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A Study for the Implementation of an Economic Optimization of Coking Plants

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Lately, the processing of low quality crudes is more frequent than ever and the heavy feedstock is providing high quantities of vacuum residue. The delayed coking unit is the main converter of residue and usually a refinery that owns a Coker is considered to be residual free (Sawarkar et al., 2007). Our work investigates the delayed coking process due to the new environmental conditions for end products and because economic optimization based on model predictive control (MPC) is required for plant flexibility.

1. Introduction

During the last few years a new trend of processing heavy crudes appeared because of their prices and availability, despite all the problems generated by the procedure. The delayed coking is one of the few processes able to convert heavy products (atmospheric and vacuum residues) into lighter ones with high economic value: gases C1 – C3, gasoline, gas oil, heavy distillate and coke. Also, in some situations the delayed coking process produces needle coke (used in graphite electrode manufacturing), when the feedstock and certain operating parameters (residence time, reaction temperature and pressure) are appropriate. Figure 1 shows the pricing evolution for needle coke. Apparently its demand will increase constantly as long as steelmaking continues to grow (electric arc furnaces are an efficient tool in steelmaking from an economic and environmental point of view). In this situation, a proper operating strategy and a real-time optimization system will assure a perpetual profit.
Some new terms are used lately in the process control community associated to the control structure design (including economic optimization), like: plant-wide control (Larsson and Skogestad, 2000), enterprise-wide optimisation (Grossman and Furman, 2009), site-wide optimisation (Zhang and Zhu, 2006), one-layer strategy and RTO with MPC (Souza et al., 2010).

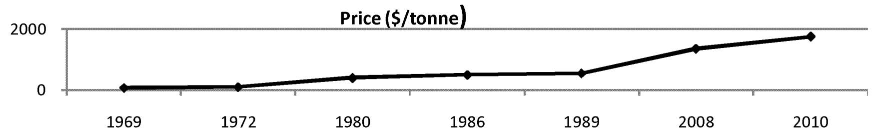

Figure 1: Pricing evolution for needle coke (Sawarkar et al., 2007)

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In practice, the overall control system is fragmented into several layers (Larsson and Skogestad, 2000) structured in a hierarchy: regulatory control (seconds), predictive control (minutes), real-time optimization (day) and enterprise-wide optimization (weeks).

There are several papers in the literature about the optimization and advanced control of delayed coking plants. Most of them represent real-world case studies, successful implementations of different techniques (Chang et al., 2001; Elliott 2003; Wodnik and Hughes, 2005; Biondo et al., 2004, Friedman, 2005; Haseloff, et al., 2007, Nam et al., 2010; Pordal, 2001). On the other hand, there are only a few papers focusing on the energy recovery systems (Chen et al., 2004; Gareev et al., 2001).

2. The delayed coking process

Figure 2 is a schematic representation of a delayed coking unit from a Romanian refinery. The vacuum residue (1) represents the fresh feed which is heated up in the convection zone of the furnace (CF) at 315-320^{0}C and then it is inserted at the bottom of the fractionator (F). Here, some lighter fractions are removed as side streams (naphtha (4), LCGO (3a), HCGO (3b)) and the mixed feed from the fractionator's bottom (2a) is redirected to the radiation section of the furnace (CF), where it is heated up to a high cracking temperature (490-495^0 C). In order to avoid the coke deposition on radiation tubes, high pressure steam is introduced in the furnace tubes, with a minimum flow of 500 kg/h. The outcome of the radiation section of the furnace (2b) is a partially cracked product which enters the coke drums (CD) where cracking continues. Each drum is filled with coke for approximately 20 hours, depending on the input flow and CCR (Conradson carbon residue). The remaining coke is periodically removed from the drum.

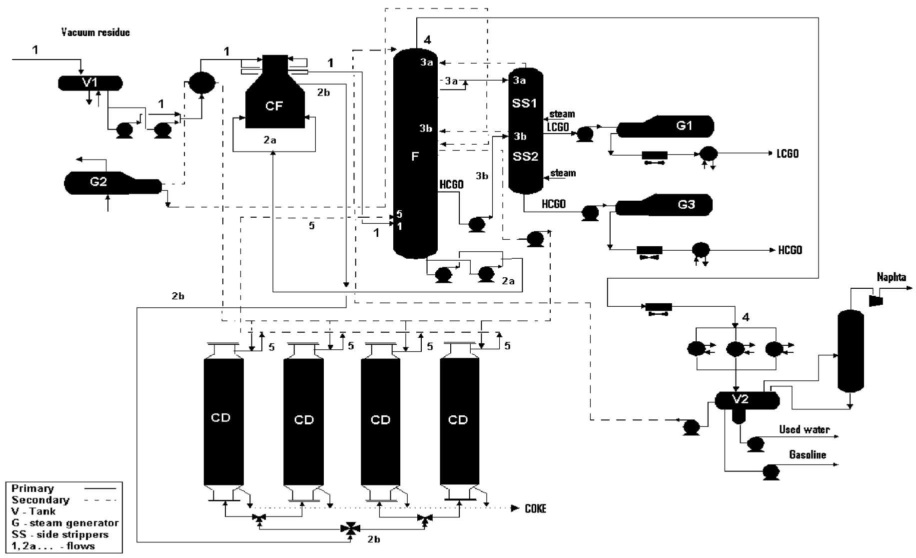

Figure 2: A schematic representation of a delayed coking unit

Although such an installation is designed to work with two parallel drums simultaneously, this one is incapable of processing the maximum flow of about 140t/h, so it uses only one drum at a time. The two main units that are under the influence of drum disturbances are the furnace (CF) and the fractionator (F).

In order to implement an economic optimization using MPC, a steady-state mathematical model is a minimal requirement. We have built a general optimization structure in DeltaV and PredictPro which uses a steady-state "yield matrix" to maximize the objective function, within the constraints.

A generally accepted modelling practice is to express the feed composition of the furnace and fractionating tower in terms of a finite number of pseudo-components. We will assume in our model a 25 pseudo-components feed.

2.1 Thermal cracking furnace

The cracking furnace is the centre of the coking plant and its outlet temperature influences directly the quality of end products and the coke deposits on the walls. The major concern about the furnace is its negative economic impact because it needs decoking every few months. The primary issue is to maintain a constant outlet temperature in order to decrease the deposits on the walls and implicitly maintain the metal tube temperature and internal pressure constant.

The mathematical description of a Coker as a one-dimensional plug-flow reactor is represented next by a set of coupled ordinary differential equations, considering a laminar regime, axial dispersion neglected, ideal gas behavior and inert steam diluents:

Material balance:

$$\frac{dF_i}{dz} = -r_i \cdot A_{liq}, \left[i = 1 \dots n_{pseudo}\right], n_{pseudo} = 25; \tag{1}$$

Enthalpy balance:

$$\frac{dT}{dz} = \frac{1}{G \cdot \overline{C_p}} \cdot \sum_i r_i H_i + \frac{4}{D \cdot G \cdot \overline{C_p}} \cdot \left[C_1(T_w^4 - T_e^4) + C_2\left(T_g^4 - T_e^4\right) + U(T_g - T_e)\right] \tag{2}$$

Momentum balance:

$$\frac{dP}{dz} = \frac{x^2}{\rho^2 \cdot g} \cdot \frac{d\rho}{dz} - \frac{2f \cdot x^2}{\rho \cdot g \cdot D} \tag{3}$$

2.2 Fractionation column

There are several possibilities to create the optimizing model for the fractionation tower. By using the MESH equations (Material balances, Equilibrium relations, Summation equations -i.e. molar fractions sum to unity- and entHalpy balances) for each tray we can obtain a steady-state model by setting the derivative term of all of the differential equations to zero (Cheng, 2006).

So, the equations for a multi-component mixture, for a general stage to be used in the steady state model will be (Skogestad, 1997; Grassi, 1992):

Overall material balance:

$$0 = F + L_{j+1} + V_{j-1} - L_j - V_j - S, [j = 1..n_{trays}] \tag{4}$$

Component balance, considering $[i = 1..n_{pseudo} - 1; j = 1..n_{trays}]$:

$$0 = z_{i,f}F + x_{i,j+1}L_{j+1} + y_{i,j-1}V_{j-1} - x_{i,j}L_j - y_{i,j}V_j - z_{i,s}S, \tag{5}$$

$$\sum_{i=1}^{NoComp} x_i = 1, [i = 1..n_{pseudo}; j = 1..n_{trays}] \quad (6)$$

Energy balance:

$$0 = h_f F + h_{j+1}^L L_{j+1} + h_{j-1}^V V_{j-1} - h_j^L L_j - h_j^V V_j - h_s S \pm Q_j, [j = 1..n_{trays}] \quad (7)$$

Vapour-Liquid Equilibrium:

$$y_{i,j} = K_{i,j} \cdot x_{i,j};\ K_{i,j} = \frac{\gamma_{ij} \cdot P_{i,j}^S}{P_j} \quad (8)$$

A set of simplification assumptions have been made: constant pressure, no vapour hold-up and perfect mixing on the trays. The equations could be used as presented here, or by applying several ways of grouping the variables in order to find the numerical solution of the set of nonlinear algebraic equations. By applying these reduction methods the computational time is reduced and the optimization requires less time to converge.
The implementation of these models is able to provide a "gain matrix" to be used in defining the model inside the MPC.
We have built the structure from Figure 3 in DeltaV Control Studio which is used as an on-line economic optimizer system composed of several modules, with the MPC (OPTIMIZER) in the center of the application. The system represents a framework for almost any type of economic optimization for processes that can be described by a "yield matrix". We can also simulate nonlinearities by manipulating the gain matrix at every optimization scan, using the vectors GAIN_FACT and the module NON_LIN.

3. Control structure design and Real-Time Optimization (RTO)

The main reason why delayed coking unit's operation, control and optimization are so complicated is that drums (CD's from Figure 2) are accumulating coke and afterwards are switched off for coke removal, so the continuous process becomes a batch one (Friedman, 2005 and Nam et al., 2010). Also, the unit operates at very high temperatures and when an empty drum is connected to the system some problems appear because it's heated using hot cracked vapours from the other drum, so the quantity of vapours sent to the fractionator (F) is decreased.
According to Nam et al. (2010), the coking plant should work properly in two different situations, at drums switch (case 1) and at normal operation - steady state (case 2).

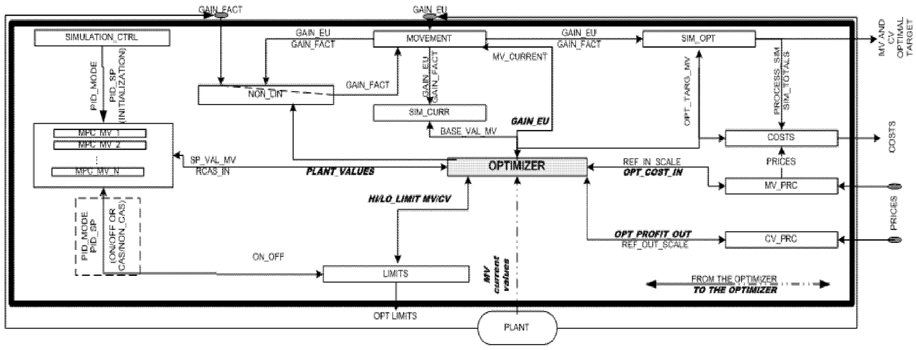

Figure 3: Optimization System's Modules

In both cases, the main control objectives are to maximize feed/throughput and also the middle distillate yield, while rejecting the periodic drum switch disturbance (case 1) and keeping the unit within equipment constraints (case 2). As handles, we use the fuel-oil feed to the furnace, the pump-around duties, pump-down and recycles to the fractionator in order to control temperatures, flows and levels.
For the economic optimization layer, we use "recipes" based on the feedstock's quality and the needle coke market demand, choosing to maximize either the liquids or coke yield, depending on their economic value.
Building a power-plant which uses undesired coke to supply steam and electricity to the entire refinery is a recently proposed idea (Sawarkar et al., 2007), that should be taken into consideration and in this case, the power-plant's economic optimization should be a part of the overall optimization strategy.
The structure form Figure 3 is initiated in DeltaV PredictPro, where the steady-state gain matrix is loaded and multiple objective function (up to five objective functions are available) are configured.

4. Discussions and conclusions

From an economic point of view, the delayed coking process is a valuable solution to the problem of decreasing residual fuel demand. It also generates a variety of fuels and in some cases a considerable amount of high quality coke (needle coke), while eliminating environmentally unfriendly streams that often involve a disposal cost.
Implementing advanced process control and optimization on a coking plant is quite a difficult task but the results could be remarkable: energy savings, maximized throughput, decreased CO emissions and improved yields while increasing the overall profit of the refinery. Our DeltaV system is able to manage the economic optimization of this process described as a "yield matrix" of a maximum of 40x80 variables.

Nomenclature

w_j (kg/kg) - Mass fraction
w_{j_0} (kg/kg) -Mass fraction of ethane
z (m) - Length along coil
t (s) - Scanning rate
S(i,j) - Stoichiometric constant
V (m/s) - Fluid velocity
r_i (kmol/m^3·s) – Reaction rate
$\overline{C_p}$ (kcal/kg·K) – Process gas specific heat
H_i (kcal/kmole) - Heat of reaction
A (m^2) - Transfer tube area
T_w (K) - Refractory wall temperature
T_g (K) - Flue gas temperature
k_{tg} (W/m·K) - Gas thermal conductivity
k_t (W/m·K) - Tube thermal conductivity
σ (W/m^4·K^4) - Boltzmann P (kgf / m^2) - Pressure
ρ (kg/ m^3) - Process gas density
G_3 (kg·m/kgf·s^2) - Dimensional constant
F - Friction factor
A (s^{-1}) - Frequency factor
E (kcal/mol) - Activation energy

Subscripts
0– initial
i – reaction index;
j – component index

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