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Economic Model Predictive Control (EMPC) of an Industrial Diesel Hydroprocessing Plant

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Abstract: Diesel hydroprocessing is a refinery process by which the sulfur impurities are removed by hydrodesulfurization and the main product diesel is obtained by hydrocracking. The industrial Diesel Hydroprocessing Plant considered in this study consists of two hydrodesulfurization reactors and one hydrocracking reactor in series. The feed to the plant is a blend of four different raw material streams which are heavy diesel (HD), light diesel (LD), light vacuum gas oil (LVGO) and imported diesel from another refinery. A two-layer, hierarchical Economic Model Predictive Control (EMPC) structure is proposed to maximize the profit of the plant. The plant-wide profit is maximized by computing the optimal set-points by the upper economic model predictive control layer while these set-points are tracked by the regulatory model predictive controllers in the lower level. Set-point tracking and disturbance rejection performances of the proposed EMPC structure are tested through closed-loop simulations.

Keywords: Economic model predictive control, non-linear control, hierarchical control, real-time optimization, hydrodesulfurization, hydrocracking.

1. INTRODUCTION

The industrial diesel hydro-processing plant (DHP) subject to this study consists of two catalytic hydro-desulfurization (HDS) reactors and a hydro-cracking (HC) reactor in series as shown in Fig.1. The feedstock of the unit is a blend of four streams: HD (straight run heavy diesel), LD (straight run light diesel), LVGO (light vacuum gas oil) and an imported diesel. HD and LD streams are obtained from a crude distillation unit, and LVGO stream is derived from a vacuum distillation unit. These streams are blended in order to obtain a desired T95 value (the temperature at which 95% of the distillate is collected e.g. by ASTM D86 distillation) for the reactor feed. In the first two HDS beds, the organic sulfur impurities are removed. Hydrocracking (HC) occurs in the last bed where heavier hydrocarbons are cracked to lower molecular weight petroleum fractions. Inter-stage cooling by quench hydrogen is used in both reactors to control the bed exit temperatures. Reactor effluent is next fed to the separation unit where the end products naphtha and diesel are obtained.

The industrial diesel hydroprocessing plant operates with varying feedstocks and large throughputs. Also, changing market conditions have significant effects on product specifications. Therefore, slight improvements on process conditions may result in high profits. In the presence of such a dynamic environment, the diesel hydroprocessing plant must be controlled in the most profitable and safe way. The main operational objectives are to maximize the overall profit of the plant and to keep the sulfur content of diesel below 10 ppm. Therefore, the optimum operating conditions have to be calculated using an economic objective function, and a proper control configuration has to be implemented. In this study, a nonlinear, plant-wide, hierarchical EMPC structure is designed. Nonlinear first principles and empirical models for both blending, reaction and separation subsystems have been developed and validated using industrial data in our earlier work (Aydın et al., 2015).

2. EMPC DESIGN

In classical plant-wide real-time optimization and control (RTO), steady-state set-points are determined first, and these set-points are tracked by the regulatory level controllers. However, a closed-loop implementation of dynamic optimization maximizing the profit over a specified time horizon provides better economic performance since it also maximizes the transient profit. Economic Model Predictive Control (EMPC) is such a strategy (Angeli et al. (2012); Amrit et al.(2013); Würth et al. (2009)). EMPC converts the open-loop dynamic optimization into a feedback control strategy by performing it at each sampling time after updating the initial state based on plant measurements. For large scale complex industrial processes, this can be computationally demanding especially when large prediction horizons have to be used to enhance stability and performance. In order to cope with these disadvantages,

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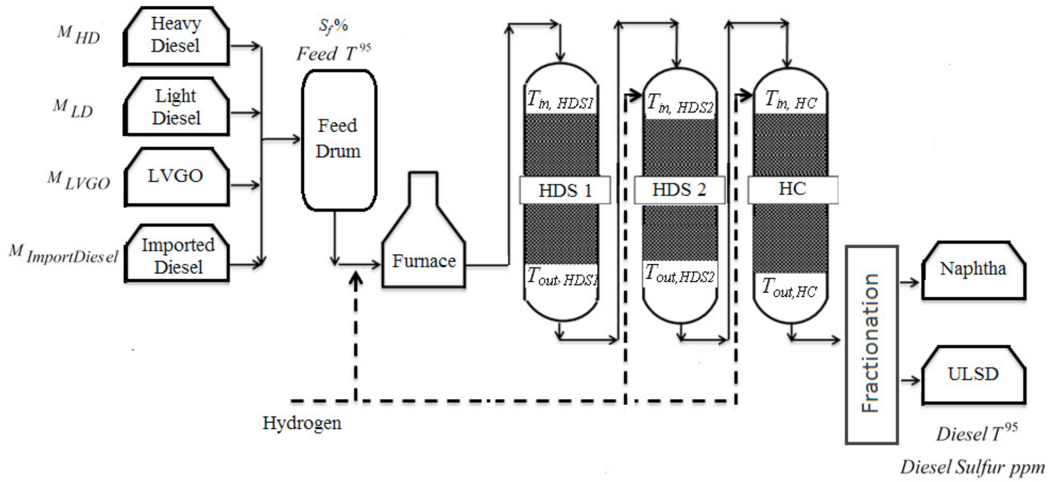

Fig. 1. Simplified DHP flowsheet.

a two-layer real-time implementation of EMPC has been proposed (Helbig et al. (2000); Kadam et al. (2002); Ellis and Christofides (2014)). Sildir et al (2014) has proposed a two-layer hierarchical EMPC strategy for an industrial FCC unit. Here EMPC is designed as a plant-wide controller which supervises the local decentralized MPCs. Similarly, the hierarchical EMPC structure proposed for the industrial DHP plant is shown in Fig. 2. The overall DHP plant consists of Blending, DHP reactors and separator subsystems. Each control layer in the hierarchy has a specific optimizing or regulatory control objective as described next. In this hierarchical approach, EMPC acts as a coordinator by supplying economically optimal time-varying set-point trajectories to the local RMPCs (regulatory MPCs). RMPC follows these trajectories by adjusting their control inputs.

2.1 EMPC Layer

The nonlinear constrained dynamic optimization performed in the EMPC layer is given as follows:

$$\underset{(u_k \;\; k=1,2,..M)}{Max} \{\sum_{k=1}^{N} P_{Diesel}M_{Diesel} + P_{Naphtha}M_{Naphtha} - C_{HD}M_{HD} - C_{LD}M_{LD} - C_{LVGO}M_{LVGO} - C_{ImportDiesel}M_{ImportDiesel}\}_k$$

s.t.

$$x_{k+1} = f_{regulatory}(x_k, u_k), \qquad y_k = g(x_k, u_k) \tag{1}$$

$$T_{out,HC} \le 655\,K, \quad T_{out,HDS1} \le 655\,K, \quad T_{out,HDS2} \le 655\,K$$

$$S_{conv} > 99.7\,\%, \quad 378\,°C \le FeedT^{95} \le 385\,°C,$$

$$350\,°C \le DieselT^{95} \le 360\,°C$$

$$-200 \le \Delta M_{HD} \le 200, \; -150 \le \Delta M_{LD} \le 150, \; 0 = \Delta M_{LVGO},$$

$$-200 \le \Delta M_{Im.Diesel} \le 200, \qquad 400 \le M_{HD} \le 600,$$

$$600 \le M_{LD} \le 800, \qquad 1300 \le M_{Im.Diesel} \le 1500,$$

$$\Delta M_{HD} + \Delta M_{LD} + \Delta M_{Im.Diesel} = 0,$$

$$-7\,K \le \Delta T_{in,HDS} \le 7\,K, \qquad -7\,K \le \Delta T_{in,HC} \le 7\,K$$

Optimization determines the optimal values of the set-points:

$$u_k = ({Feed\,T^{95}}^{sp}{}_k,\, {T_{out,HC}}^{sp}{}_k,\, {T_{out,HDS1}}^{sp}{}_k,\, {T_{out,HDS2}}^{sp}{}_k)$$

which are supplied to the lower layer regulatory MPCs. P_i's are the prices of the products and C_i's are the costs of the raw materials, both of which are set by the refinery management. The functional $f_{regulatory}$ represents the non-linear closed-loop dynamics of the regulatory layer. The refinery management also specifies the daily total flow rate of the unit, which is not allowed to change through daily operation. Since the total feed flow-rate is constant during the period of optimization, utility cost can be assumed as constant. The sampling time of EMPC layer is 100 min, Move horizon M=1 and the length of the prediction horizon N is set to 100 min, which is close to the settling time of the plant.

All the constraints are well defined by the plant management considering equipment, catalyst capacities and safety regulations. LVGO flow rate is not allowed to change. Other raw materials flow rates are subject to the restrictions imposed by the upstream distillation columns. The main product Diesel must have its T^{95} value between 350 and 360 ^0C . Sulfur conversion (*Sconv*) is constrained in order to reach a Diesel ppm level less than 10 ppm.

2.2. Regulatory MPCs

Regulatory MPCs are the decentralized model predictive controllers of the reactors (Regulatory Reactors MPC) and the blending unit (Regulatory Blending MPC). Sampling time of these layers is 6 seconds which is much smaller than the sampling time of the optimization layer to be able to reject