# DYNAMIC REAL-TIME OPTIMIZATION OF A FCC CONVERTER UNIT

**Euclides Almeida Neto[1], Argimiro R. Secchi[2]**

*Grupo de Modelagem, Simulação, Controle e Otimização de Processos (GIMSCOP)*
*Departamento de Engenharia Química – Universidade Federal do Rio Grande do Sul*
*Rua Sarmento Leite, 288/24 – CEP: 90050-170 – Porto Alegre –RS – Brazil*
*Phone: +55-51-3316-3528 – Fax: +55-51-3316-3277*
*E-mail: { [1]eanet, [2]arge}@enq.ufrgs.br*

Abstract: Fluidized-bed Catalytic Cracking (FCC) is a process subject to frequent variations in the operating conditions and changes in the feed quality and feed rate, due to the attempts to maximize LPG and gasoline. This fact makes the FCC converter unit an excellent opportunity for real-time optimization. The present work aims to apply a dynamic real-time optimization (D-RTO) into a simulation of an industrial FCC converter unit, using a mechanistic dynamic model. The algorithms that solve D-RTO problems need to deal with large-scale problems due to the full or partial system discretization along the optimal trajectory. In this work a simultaneous approach, present in the IPOPT solver, was used to discretize the system and solve the resulting large-scale NLP problem.
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## 1. INTRODUCTION

The Fluidized-bed Catalytic Cracking Unit (FCC) is one of the most profitable process units of a petroleum refinery. The FCC converter is part of the reaction section of the unit, where it transforms the low-value raw-materials into commercial products of high-aggregated value.

The FCC converter is a flexible equipment, where the operating conditions can be adjusted to obtain higher yields of LPG (liquefied petroleum gas). When the price of the gasoline is favorable, it can be adjusted to maximize the yield of cracked naphtha, whereas LCO (light oil of recycle) is maximized when the "spread" is favorable to the diesel production. Due to the high profitability of this unit, it should be used its maximum capacity, operating at its maximum feed rate, pushing the big machines, as the gas compressor and air blower, to their upper limit.

Frequent operating-points transitions occur in the converter due to variations in the feed quality, such as variations in the raw-material quality or blends of different streams (coke of gasoil, naphtha, or atmospheric residue) to compose the feed. Frequent changes also happen in the production planning, moving LCO for gasoline or gasoline for LPG, in order to maximize the profitability of the supply chain of the refinery. This process unit is also subject to disturbances in the environmental conditions and limitations of equipments capacity in other process areas.

These facts suggest the use of dynamic real-time optimization (D-RTO) of this system trying to find interesting solutions to optimize the unit, subject to frequent changes in the process operating conditions and production planning. This study seeks to analyze the benefits and limitations of applying dynamic optimization to address this kind of problem. Besides, the critical factors of success of the use of D-RTO should be evidenced to obtain the whole financial potential of this process unit.

Control and optimization of FCC converters has been subject of many studies. The optimization of these processes has been made through MPC's (Odloak *et al.* 1995) and steady-state RTO's (Chitnis and Corropio 1998; Zanin *et al.*, 2000a). NMPC has also been applied (Ali and Elnashaie, 1997) and other strategies of RTO, as optimization in the same layer of advanced control (Odloak *et al.*, 2002; Gouvêa and Odloak, 1998). Zanin *et al.* (2000b) made a comparative study of the use of different optimization strategies in FCC converters. Recently Kadam *et al.* (2005) have been studying dynamic optimization using as example an FCC unit.

## 2. PROCESS MODEL

The FCC conversion area is composed by a furnace for feed pre-heating, a system of reactor-regenerator, air blower, main fractionating tower, and gas compressors. The cracking process constitutes of breaking heavy molecules in a high-temperature tubular reactor, producing fuel gas, LPG, cracked naphtha (gasoline), LCO, decanted oil and coke. During the reaction, deposit of coke occurs in the catalyst surface causing its deactivation and, therefore, its regeneration is mandatory, making part of the process. During the regeneration process there is a heat recovery used to heat the feed up to the cracking reaction temperature.

The FCC converter model used in this work, and developed by Secchi *et al*. (2001), is constituted of the following parts: riser model, separator model, gas compressor model, regenerator model, and valves and controllers models. These models describes a FCC UOP stacked converter, Figure 1, used by PETROBRAS in the Alberto Pasqualini refinery (REFAP S/A). The model was adjusted to the operating conditions of this process unit, describing reasonable well its dynamic behavior.

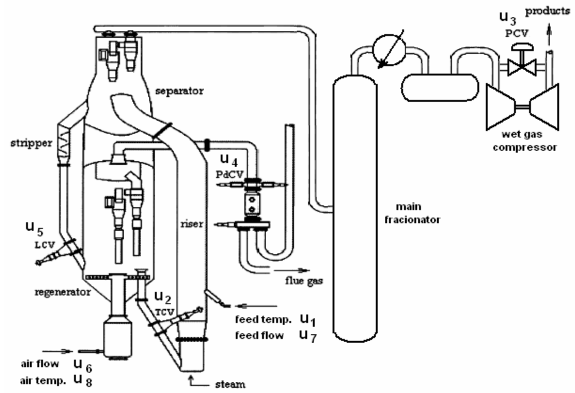


Fig. 1 FCC UOP Stacked Converter (Secchi *et al.*, 2001).

### 2.1 Riser Model

The Riser is modeled as an adiabatic plug flow reactor, with the kinetics described by the ten lumps model of Jacob *et al*. (1976), and using catalyst deactivation and coke formation tendency functions.

The dynamic model of the riser is represented by the mass balance of each lump and coke, using the reaction kinetics of formation of each species, and the energy balance. The resulting partial differential equation was discretized using backward finite-difference technique, with a log-scale non-uniform mesh. A mesh of 20 points was shown satisfactory.

### 2.2 Separator Model

The separator is assumed to be a continuous stirred tank, where catalyst and vapor products (hydrocarbons) are separated. The model of this equipment, based on mass and energy balances, focuses on the prediction of the catalyst level in the separator, the coke content in the spent catalyst, and the catalyst temperature in the separator. The pressure dynamics in the separator is established by a momentum balance.

### 2.3 Gas Compressor Model

The gas compressor is modeled as a single stage centrifugal compressor, driven by a constant speed. The polytropic flow model predicts the suction pressure of the compressor that establishes the pressure in the main fractionating tower and in the separator. There is a recycle stream around the compressor to control the suction pressure, and the mass balance is given by assumed dynamics.

### 2.4 Regenerator Model

The catalyst regeneration is carried out by burning the coke in the catalyst in a fluidized-bed reactor. The fluidized bed is modeled as emulsion and bubble phases that exchange mass and heat. The bubble phase is assumed to be at the pseudo steady-state condition. The disengagement section is modeled as two serial continuous well-mixed tank reactors, corresponding to the diluted and flue gas phases, according to the Figure 2.

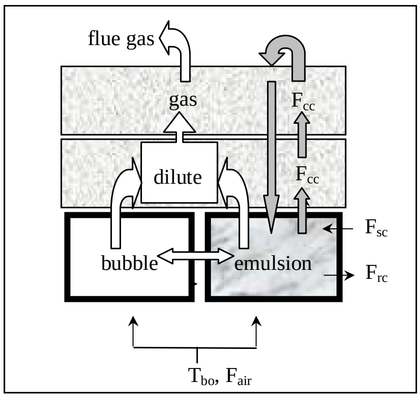


Fig. 2 Regenerator phases (Secchi *et al.*, 2001).

In the regeneration kinetics is used the assumption that the combustion reactions of coke occur in the emulsion, diluted, and gas phases. Component mass balances for $O_2$, CO, $CO_2$, $H_2O$, and coke describe the dynamic behavior of these reactions, resulting in five state equations for each phase of the regenerator. The catalyst inventory in the regenerator is modeled by the overall mass balance for catalyst. The pressure change behavior in the regenerator is obtained through the global mass balance in the gas phase. To predict the dynamic behavior of the temperatures in the regenerator, energy balance was applied in each phase. Considering that the catalyst loss in the regenerator is negligible, the whole catalyst that enters in the regenerator is accumulated or sent to the riser. The coke content in this catalyst is burned mainly in the emulsion phase, but it also suffers reaction in the diluted and gas phases.

*2.5 Valves and Controllers Models*

This type of FCC converter has four degrees of freedoms to provide stability to the system. These degrees of freedom are eliminated by placing regulatory PI controllers in their respective positions:

- Compressor suction pressure controller, using a control valve (PCV) in the compressor recycle stream;
- Reaction temperature controller, using a control valve (TCV) in the stand-pipe to the riser;
- Pressure drop controller between the reactor and the regenerator, using a control valve (PdCV) in the hole chamber of the combustion gases of the regenerator;
- Catalyst level controller in the separator, using a control valve (LCV) in the stand-pipe to the regenerator;

The dynamics of the valves openings are determined by their respective time constants. Additionally, each PI controller has one state equation to describe the integral action. The reaction temperature control will be done by the dynamic optimizer, through a supervisory action directly on the slide-valve. Therefore, only three PI controllers were used.

*2.6 Empirical Correlations for Product Yields*

The FCC converter model does not supply directly all the outputs of interest to analyze and optimize the process. Usually the predictions of products yield and conversion are important to carry out these studies. Empiric correlations were used to supply such desired information. In this case, the volumetric conversion and the yields of fuel gas, LPG, gasoline (GLN), light oil of recycle (LCO), decanted oil (OCLA) and coke (CK).

## 3. FORMULATION OF DYNAMIC OPTIMIZATION PROBLEM

The dynamic optimization problem of a process has the following general form:

$$\min_{u(t)} \varphi(z(t), y(t), u(t)) \tag{1}$$

subject to:

Dynamic Model (EDO):

$$F\left(\frac{dz(t)}{dt}, z(t), y(t), u(t)\right) = 0 \tag{2}$$

Algebraic Equations (EA):

$$G\left(z(t), y(t), u(t)\right) = 0 \tag{3}$$

Initial Conditions:

$$z(0) = z^0 \tag{4}$$

Bounds:

$$\begin{aligned} z^L &\le z(t) \le z^U \\ y^L &\le y(t) \le y^U \\ u^L &\le u(t) \le u^U \end{aligned} \tag{5}$$

*3.1 Solution of Dynamic Optimization Problem*

The infinite dimension dynamic optimization problem can be solved through variational methods, using the Pontryagin's maximum principle and solving the resultant two-point boundary value problem (TPBVP), or approximating to a finite formulation, with predefined functional forms for the control variables. In this last case, the resultant NLP problem can be solved by sequential or simultaneous approaches. In the sequential approach only the control variables are discretized or parameterized, while in the simultaneous approach the whole system is discretized in the time domain, usually using orthogonal collocation techniques. See the work of Biegler *et al*. (2002) for a deeper revision on these methods.

In this work the simultaneous strategy has been used, where the continuous problem is converted in a nonlinear programming problem (NLP) when approximating the state and control profiles by a family of orthogonal polynomials on finite elements (Cervantes, 1998). For first order differential equations the approximation results in:

$$z(t) = z_{i-1} + h\sum_{q=1}^{ncol} \Omega_q\left(\frac{t - t_{i-1}}{h_i}\right)\frac{dz}{dt_{i,q}} \tag{6}$$

The control profiles and algebraic equations are approximated in a similar way and the equation takes the following form:

$$y(t) = \sum_{q=1}^{ncol} \Psi_q\left(\frac{t - t_{i-1}}{h_i}\right) y_{i,q} \tag{7}$$

$$u(t) = \sum_{q=1}^{ncol} \Psi_q\left(\frac{t - t_{i-1}}{h_i}\right) u_{i,q} \tag{8}$$

Usually, in dynamic optimization problems the number of control variables is small and the number of state variables is very large. In this case the rSQP algorithm (reduced SQP) is efficient (Waanders *et al*., 2002). The solution of these problems is also efficient using the interior point algorithms, however they require improvements, and many of them have been proposed. The following ones can be highlighted: the use of the preconditioned conjugated gradient method (PCG) to update the control variables (Cervantes and Biegler, 2001), and the introduction of a filter in the strategy of the line search, where the objective function compete with the infeasibility of the problem (Wächter, 2002). In the interior point algorithm the original NLP problem can be written as:

$$\begin{aligned} \min \quad & f(x) \\ s.t. \quad & c(x) = 0 \\ & x \ge 0 \end{aligned} \tag{9}$$

The barrier function is added to reduce the dimension of the problem, and then the problem takes the form:

$$\min \quad \varphi_\mu(x) = f(x) - \mu \sum_{i=1}^{n} \ln(x^i) \tag{10}$$
$$s.t. \quad c(x) = 0$$

All of these features were implemented in the IPOPT algorithm developed by Carnegie Mellon University (CAPD Report, 2003; Lang and Biegler, 2005).

### *3.2 Configuration of the Objective Function*

In the optimization of FCC converters there are some concurrent production objectives. The maximization of the operational profit is a common objective; however there are moments where some specific product needs to be maximized. This is due to the optimization of the refinery supply chain. There are situations where the local optimum of an isolated unit of process is not the global optimum of the supply chain. In order to attend these situations, multiple objectives can be adopted. In this dynamic optimization problem, the integral of different factors along a day ($t_f$ = 24 h) were maximized. The most common situations are the following ones:

*Maximization of the operational profit:*

This is the most common objective function; however it is more difficult for the operators to analyze the results from the optimizer.

$$Profit = Revenue - Costs$$

$$Revenue = m_{FG}\,\Pr_{FG} + m_{CK}\,\Pr_{CK} + V_{LPG}\,\Pr_{LPG} + V_{GLN}\,\Pr_{GLN} + V_{LCO}\,\Pr_{LCO} + V_{OCLA}\,\Pr_{OCLA} \tag{11}$$

$$Costs = V_{Feed}\,\Pr_{Feed} + m_{Cat}\,\Pr_{Cat} + Q_{PreH}\,\Pr_{Q} + Q_{Proc}\,\Pr_{Fuel} + Pot_{Blwr}C_{Blwr} + Pot_{Compr}C_{Compr} \tag{12}$$

$$FObj_1 = -\int_0^{t_f} Profit.dt \tag{13}$$

*Maximization of the total conversion:*

The maximization of the total conversion leads to the use of the maximum capacity of the converter, breaking the molecules into more important products. The disadvantage of this approach is that the conversion does not focus in highest price products. The average conversion is calculated in the following form:

$$FObj_2 = -\frac{\int_0^{t_f} Conv_v V_{Feed}.dt}{\int_0^{t_f} V_{Feed}.dt} \tag{14}$$

*Maximization of LPG production:*

This objective is adopted when there is a clear advantage in the maximum conversion in LPG product.

$$FObj_3 = -\int_0^{t_f} \frac{\eta_{LPG}}{100} V_{Feed}.dt \tag{15}$$

*Maximization of gasoline production (GLN):*

This objective is adopted when there is a clear advantage in the maximum conversion in gasoline.

$$FObj_4 = -\int_0^{t_f} \frac{\eta_{GLN}}{100} V_{Feed}.dt \tag{16}$$

*Maximization of LCO production:*

The maximization of production of LCO is adopted when there is a clear advantage in the maximum conversion in LCO. In this case the LCO is an intermediary product and it is incorporated to the diesel pool. When LCO is a diluent, there is only interest in maximize it when displace some part of kerosene from the diluent pool to the jet fuel pool.

$$FObj_5 = -\int_0^{t_f} \frac{\eta_{LCO}}{100} V_{Feed}.dt \tag{17}$$

The specific productions objectives are mutually exclusive. When you need to maximize the production of a specific stream, the weights of the other objectives should be zero.

*Objective function formulation:*

In general case each production objective can be represented in the following way:

$$FObj_i = -\int_0^{t_f} OBJ_i.dt \tag{18}$$

The multi-objectives problem can be written as a weighted sum of each specific objective:

$$\varphi = \sum_{i=1}^{n} k_i FObj_i \tag{19}$$

The constraint based multi-objectives strategies (ε-constrained and goal attainment) will be studied in future works, and was partially adopted here.

The integral in each specific objective is manipulated by differentiating the original objective function and creating a new state $\varphi$ added to the set of differential equations. Therefore, the objective function assumes the following form:

$$\frac{d\varphi}{dt} = -\sum_{i=1}^{n} k_i OBJ_i \tag{20}$$

### *3.3 Additional Constraints*

Besides the constraints usually imposed to the states and the control variables, supplementary constraints were added that represent bounds in the production objectives to guarantee the feasibility of the solution in the optimization problem. The additional constraints are the following ones:

Minimum daily profit:

$$Profit \geq Profit_{\min} \tag{21}$$

Minimum conversion in the riser:

$$Conv_v \geq Conv_{\min} \tag{22}$$

Minimum and maximum LPG production:

$$V_{LPG}^{\min} \leq \frac{\eta_{LPG}}{100} V_{Feed} \leq V_{LPG}^{\max} \tag{23}$$

Minimum and maximum GLN production:

$$V_{GLN}^{\min} \leq \frac{\eta_{GLN}}{100} V_{Feed} \leq V_{GLN}^{\max} \tag{24}$$

Minimum and maximum LCO production:

$$V_{LCO}^{\min} \leq \frac{\eta_{LCO}}{100} V_{Feed} \leq V_{LCO}^{\max} \tag{25}$$

## 4. CASE STUDIES AND RESULTS

The dynamic optimization problem has been solved applying the IPOPT algorithm through the software of dynamic optimization DynoPC developed by Carnegie Mellon Univ. (Lang and Biegler, 2005). The several alternatives of production objectives studied in this work are presented in Table 1.

Table 1. Case studies.

| ***Case*** | ***Production Objective*** |
|---|---|
| 1 | Maximum Profit |
| 2 | Maximum Feed Rate |
| 3 | Maximum Conversion |
| 4 | Maximum LPG Production |
| 5 | Maximum Gasoline (GLN) Production |
| 6 | Maximum LCO Production |
| 7 | Maximum Profit with Max. Conversion |
| 8 | Maximum Profit with Max. Feed Rate |
| 9 | Maximum Profit with Max. LPG Prod. |
| 10 | Maximum Profit with Max. GLN Prod. |
| 11 | Maximum Profit with Max. LCO Prod. |

The maximization of the profit is the more common production objective, and it is usual to associate it to a specific objective, as maximum conversion or some product that the scheduling people defines as priority. This prioritization can also be made putting bounds in secondary objectives, for example, the minimum conversion (ε-constrained approach).

### 4.1 *Dimension of the Optimization Problem*

As the definition of the problem described above, the number of variables involved in the problem are given in Table 2.

Table 2. Number of variables in the formulation.

| | |
|---|---|
| Number of differential variables (nz) | 274 |
| Number of algebraic variables (ny) | 21 |
| Number of control variables (nu) | 8 |
| Number of finites elements (ne) | 40 |
| Number of collocation points/element (ncol) | 3 |
| Total number of discretized variables | 47642 |
| Total number of constraints | 47594 |
| Total number of lower bounds | 14440 |
| Total number of upper bounds | 14440 |

### 4.2 *Case Studies*

Due to space limitation, the obtained results are analyzed for the maximization feed rate case. The optimization problem was solved in an Intel Pentium IV, 2.8 MHz computer and spent 30 - 45 min of CPU time. This time consumption is compatible with the interval per control action (around 4 to 6 per day).

To obtain accurate and numerically stable results, it was necessary to tune the discretization parameters as the number of collocation points and finite elements. Also, in order to reduce the number of control actions some finite elements were grouped (Lang and Biegler, 2005). This procedure provided a more robust solution of the optimization problem and with a better performance.

*Case 2 – Maximum Feed Rate*

The maximization of the feed rate is prioritized when it is necessary to use the total capacity of the process unit. In this case, it is reached the limit of catalyst circulation or the limit of capacity of a main machine (air blower or gas compressor). Notice that the optimizer increased the feed flow rate, opened the catalyst valve to the maximum, dropped the suction pressure of the gas compressor, and reduced the pressure drop between the reactor and regenerator (Figs. 3 to 6). In order to supply the additional energy demanded by the system, the regenerator and riser temperatures were increased (Figs. 7 and 8).

As result of the dynamic optimization, the profit operation was increased by the order of 5.5 thousand dollars a day ($0.20/bbl). It is the normal potential of benefit with the advanced control and RTO applications (Fig. 9). It also can be observed that there was an increase of volumetric conversion (Fig. 10) and yields of gasoline (GLN) and LCO and a reduction in the decanted oil yield (OCLA), which is a less valuable product (Fig. 11).

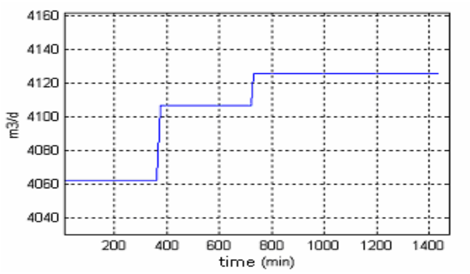

Fig. 3 Feed flow rate ($u_1$).

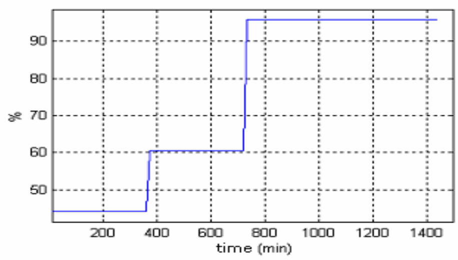

Fig. 4 TCV control signal ($u_2$).

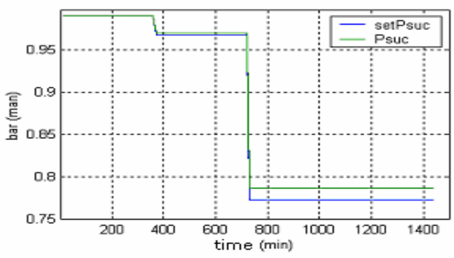


Fig. 5 Suction pressure of gas compressor ($u_3$).

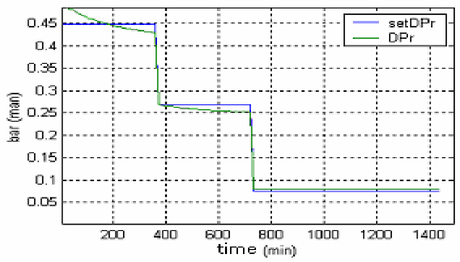


Fig. 6 Differential pressure reactor/ regenerator ($u_4$).

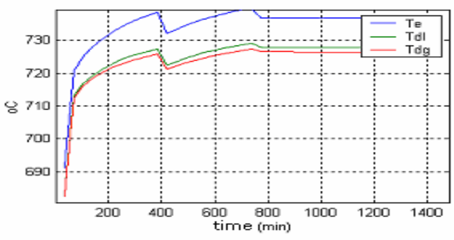


Fig. 7 Regenerator's temperatures.

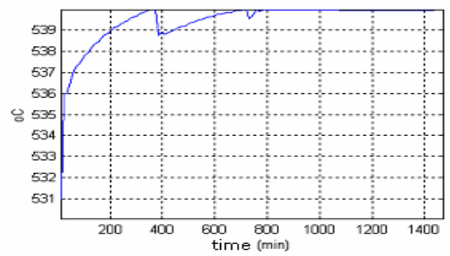


Fig. 8 Riser temperature (reaction).

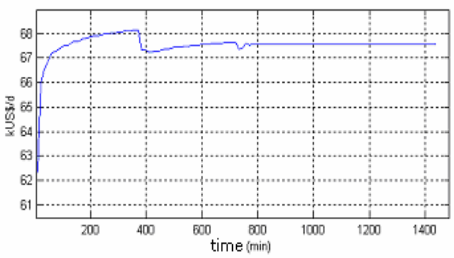


Fig. 9 Operation profit.

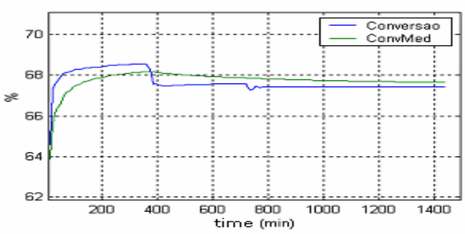


Fig. 10 Volumetric conversion.

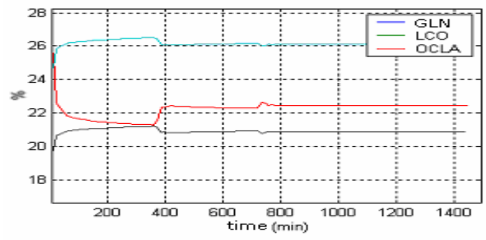


Fig. 11 Volumetric yields.

## 5. CONCLUSIONS

The dynamic optimization of the FCC converter has obtained coherent results with the expected in an industrial unit. The results demonstrate that the application of D-RTO in this kind of unit can bring significant benefits. The simultaneous approach has shown to be effective for the solution of the problem, but it demanded a lot of time to tune the discretization parameters of the control variables. The strategy of grouping intervals for the control variables was the one that presented better performance. Due to space limitation, the other analyzed cases will be presented in the symposium.

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