, iso and cyclo-paraphins and some olephins) that is preheated in a heat exchanger (REC) which takes advantage of a lateral extraction of the Redistillation column (T-2). The light hydrocarbons (less than C-10) and sulfur are separated in the stripper (T-1) at the top (Naphta). The stripper's bottom is fed to the Re-distillation column, where the main product (Light) containing lineal hydrocarbons (C-10 to C-14) is obtained at the top and heavy kerosene is obtained as byproduct at the bottom. The interaction between the two columns is due to the energy exchange between their streams and the connection of the stripper's bottom, stream to the re-distillation column. Further details of the process are withheld for confidentiality reasons.

Steady State Model

The steady state model has been developed using Hysys.Plant® sequential modular process simulator. The model has been conceived for process simulation under the main input: feed conditions (flow, temperature, pressure and composition) and the following specifications: for the Stripper (T-1), bottom and top pressures and heat to reboiler (*QReb-1*); for the Redestilation Column (T-2), bottom and top pressures, the lateral extraction flow (*Draw*) and finally the split fraction corresponding to the draw extracted (Sf_E). Additional "planning decision" specifications are the mass flow ratios of the three final products with respect to the feed. They are introduced as column specifications in T-1 (Naphta) and T-2 (Heavy), while the last one (Light) is automatically satisfied meeting the global mass balance. Besides, there are other planning specifications related with product quality caraterization [API gravity index, Normal Boiling Point (NBP) and mean molecular weight (MMW)].

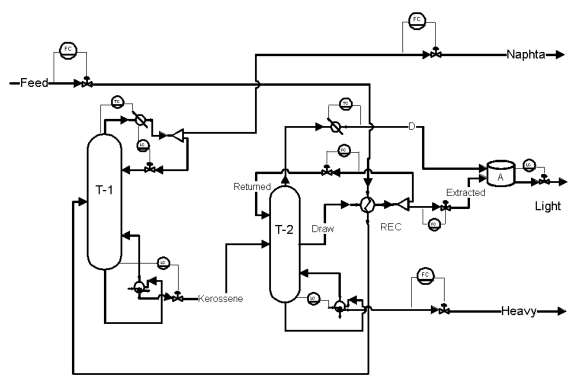

Figure 1 Simplified Process Flowsheet

The Performance Model

The performance model describes the influence of operating conditions over the process economical and quality performance. Although the problem is multi-objective in nature, both aspects were aggregated in a single objective function. The instantaneous objective function (IOF), to be maximized is:

$$IOF = R - (H + C + RM) \tag{1}$$

where: R are the revenues obtained for selling the products ($/h); H represents the heating costs ($/h); C the cooling costs ($/h); RM the raw material costs ($/h). To take into account the products quality, a quality index is computed using the quadratic difference between the nominal and the current value of the quality-characterizing parameter (API, NBP and MW), affecting the revenues term in penalizing economic terms. Thus, the selected IOF reflects the trade-off between energy consumption and product quality in a simplified way.

The Optimization

The optimization objective is to maximize the selected objective function, by modifying the specifications (decision variables) and observing a set of boundary constrains, for decision variables and product quality indexes. Besides, an additional operation constraint establishes that at least 20% of the light product should come directly form the tower T-2 (see stream D in the flowsheet). This latter constraint has been used to link two of the decision variables in the form of an inequality ($Sf_{E,}$ the Extracted split fraction, and *Draw*). However, a sensitivity analysis indicates that this inequality is always active at the optimum point. Therefore, it has been explicitly included as an equation to reduce the degrees of freedom from three to two (*Qreb-1* and *Draw)*. Both variables have been included as ratios to the current feed flow, to avoid non-desired variability to changes in the most frequent disturbance. Feed conditions are the main disturbances considered from the optimization standpoint.

Off-line Results

The base (reference) case corresponds to the nominal operating point, where the objective function has shown to be convex. Additionally, the sensitivity of the optimal operating conditions has been evaluated off-line introducing variations on the feed conditions. A pronounced change of the optimal operating point has been found with the variability in feed temperature and composition.

On-line Results

A dynamic first principles model is used to emulate "on line" data, which are consequently already validated, filtered and reconciled. The RT system includes the following components: the steady state detector used for model updating, the steady state process model and its associated performance model, the solver (in RTO is an optimization algorithm while in RTE it is just the improvement algorithm) and finally the implementation block that sends the generated set points to the plant, if they are acceptable.

The dynamic model has been developed using Hysys.Plant, and for the communication with the RT system block the DCS interface has been used. It should be noted that the development of the dynamic model, besides a considerable degree of effort and expertise, requires the specification of a significant number of additional parameters, mainly related to the relationships between flow and pressure changes (an aspect rarely considered in steady state models although it is a potential source of plant-model mismatch). The dynamic model includes the whole control layer. All controllers are proportional-integral. For the sake of simplicity, some assumptions were made, that do not compromise the simulation results: perfect control for loop QC-Reb1, perfect control for top pressures in both columns, using two vent streams in T-1 condenser and tank A, and perfect control for the heat exchanged in REC (QRec).

Several experiments have been performed, simulating disturbances in input conditions and comparing the results obtained when using RTO and RTE, and when taking no optimizing action. Following, the case and results corresponding to a step change in compositions (10 % in average) and set point of feed flow (5 %) are commented.

Without Optimizing Action

This situation does not correspond to keeping every controller with a constant set point. It corresponds to maintaining constant the *decision variable* values rather that the set points. For this case study, the decision variables are related to the feed flow, and therefore the set points for the controllers: QC-Reb (T-1), FC-Returned, FC-Extracted and the QRec will change proportionally to the feed flow in order to keep constant the decision variables (*Qreb1*, *Draw* and Sf_E). In practice, this situation is handled by the plant operators or more commonly by ratio controllers at the supervisory control level. The latter approach has been used during the simulations. The situation for the temperature controllers is similar, but in this case the model used to evaluate the set point changes has been the same steady state model for optimization instead of a ratio controller (i.e. a nonlinear correlation).

Simulation results indicate that the system performance is reduced as a consequence of the disturbance, and that approximately after 200 minutes the system smoothly reaches the new steady state.

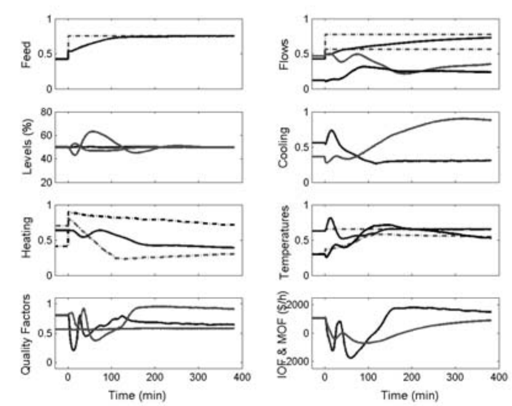

Figure 2: RTE

RTO

Approximately after 200 minutes, the steady state is detected, and optimization takes place. The resulting optimal operating point is implemented as a bounded step in the associated set points.

It is worth noting that the simultaneous implementation of set points may produce a more significant system perturbation than the produced by the individual changes, besides the increase in the settling time (now about 400 min). An important consequence has been the appearance of small amount of vapor phase in some liquid streams, which compromise the performance of the corresponding flow controllers as have been seen in a noisy behavior of some variables' curves. Otherwise, the plant performance, in IOF terms, is substantially increased after 250 min.

RTE

RTE has been executed every 4 minutes, allowing a maximum decision variable change of 1%. Note that immediately after the disturbances occur (Figure 2), the set points are updated by the ratio controllers. Then the system is progressively improving the operating points, (not necessarily by straight lines). Besides, like in the RTO case, the process performance has been improved and an increase in the settling time can be also observed; although lower than that of the RTO case.

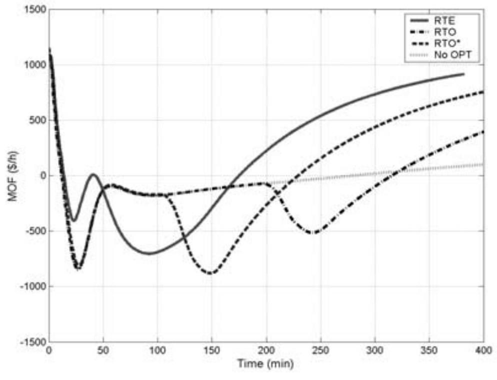

Figure 3: MOF's

Final comments

Figure 3 shows the Mean Objective Function (MOF), defined as:

$$MOF = \frac{\int_{t_0}^{t} IOF(t) \cdot dt}{t - t_0} \qquad (2)$$

(in other words, the mean performance during the interval *[t_0, t]*) for the three situations: without optimizing action, RTO and RTE. It is clear that as time goes by both RTE and RTO strategies improve the process performance. However, the improvement produced by RTE is obtained sooner, thus making the system ready to deal with other disturbances without deteriorating the process performance. There is an additional curve in Figure 3 describing the MOF behavior for an RTO system with a less "strict" steady state detector (dashed line, RTO*). Although the process performance is improved faster, the minimum is more pronounced, which reveals a more aggressive behavior of introducing a new disturbance when the system has not completely overcome the initial one.

Conclusions

This work presents briefly some results obtained in the validation of an on-line optimization system for an existing scenario. The proposed validation scheme, involving dynamic simulation, gives a complementary vision of the system performance along with the typical sensitivity and uncertainty steady state analysis. It has been illustrated how the RT system introduces by itself a disturbance on the system, which may compromise the process performance without actually risking the process operation. Besides, it has been also shown the superior economical and operational behavior of the RTE approach, in a real industrial scenario. The system analysis has shown that the high degree of integration of the stripper and redistillation columns makes the process transient too long, and therefore, an improved control strategy may help the use of on-line optimization, becoming it more adequate and giving it the ability of coping with more frequent perturbations. Thus, future work includes the re-engineering of the control layer

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