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Improvements on Model Predictive Control for a Pulp Mill Process

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Abstract: This work constitutes a contribution to the previous one presented by Castro and Doyle (2004a). They decided the incorporation of four Model Predictive Control (MPC) for specific parts of the complex chemical Pulp and Paper plant to improve its global dynamic and economic performance. Meanly the authors supported the decision of including MPC based on the RGA information. In this paper, a deep analysis about each MPC implementation is performed so as to test if the used methodology could guide efficiently for adopting this kind of decisions. Initially, the study begins with a systematic procedure for adjusting the key MPC tuning parameters. The economic and dynamic performance indexes are evaluated to demonstrate for which specific cases a real benefit can be achieved. The results presented here were obtained through dynamic simulations using the computational benchmark model of 8200 states for the same scenarios evaluated by Castro and Doyle (2004b).

Keywords: pulp and paper industry, model predictive control, parameter adjustment, performance evaluation, economic evaluation, dynamic simulation.

1. INTRODUCTION

Model Predictive Control (MPC) is a control strategy for industrial use in different types of processes. In particular, chemical industry is one of the sectors where its implementation has been highly successful since its inception. Nevertheless, there are edges that represent an open problem to dedicate a space where generate satisfactory answers. MPC is based on the explicit use of an internal mathematical model of the process. Itl is utilized to predict the evolution of controlled variables over a prediction horizon. MPC actions respond to the optimization of a cost function which is related to the future behaviour of the system, predicted through the dynamic model. A complete review of MPC can be found in Maciejowski (2000).

In the present work, a deep analysis about the adopted criteria given by Castro and Doyle (2004a) for deciding the MPC incorporation is a proper technique or not. It was thought for some specific units of the large-scale chemical Pulp and Paper process. The objective was to obtain overall dynamic and economic advantages. When it is appropriate, MPC controller replaces the Decentralized Control (DEC) structure originally implemented at the supervisory control level.

The DEC's dynamic and economic capabilities which justify its replacement are compared against the new control structure through the calculation of specific indexes. Since the results are sensibly different according to the tuning parameters adopted for the MPC in this work represents one of the most important topics to be focused. Hence a proper MPC tuning parameters based on a sensibility analysis of key variables is performed. A rigorous model of the process allows support the complete analysis.

The control problem addressed here starts with the same structure given by Castro and Doyle (2004b). The study presented here tries to demonstrate that several considerations given there could drive to erroneous decisions. The conclusions obtained here encourage the development of a systematic procedure for determining a new plant-wide control strategy.

The process presents typical characteristics of large chemical plants like important time delays, multiple-loop interactions and rich dynamic variables. It consists of several operating units, resulting in a large number of controlled outputs (114 CVs), manipulated variables (82 MVs) and disturbances (58 DVs).

The whole process is implemented in Matlab 6. It consists of Simulink models, s-functions and numerous scripts (m-files and c-files) including configuration scripts for simulations given by Castro and Doyle (2004a).

In Section 2 a brief description of the Kraft process, its main objectives and restrictions are presented. Section 3 presents the implemented control architectures in both the lower and upper control layers. The various tools that allow a quantification of the benefits are discussed in Section 4. Section 5 presents the performed closed-loop simulations for the complete model, according to the various control configurations used. Finally, the results, conclusions and future works are exposed.

2. PROCESS DESCRIPTION

2.1 Kraft Process

The Kraft Pulp Mill process is shown in Figure 1 in a simplified way as that given by Castro and Doyle (2004a). The plant consists of two major areas such as the Fiberline and the Recovery Plant.

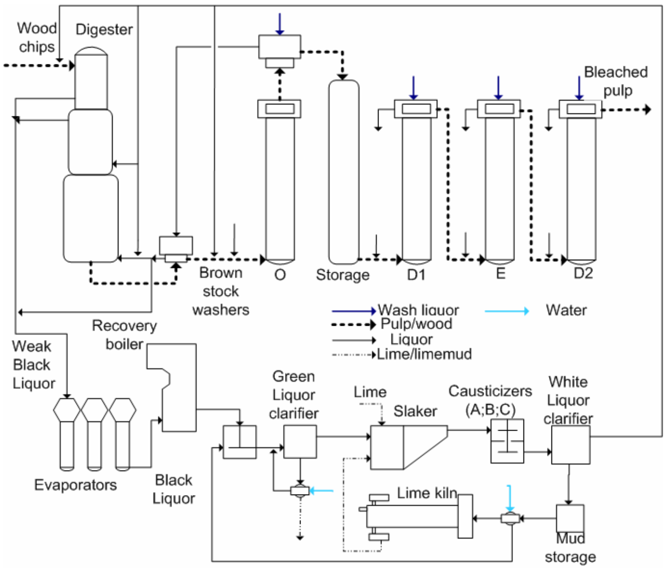

Fig. 1. The Kraft Process.

The objective of the Fiberline area is to produce fibers from wood chips at a desired production rate and quality. Major raw materials of this process are wood chips and chemicals called white liquor (WL) which consists primarily of NaOH and NaSH. They are combined in a pressurized impregnation vessel, where wood chips are saturated in the white liquor. They then enter the Digester, where lignin in the wood starts to dissolve out. The main controlled variable in this unit is the Kappa Number, which is a measure of the amount of remaining lignin in the wood. Fibers are further washed in a Brown Stock drum washing section, to remove chemicals and residual lignin. Fibers then are bleached in several Bleaching Towers, to further remove lignin and achieve a target brightness coefficient. The bleaching sequence includes post-delignification with Oxygen (O) and white liquor, Chlorine Dioxide (D1), Sodium Hydroxide (E), and brightening via Chlorine Dioxide (D2). At the end of each bleaching stage, pulp is washed to removed chemicals and lignin content before going to the next bleaching stage. On the other hand, the exit of the chemical streams from the Digester and washing stages now has many organic residuals and brown colour and, hence, are called ”Weak Black Liquors” (WBL). To recover chemical components and energy from these streams, the weak black liquors are sent to the Recovery Plant.

The most important objectives of the Chemical Recovery area are to obtain energy from the combustion of black liquor and regenerate the NaOH and Na2S from the weak black liquor coming from the Digester, extract liquor flows and the brown stock washing system. This regeneration procedure becomes the overall process economically feasible. In the multi-effect evaporation system the weak black liquor is converted in black liquor. The black liquor is sent to the recovery boiler where the combustion of the organic liquor provides the energy to produce high pressure steam and to carry out the reduction reactions to recover Na2S from Na2SO4 and other sulfur-based salts, and to recover Na2CO3. The black liquor solids is mixed with weak wash water to produce green liquor. The green liquor goes through the slaking/causticizing reactions to produce white liquor. This white liquor is sent back to the digester and the oxygen reactor. The lime mud from the white liquor clarifier is sent to the lime kiln to recover the lime.

2.2 Main Objectives

The control strategy should consider the following points:

a) Minimize the error of controlled variables (difference between its value and setpoint), respecting the following constraints:

- Produce pulp with a defined production rate and quality (brightness and Kappa No.) in the Fiberline. The brightness of the pulp should not exceed ±1% of its nominal value (quality constraint).

- The percentage of Black Liquor solids in the Evaporator should be maintained between 60-70%. This ensures proper combustion of the liquor without causing excessive heating of the evaporation system.

- The oxygen concentration in the kiln should be at least 1% (environmental constraint).

The manipulated variables must also abide its restrictions.

b) Reject disturbances as quickly as possible.

c) Minimize the use of wood chips, chemicals and other raw materials, and energy in general, in order to minimize existing costs.

3. IMPLEMENTED CONTROL ARCHITECTURES

3.1 Lower control layer

Based on the study of the multiple variables interaction by Relative Gain Array (RGA) analysis and assumptions from the engineering criterion, Castro and Doyle (2004b) established the input-output pairs for Decentralized control of the secondary variables (which are related to pressures, levels, flows, temperatures, etc.). To design the loops, the

CV-MV matching were chosen so that the RGA matrix results as diagonal as possible. This means decoupling the system, which is the purpose of Decentralized control.

3.2 Upper control layer

Two configurations are analyzed through simulations to control the primary variables (which are related to quality, production, safety, environment and operation of the process) and the economic variables (they are chosen in order to improve cost-benefit ratio of the process):

a) Decentralized Control (DEC): Consists of conventional PI loops and cascade loops. Some of them contain special controls to improve the dynamics of the primary controls: 3 control loops include Ratio Control, 4 control loops include Factor Kappa Control, 2 control loops include Smith Predictors and 1 control loop includes Feedforward Control.

b) Model Predictive Control (MPC): In this case the problem is approached from a more wide point of view. All primary controlled variable measurements and measured disturbances enter to the controller, and the manipulated variables are calculated. This is a strategy that monitors the main controlled variables in order to comply with certain objectives and constraints of the process. These objectives and restrictions determine the limits on controlled and manipulated variables, and also the setpoints for the primary variables. Further details of the MPC control can be found in Maciejowski (2000).

The MPC strategy developed to control the process consists of four MPC controllers. This partition was realized taking into account the RGA analysis in the portions in which the degree of interaction was very significant. Castro and Doyle (2004b) proposed to distribute the MPC in the following areas:

- MPC 1: Digestor and Oxygen Tower

- MPC 2: Bleaching Plant

- MPC 3: Evaporators

- MPC 4: Lime Kiln and Recaustificators

Every MPC action can directly enter to the process as a manipulated variable, or represents a setpoint associated with a PI controller or a Kappa controller.

The corresponding control algorithm is based on state-space interpretation of MPC: a state-space model expressed in terms of step response parameters is used. A complete review of this can be found in Lee et al (1994).

4. QUANTIFICATION OF BENEFITS

As tools to quantify and compare the capabilities of each of the control strategies, EIP and PIP indexes were mainly used. The first one reflects the dynamic performance, whereas the second one concerns the economic benefits. These indexes are based on the time evolutions of the dynamics of the controlled and the manipulated variables. They are defined as follows:

- EIP ”Error Improvement Percent”:

$$EIP = 100\,\frac{IAE^{base} - IAE^{new}}{IAE^{base}} \qquad (1)$$

While greater (and positive) is the calculated value of EIP for a variable, then smaller is the error between the output value of the variable and its desired value. Which translates into better dynamic performance for the new control structure employed. The superscript ”base” refers to the decentralized control strategy and the superscript ”new” refers to the proposed new control strategy for the system.

- IAE ”Integral Absolute Error”:

$$IAE = \sum_{k} \left| k(k) - y(k) \right| \qquad (2)$$

Where ”y” is the output and ”r” corresponds to the desired value.

- PIP ”Profit Improvement Percent”:

$$PIP = 100\,\frac{TOP^{new} - TOP^{base}}{\left|TOP^{base}\right|} \qquad (3)$$

The PIP reveals how much the economic benefits were increased by the utilization of the new control strategy with respect to the one used as base. Therefore, while greater is its value for a particular area of the process, then the operation is more profitable when it is controlled with the new structure.

- TOP ”Total Operating Profit”:

$$TOP = Sales - Penalty\ \ Costs - Raw\ \ Costs \qquad (4)$$

It is the economic benefit calculated for each unit of the process:

-”Raw Costs”: costs of raw materials used

-”Penalty Costs”: costs relative to penalties from environmental regulations or violations of product quality

- ”Sales”: proportional value to the production of the area

Similarly, the above values are also calculated for the entire process.

5. PERFORMANCE OF CONTROL CONFIGURATIONS. METHODOLOGY.

For evaluate and compare the performance of the different control configurations at the supervisory level (upper control layer), the following set of simulations were performed. For all the simulations, lower control layer is implemented with DEC strategy.

- Simulation 1: upper control layer implemented with DEC strategy.

- Simulation 2: upper control layer corresponding to Digestor unit controlled with MPC1, and DEC strategy for the remaining units.

- Simulation 3: upper control layer corresponding to Bleaching Plant controlled with MPC2, and DEC strategy for the remaining units.

- Simulation 4: upper control layer corresponding to Evaporators unit controlled with MPC3, and DEC strategy for the remaining units.

- Simulation 5: upper control layer corresponding to Lime Kiln unit controlled with MPC4, and DEC strategy for the remaining units.

- Simulation 6: upper control layer corresponding to Fiberline area controlled with MPC (MPC1+MPC2), and DEC strategy for the remaining units.

- Simulation 7: upper control layer corresponding to Chemical Recovery area controlled with MPC (MPC3+MPC4), and DEC strategy for the remaining units.

- Simulation 8: upper control layer corresponding to Digestor and Lime Kiln units controlled with MPC1 and MPC4 respectively, and DEC strategy for the remaining units.

- Simulation 9: same conditions as Simulation 8 with the addition of the adjustment of MPC1 parameters.

- Simulation 10: upper control layer completely controlled with MPC (MPC1+ MPC2+ MPC3+ MPC4)

The simulations are based on applying a typical sequence of inputs consisting of setpoint changes and disturbances affecting the full process model, as proposed by Castro and Doyle (2004b). Table 1 shows the sequence of setpoint changes and disturbances applied.

Based on the obtained time evolutions of the CVs and MVs, EIP indexes for primary variables and PIP indexes for the process units were calculated.

6. RESULTS

Within the set of simulations, the most profitable results with one or more MPC implemented are presented, confronted to the DEC strategy implemented in the whole upper control layer. Different combination of control structures allowed to obtain good performance for each of the typical process units. Subsequently, the search was focused to the best strategy for the entire process, considering the results already found.

In the Fiberline area, the best dynamic and economic performance was obtained when upper control layer corresponding to Digestor unit was controlled with MPC1, and DEC strategy for the remaining units.

The greatest error reduction of the variables of the Chemical Recovery area was obtained when upper control layer corresponding to Lime Kiln unit was controlled with MPC4, and DEC strategy for the remaining units. But it did not show economic benefits.

Table 1. Sequence of Setpoint Changes and Disturbances applied to the Process

Time [hr]	Event	Value (unscaled)
0	O caustic make-up flow change	-1.0
0	Kiln O2 % change	0.015*
0	Kiln CaCO3 % change	-1.0
0	Minimization of O tower steam	-1.0
0	Minimization of fresh lime	-1.0
8.3	Wood chips moisture change	-1.0
25	Lignin density change	1.0
25	Cellulose density change	-1.0
41.6	ClO2 composition change	-0.5
56.6	Kiln O2 % change	0.0125*
58.3	ClO2 composition change	0.0
75	Ambient temperature change	-1.0
76.6	Slaker temperature change	0.25
83.3	Displacement ratio of recov. filters	-1.0
91.6	Bleach pulp production change	756*
100	E Kappa no. Change	-0.83
100.2	E tower temperature change	1.5
101.5	D2 tower temperature change	1.5
101.5	D2 brightness change	5.0
(*): scaled values		

Therefore, the performance of the combination MPC1+MPC4 together with DEC strategy for the remaining units was examined. From the Simulation 8 we obtained a very good dynamic performance for the entire process. This is evidenced through Figures 2 to 5, corresponding to some of the most important primary controlled variables that are related to the restrictions. In Figures 2 to 5, the setpoint is represented with green, the DEC strategy with blue, and the new strategy (MPC1+MPC4) with red.

Figure 2 shows the dynamics of the variable ”D2 Production Rate”, belonging to the Fiberline. With a setpoint change from 630 to 756 tons/day at 91.6 hours, it shows that the variable which adjusts quickly to the new value is generated by the structure with MPC.

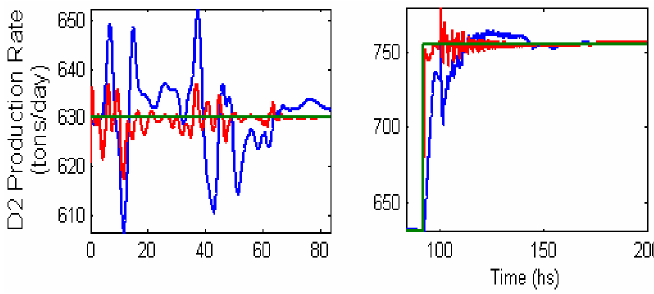

Fig. 2. CV3 – D2 Production Rate.

Figure 3 shows the behavior of the variable ”Percentage of Dissolved Solids in Black Liquor”. It shows that the control action, generated by the implemented MPC, makes the controlled variable goes faster to its setpoint, after a disturbance enters to the system at 91.6 hours.

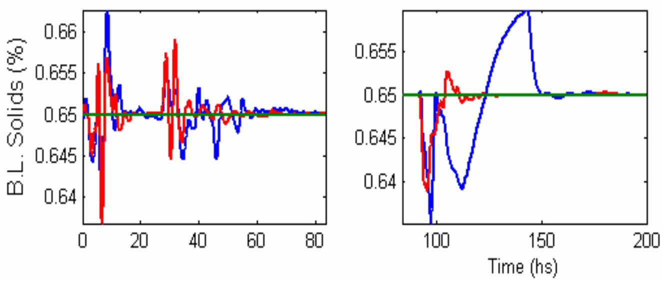

Fig. 3. CV44 – % of Black Liquor Solids.

In Figure 4, the primary controlled variable under study is "Kiln O2 Mass Fraction," which corresponds to the Chemical Recovery area. With two setpoint changes, the first from 0.035 to 0.015 at 0 hour and the second from 0.015 to 0.0125 at 56.6 hours, and with a disturbance at 91.6 hours, the best response is that of the structure MPC1+MPC4.

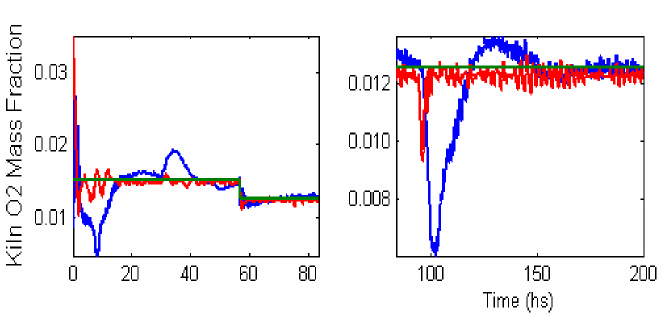

Fig. 4. CV79 – Kiln O2 Mass Fraction.

Figure 5 shows the behavior of the variable "Digester Kappa No." which corresponds to the Fiberline. With the entry of a disturbance at 91.6 hours, a low overvalue takes place and stabilizes faster with the new strategy.

The average EIP for the Chemical Recovery area variables increased from 29% (corresponding to the MPC4 only) to 38.6%. All graphs show that the total error is much smaller with the new strategy. This is reflected in the EIP values shown in Table 2.

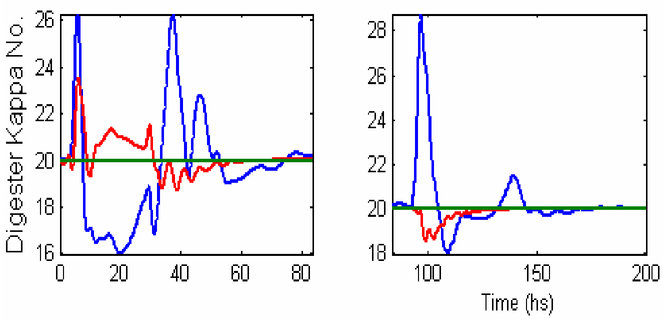

Fig. 5. CV4 – Digester Kappa No.

The disadvantage of this combination is that it does not produce economic benefits with respect to upper control layer implemented with DEC strategy.

Table 2. EIP Indexes Comparison

CV Nº	CV Description	IAE (Dec)	EIP % MPC1+4	EIP % Mpc1+4 (adj.)
3	D2 Production Rate	141.1	74.1	73.3
39	Upper Extract Ef. Alkali (est)	202.8	72.1	35.5
40	Lower Extract Ef. Alkali (est)	67.7	23.2	68.0
4	Digester Kappa no.	453.3	72.2	74.6
19	O Tower Kappa no.	203.4	59.2	57.2
22	E Tower Kappa no.	68.9	24.2	22.4
24	E Washer [OH-]	117.7	57.8	65.7
26	D2 Brightness	507.2	0.4	-1.2
44	% BL Solids	114.5	52.7	66.1
62	Slaker Temperature	12.7	7.6	15.5
79	Kiln O2 Mass Fraction	134.3	61.4	62.0
81	Kiln CaCO3 (est.)	150.3	32.6	35.9

As another principal objective is to obtain economic benefits with regard to the DEC strategy, then the values of the following set of parameters belonging to the MPC1 were readjusted: Qi (parameters belonging to the weight matrix that penalizes errors of the controlled variables with respect to their setpoints) and Ri (parameters belonging to the weight matrix that penalizes the control movements made).

Several preliminary simulations were done for choosing the parameters to adjust. In each simulation only one parameter value (Qi or Ri) were changed. After each simulation the PIP index relative to the whole process was evaluated. The parameters considered were those that appreciably increased the value of PIP = -0,53% originally obtained with MPC1+MPC4. The utilized criterion to define which parameters were finally adjusted, and their values, was to gather those Qi and Ri such that their individual effects of increasing the PIP were enhanced by modifying them together. To sensitively increase the economic benefits, this adjustment of parameters should increase the sales and/or reduce the costs in the process units which have major impact on total costs (Bleach Plant, Digester, and Evaporators). In principle, an error increase in some controlled variables was permitted, in order to obtain greater economic benefits.

Finally a new simulation (Simulation 9) was executed with MPC1+MPC4 and the adjusted parameters of MPC1, with the following favourable results. Table 2 shows the calculated EIP values. Table 3 shows the adjusted MPC1 parameter values.

Table 3. New Values for MPC1 Parameters

Parameter	Original Value	New Value
Q(1)	1.50	0.70
Q(3)	0.75	0.35
Q(4)	0.75	1.20
Q(6)	0.00	0.60
R(1)	30.00	15.00
R(3)	20.00	10.00
R(4)	20.00	30.00
R(6)	10.00	20.00
R(11)	10.00	20.00

By reducing the value of the parameter Q1 (which is the one that penalizes the error in the variable CV3: "D2 Production Rate") there was a corresponding decrease in the EIP. However, when a setpoint change was applied in the variable, it adjusted more quickly to its new value, increasing this way sales in the Bleach Plant.

In the case of the Digester, by increasing the value of two of the Ri parameters (R4 and R6), then the expense of control associated with two of the variables of this zone was decreased. Since these control actions represent temperature setpoints in Digester loops, then a reduction of them implies a reduction of the manipulated variables "steam flow". This reduction of Digester costs justifies the PIP index increase. It changes from -0.21% to -0.12% in that unit.

For the Evaporators area there was a PIP index increase from 2.39% to 2.44%, due to the cost reductions achieved.

An improvement was obtained in the dynamics of the controlled variable "Kiln O2 Mass Fraction" in the Lime Kiln. The restriction violation time of this variable was despicable. Therefore penalty costs disappear in this area, increasing the benefits. Also there was a small reduction in fuel consumption.

The new strategy is successful in improving PIP indexes for the more profitable areas of the process, with regard to the MPC strategy without the adjusting of MPC1 parameters. Table 4 shows the PIP values for both simulations.

The positive value of the PIP index obtained with the MPC1 (adjusted)+MPC4 configuration for the whole process reflects an overall increase of the economic benefits with regard to the DEC strategy.

Table 4. PIP Indexes Comparison

Operation Unit	PIP % MPC1+4	PIP % MPC1(adj.)+4
Digester	-0.21	-0.12
Brown Stock	0.47	0.70
Oxygen Tower	-0.25	-0.53
Bleach Plant	-0.31	0.16
Evaporators	2.39	2.44
Recaust	-1.00	-1.19
Lime Kiln	0.58	0.72
Total	**-0.53**	**0.41**

7. CONCLUSIONS AND FUTURE WORKS

7.1 Conclusions

Remarkable economic improvements were obtained by the implementation of MPC1 (Digester zone), MPC4 (Lime Kiln zone) and DEC in the remaining units, when the values of the MPC1 parameters were adjusted. The criterion of adjustment was based in selecting and handling a limited number of parameters of the MPC1 weight matrices. These parameters are related to the Digester, Bleach Plant and Evaporator economics, which are those of major impact on total costs. Comparing against the DEC strategy, the obtained PIP index for the whole process with the new control configuration was PIP = 0.41%. Although initial hypothesis for the adjustment allowed a reduction of the positive margin of the average EIP index available up to that moment, finally remained its high value after the adjustment, with an average EIP of 48%. Although the "D2 Brightness" controlled variable is the only one that increases its error comparing with the DEC strategy, it resulted PIP = 0.16% for the Bleach Plant.

7.2 Future Works

Starting with the stabilization of the open-loop plant including some level control loops, a simplified dinamic model of the complete process was obtained. These previously tasks represent the stating point of designing and implementing a new plant-wide control structure for the process such as that given by Molina et al (2009). The idea is to get a simpler DEC structure with a fewer number of control loops than those presented previously by Castro and Doyle (2004a), fulfilling the main objectives of keeping the process with a reduced energy consumption and reject the disturbances. This methodology involves obtaining in a systematic way and without using any heuristic concepts, a set of controlled variables minimizing some defined index which accounts both setpoint and disturbance changes. Then, the use of the relative gain array to find the MV-CV pairings. The resolution of this large-scale combinatorial problem will be efficiently done through a proper optimization algorithm.

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