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LP-DMC CONTROL OF A CHEMICAL PLANT WITH INTEGRAL BEHAVIOUR

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Abstract: The purpose of this paper is to describe the implementation of a hierarchical distributed control system for a chemical pilot plant using industrial standard components. The control system consists of three components: regulatory control (mainly PID controllers), multivariable control (Dynamic Matrix Control controller) and the optimization and economic layer (Linear Programming). This work also includes details of the implementation tools: PROFIBUS, LabVIEW and data acquisition modules. The paper concludes with two possible solutions that deal with plants with integral behaviour. *Copyright (c) 2002 IFAC*

Keywords: Distributed control, Fieldbus, Integral behaviour, Predictive control, Linear programming.

1. INTRODUCTION

In order to implement distributed control systems in industrial environments robust communications tools are needed. These tools consist of a physical interface (number of wires, connectors, maximum distance and speed, etc.) and communication protocols. They are the so-called industrial buses or fieldbuses (J. Ayza, 2000 (In Spanish)).

There exist different commercial fieldbus products in the market, each of them with different physical interfaces and communication features. Profibus-DP ((PROFIBUS Nutzerorganisation E.V., 1999)) is the one used in this work. This fieldbus is mainly used in Europe, and there exist thousands of industrial applications and products (data acquisition modules and sensors). The physical interface of Profibus-DP is based on *RS*485 and it is mainly used for cycled transmission of data from the control system to the peripherals (sensors and actuators) and vice versa. Bus access of the different modules is implemented in a master-slave approach. Profibus-DP is suitable for schemes with one master node and several slave modules (slave demands can be attended in less than 1ms). Almost any control technique can be implemented on the hardware-software platform provided by Profibus-DP.

One of the most popular techniques in the chemical industries is the Dynamic Matrix Control (DMC)(Cutler and Ramaker, 1980),(Prett and Morari, 1986). This approach has been accepted by practitioners around the world from the end of the seventies. Nowadays, DMC is more than a control algorithm, commercial products that include the DMC controller also include identification packages and plant wide optimization programs (VanDoren, 1997),(Babatunde, 1986).

Some of the main features of the DMC algorithm are:

- Flexibility, due to the tuning parameters that the operator can modify on-line.

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- Anticipation. Future setpoint or disturbance changes and large delays can be taken into account using the prediction model.
- Constraints in the manipulated and controlled variables are considered explicitly.
- Coupling is minimized when dealing with MIMO systems.

1.1 *The three level approach.*

The implementation of the DMC algorithm in an industrial environment is not straightforward, questions as security and economic factors must be also considered. In this way the DMC is embedded in a hierarchical frame consisting of three levels (see figure 1).

In the upper level, optimization tools are used to obtain the steady-state values of the manipulated and controlled variables (MVs and CVs) of the plant. Economic factors as profit and cost involved in the process, and operational and security constraints are used to obtain economic indexes. These indexes will be minimized by optimization tools as Linear Programming (LP). The inclusion of the LP has been carried out in (Morshedi *et al.*, 1985), (Yousfi and Tournier, 1991),(Ying *et al.*, 1980) and (Sorensen and Cutler, 1998). LP-DMC stability is guaranteed when the process model is accurate and constraints are only considered in the LP block as shown in (Ying *et al.*, 1980). The optimization is performed using the steady-state model, the current values of the MVs $u(k)$ and the steady-state values of the CVs $\hat{y}_{ss}$. The results of the optimization are the *increment* steady-state values of the MVs (Δu_{ss}) and CVs (ΔSP), see figure 1.

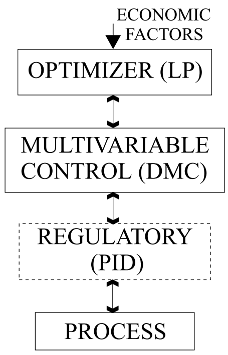

Fig. 1. The three level scheme.

In the intermediate level or multivariable level the DMC algorithm is carried out. In this level the values of the MVs are computed using the dynamic model of the plant together with the information coming from the upper level. Finally these values are sent to the regulatory level. DMC can be implemented without considering constraints because they are taken into account in the LP block.

In the regulatory level the information coming from the intermediate level can be used in two different ways: (a) as a setpoint for a PID controller (or any other SISO controller), (b) as the control action for the actuator. This level appears mainly as a security level in the plant. If the multivariable controller fails the regulatory system must move the plant to a safe operating point.

2. CONTROL SYSTEM IMPLEMENTATION

2.1 *Process description.*

The process is a chemical pilot plant where several industrial subprocesses can be found. The plant consists of three coupled subprocesses: temperature, level, and flow as shown in figure 2. The CVs and the operating points are:

Flow sensor (FT1/F)	50% of the maximum flow.
Level sensor (LT1/L)	50% of height of the tank.
Temperature sensor (TT1/T)	33^oC

The MVs and the operating points are:

Electrovalve (FV1/UF)	43.5%
Electrovalve (LV1/UL)	29%
Fan speed (TF1/UT)	30.8%

The disturbance variable is the temperature of the liquid in the main tank (D1). This temperature can be modified by a heating resistance. The operating point considered is 35^oC.

2.2 *The plant model.*

Although the plant is non-linear a linear process can be obtained around the operating point above. The model has been computed from data coming from a step-test.

$$\begin{pmatrix} F\,(\%) \\ L\,(\%) \\ T\,(C) \end{pmatrix} = \begin{bmatrix} G_{11} & 0 & 0 \\ G_{21} & G_{22} & 0 \\ 0 & G_{32} & G_{33} \end{bmatrix} \begin{pmatrix} UF\,(\%) \\ UL\,(\%) \\ UT\,(\%) \end{pmatrix} \quad (1)$$

$$G_{11} = \frac{2.3e^{-0.3s}}{0.73s+1} \quad G_{21} = \frac{0.027e^{-0.3s}}{s}$$
$$G_{22} = \frac{-0.021e^{-0.6s}}{s} \quad G_{32} = \frac{0.075e^{-15s}}{15.3s+1}$$
$$G_{33} = \frac{-0.095e^{-11s}}{25s+1}$$

Transfer functions G_{21} and G_{22} have integral behaviour, this will be considered in future sections. The actuators present two kind of non-linearities: saturation and dead-zone. These non-linearites can be taken into account as constraints in the MVs.

$$\begin{pmatrix} UF \\ UL \\ UT \end{pmatrix}_{max} = \begin{pmatrix} 58\% \\ 80\% \\ 100\% \end{pmatrix} \quad \begin{pmatrix} UF \\ UL \\ UT \end{pmatrix}_{min} = \begin{pmatrix} 19\% \\ 15\% \\ 0\% \end{pmatrix} \quad (2)$$

Due to security questions the CVs must be inside the following limits:

$$\begin{pmatrix} F \\ L \\ T \end{pmatrix}_{max} = \begin{pmatrix} 60\% \\ 60\% \\ 34C \end{pmatrix} \quad \begin{pmatrix} F \\ L \\ T \end{pmatrix}_{min} = \begin{pmatrix} 40\% \\ 40\% \\ 32C \end{pmatrix} \tag{3}$$

2.3 *Control objective.*

The pilot plant can be considered as a production process. During this process inlet stream is heated in the main tank (D1). After that, the liquid is moved to tank (C1) where it rests and finally it is cooled before leaving the system (see figure 2).

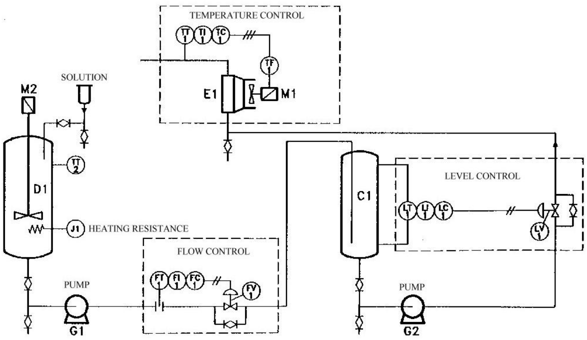

Fig. 2. Plant diagram.

The objective is to produce the maximum quantity of material, that means to increase the flow of the outlet stream to the maximum. At the same time the quality of the final product must be within certain limits (the quality of the product is related to the temperature of the outlet stream).

The variables involved in the production are:

F: is the flow measurement between tanks D1 and C1. If the level in tank C1 is maintained in a fixed position then F is also the flow of the outlet stream. Therefore F must achieve the maximum possible value and so the MV (UF).

L: is the level of the product in tank C1. This level should be fixed to make sure that F is also the flow of the outlet stream. At the same time, possible overflows or underflows must be avoided.

T: is the temperature of the final product. This temperature must be low (due to the required quality of the final product) therefore the energy consumption of the cooler fan must be taken into account.

2.4 *Programming and system tools.*

The main elements of the control system are:

- An industrial computer with the communications card CIF-30 DP by Hilscher. This module is a master node in the network and performs tasks of configuration of the network and cycled exchange of information with the slave nodes.
- A BK3100 I/O module by Beckhoff. This module is a slave node in the network and performs tasks of data acquisition.

Some of the parameters of the network are:

- Speed: $12 Mbits/s$.
- Poll cycle (from master to slave): $0.95ms$.
- Slave watchdog deadline $200ms$.
- Master monitoring watchdog $1000ms$.

The application has been developed with LabVIEW (Wells and Travis, 1997). The main window is shown in figure 3 where the components of the plant are shown: tanks, pipes, pumps and valves. The operator can check the current state of any setpoint, sensor or actuator of the plant and there exits also a window to check the trends of these variables and saving all these data.

The controller is a function written in 'C' language and embedded in a dynamic library (DLL) for Windows that will be included in LabVIEW. The math routines (i.e. linear programming) come from the NAG package (NAG, 1997).

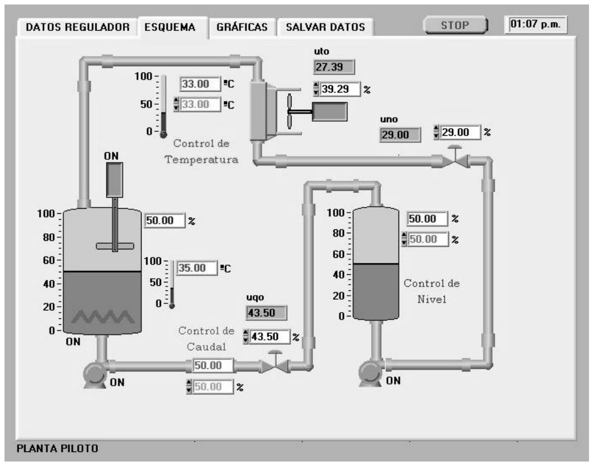

Fig. 3. Main window of the user interface.

Two possible control schemes are described in next subsections.

2.5 *Approach without PID type controllers.*

In this case no PID controllers are used in the regulatory level so the MVs computed by the DMC controller are applied to the plant.

2.5.1. *The linear program.* Cost function (Φ) will consist of two parts: profit (due to the final product) and cost (due to the energy used during the process). The final product must be written in terms of CVs and MVs. Whereas to obtain the cost only MVs are needed as the cost is normally proportional to their value.

$$\begin{array}{l} Profit = G_{11} \cdot \Delta U_L \cdot price \\ Cost = C_F \cdot \Delta U_F + C_L \cdot \Delta U_L + C_T \cdot \Delta U_T \\ \Phi = Cost - Profit \end{array} \tag{4}$$

where *price* is the price in the market of the final product, and C_F, C_L and C_T are the costs due the movement of the MVs. All the variables are considered in their incremental version (increments over the operating point).

In order to implement the previous index in a Linear Programming environment with constraints, some modifications must be carried out. Positive increments (ΔU_i^+) and negative increments (ΔU_i^-) of the variables can be penalized in a different way. The total increment is $\Delta U_i = \Delta U_i^+ - \Delta U_i^-$. Constraints on CVs will be implemented as soft constraints while MVs constraints will be considered hard. Therefore slack variables $\xi_i \geq 0$ and weighting factors W_i must be defined for every CV (positive and negative increments).

$$\begin{array}{rl} \Phi = & K_F^+ \cdot \Delta U_F^+ + K_F^- \cdot \Delta U_F^- + \\ & + K_L^+ \cdot \Delta U_L^+ + K_L^- \cdot \Delta U_L^- + \\ & + K_T^+ \cdot \Delta U_T^+ + K_T^- \cdot \Delta U_T^- + \\ & + W_1^+ \cdot \xi_1^+ + W_1^- \cdot \xi_1^- + W_2^+ \cdot \xi_2^+ + \\ & + W_2^- \cdot \xi_2^- + W_3^+ \cdot \xi_3^+ + W_3^- \cdot \xi_3^- \end{array} \tag{5}$$

Finally, there exists one quality requirement in the final product: the temperature of outlet stream must be low, hence the increments of the temperature over the operating point are penalized with the C_{temp} factor.

$$\begin{array}{l} K_F^+ = -K_F^- = C_F \\ K_L^+ = -K_L^- = C_L - G_{11} \cdot price + G_{32} \cdot C_{temp} \\ K_T^+ = -K_T^- = C_T + G_{33} \cdot C_{temp} \end{array}$$

Implementation examples of constraints on the VCs and the MVs are:

$$\begin{array}{l} G_{11} \cdot \Delta U_F^+ - G_{11} \cdot \Delta U_F^- - \xi_1^+ \leq y_{1\max} - y_{1ss} \\ -G_{11} \cdot \Delta U_F^+ + G_{11} \cdot \Delta U_F^- - \xi_1^- \leq -(y_{1\min} - y_{1ss}) \end{array} \tag{6}$$

$$\begin{array}{l} \Delta U_F^+ \leq u_{1\max} - u_1(k) \\ \Delta U_F^- \leq -(u_{1\min} - u_1(k)) \end{array} \tag{7}$$

As the optimization (Alonso *et al.*, 2000 (In Spanish)) is carried out in steady state, constraints are considered only in steady state (subindex *ss*). Finally the slack variables are also constrained:

$$\begin{array}{rclrcl} -\Delta U_F^+ & \leq & 0 & -\Delta U_F^- & \leq & 0 \\ -\xi_1^+ & \leq & 0 & -\xi_1^- & \leq & 0 \\ \xi_1^+ & \leq & \xi_{1\max} & \xi_1^- & \leq & \xi_{1\min} \end{array} \tag{8}$$

2.5.2. *Integral behaviour.* The application of the LP-DMC approach on the pilot plant requires stable behaviour in all the control loops, but the level loop presents integral behaviour. One approach is to consider a PID controller for that loop and then implement LP-DMC over that structure. Another possibility is to consider slope variables (Sorensen and Cutler, 1998). In this way the slope of the CVs is taken into account instead of the value of the CVs itself. Therefore the slope must be included in the index and also in the constraints. In the constraints case:

$$\begin{array}{l} G_{21}^* \Delta U_F^+ - G_{21}^* \Delta U_F^- + G_{22}^* \Delta U_L^+ - \\ -G_{22}^* \Delta U_L^- - \xi_2^+ = SLOPE - y_{2ss} \\ -G_{21}^* \Delta U_F^+ + G_{21}^* \Delta U_F^- - G_{22}^* \Delta U_L^+ + \\ +G_{22}^* \Delta U_L^- - \xi_2^- = -(SLOPE - y_{2ss}) \end{array} \tag{9}$$

where G^* is the static slope of the transfer function. When dealing with slope variables the constraints are considered as equality constraints. The objective is to maintain the slope of the variable close to zero so the CV is maintained in the operating point. Once the solution is obtained (Δu_{ss} and ΔSP) the new setpoint is computed as:

$$SP_{new} = SP_{old} + G \Delta u_{ss} \tag{10}$$

where $SP_{old} = y_0 + \hat{y}_{ss}$ and G is the corresponding static gain.

If the CV is a slope variable the prediction horizon N of the DMC must be taken into account:

$$SP_{new} = y_0 + (\hat{y}_{ss} + G^* \Delta u_{ss}) N \tag{11}$$

Finally, the solution of linear program Δu_{ss} is a constraint for the DMC controller:

$$\sum_{i=1}^{Nu} \Delta u_i = \Delta u_{ss} \tag{12}$$

In this way the stability of the control loop is guaranteed (this is another approach of the terminal constraint used in predictive control (Mayne *et al.*, 2000)).

2.6 *Approach with PID type controllers.*

In this case a PI controller for the level loop is used. Therefore all the subprocesses under the DMC controller are stable. The main consequence is that the new CV is not the slope of the level but the setpoint of this PI controller.

First of all a new model of the process must be obtained. This model incorporates the following PI controller:

$$Gr\left(z^{-1}\right) = \frac{Kc\left(1 + {}^{Ts}/_{Ti}\right) - Kcz^{-1}}{1 - z^{-1}} \tag{13}$$

and the new process model becomes:

$$\begin{pmatrix} F\,(\%) \\ L\,(\%) \\ T\,(C) \end{pmatrix} = \begin{bmatrix} G'_{11} & 0 & 0 \\ G'_{21} & G'_{22} & 0 \\ G'_{31} & G'_{32} & G'_{33} \end{bmatrix} \begin{pmatrix} UF\,(\%) \\ SP_L\,(\%) \\ UT\,(\%) \end{pmatrix} \tag{14}$$

$$
\begin{array}{ll}
G'_{11} = G_{11} & G'_{21} = \dfrac{G_{21}}{1+GrG_{22}} \\
G'_{22} = \dfrac{GrG_{22}}{1+GrG_{22}} & G'_{31} = \dfrac{-G_{32}GrG_{21}}{1+GrG_{22}} \\
G'_{32} = \dfrac{G_{32}Gr}{1+GrG_{22}} & G'_{33} = G_{33}
\end{array}
$$

and the new cost index is:

$$
\begin{aligned}
&\Phi = K_F^+ \cdot \Delta U_F^+ + K_F^- \cdot \Delta U_F^- + K_{SP_L}^+ \cdot \Delta SP_L^+ + \\
&+K_{SP_L}^- \cdot \Delta SP_L^- + K_T^+ \cdot \Delta U_T^+ + K_T^- \cdot \Delta U_T^- + \\
&+W_1^+ + \cdot \xi_1^+ W_1^- \cdot \xi_1^- + W_2^+ \cdot \xi_2^+ + \\
&+W_2^- \cdot \xi_2^- + W_3^+ \cdot \xi_3^+ + W_3^- \cdot \xi_3^-
\end{aligned} \tag{15}
$$

$$
\begin{aligned}
&K_F^+ = -K_F^- = C_F - G'_{11} \cdot price + G'_{31} \cdot C_{temp} \\
&K_{SP_L}^+ = K_{SP_L}^- = C_{SP_L} + G'_{32} \cdot C_{temp} \\
&K_T^+ = -K_T^- = C_T + G'_{33} \cdot C_{temp}
\end{aligned}
$$

3. RESULTS

3.1 *Approach without PID type controllers.*

The parameters of the DMC are: sampling time of DMC and LP, $T_s = 0.3seg.$, equal concern error α and move suppression factors λ:

$$
\begin{array}{ll}
\alpha = [1, 4.5, 8] & \lambda = [10, 1, 1] \\
u_0 = [43.5, 29, 30.8] & y_0 = [50, 50, 33] \\
N_1 = [1, 1, 1] & N_2 = [60, 60, 750] \\
Nu = [2, 3, 6] &
\end{array}
$$

where N_1 and N_2 define the prediction horizon and Nu defines the control horizon, and u_0 and y_0 define the operating point. Finally, the constraints and cost factors are:

$$
\begin{array}{ll}
SLOPE = 0 & price = 3000 \\
u_{max} = [58, 80, 100] & u_{min} = [19, 15, 0] \\
y_{max} = [60, 60, 34] & y_{min} = [40, 40, 32] \\
\xi_{max} = [2, 0, 1] & \xi_{min} = [2, 0, 1] \\
C_{temp} = 100000 & C_T = 1900 \\
C_L = 20 & C_F = 20 \\
W_1^+ = 200000 & W_1^- = 200000 \\
W_2^+ = 2000000000 & W_2^- = 2000000000 \\
W_3^+ = 500000 & W_3^- = 200000
\end{array}
$$

If $k_F^+ = k_F^-$ the response is in figure 4. In this example the production is at the maximum rate (maximum flow) and minimum temperature (maximum quality). At the same time the level is maintained close to the operating point (slope close to zero). Constraints are also satisfied and slack variables are always close to zero.

3.2 *Approach with PID type controllers.*

The tuning parameters of the level PI controller are: $Kc = -10$ $({}^{UL(\%)}/_{L(\%)})$ and $Ti = 40s$. In this case the sampling time of the PI controller must be smaller than the DMC controller. As the DMC works at $1.2s$ the PID controller sampling time is $0.03s$. Note that the sampling time of the DMC is different from the previous case.

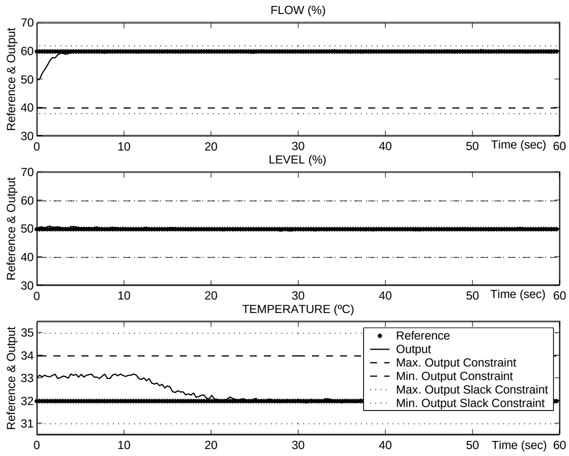

Fig. 4. System response using LP-DMC with slope variables.

The tuning parameters of the DMC are:

$$
\begin{array}{ll}
\alpha = [1, 4.5, 8] & \lambda = [10, 20, 1] \\
N_1 = [1, 1, 1] & N_2 = [30, 450, 280] \\
Nu = [2, 3, 6] &
\end{array}
$$

the operating point is:

$$
u_0 = [43.5, 50, 30.8] \qquad y_0 = [50, 50, 33]
$$

Finally the constraints and cost factors are:

$$
\begin{array}{ll}
u_{max} = [58, 60, 100] & u_{min} = [19, 40, 0] \\
y_{max} = [60, 60, 34] & 6y_{min} = [40, 40, 32] \\
\xi_{max} = [2, 2, 1] & \xi_{min} = [2, 2, 1] \\
C_{temp} = 100000 & C_T = 1900 \\
C_{SP_L} = 10 & C_F = 20 \\
W_1^+ = 200000 & W_1^- = 200000 \\
W_2^+ = 2000000000 & W_2^- = 2000000000 \\
W_3^+ = 500000 & W_3^- = 200000 \\
price = 3000 &
\end{array}
$$

The system response is shown in figure 5. These results are quite similar to the ones in figure 4.

4. CONCLUSIONS

After the application of the three level approach (LP-DMC and LP-DMC-PI versions) to the pilot plant some conclusions can be drawn:

The economic and safety information of the plant are taken into account when the control actions are computed. The process is moved to the optimum operating point.

The setpoints are changed in a smooth way, this is quite related to stability (Ying *et al.*, 1980).

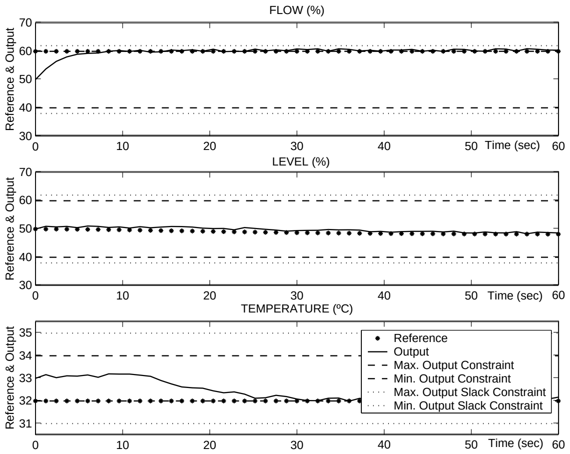

Fig. 5. Response of the LP-DMC-PI with Ts = 1.2.

Constraints are not always satisfied during the transient time. This is due to fact that only the linear program considers constraints but not the DMC. The DMC controller only tries to satisfy the terminal constraint (equation 12).

Both versions LP-DMC and LP-DMC-PI achieve similar performance: maximum quantity of product and maximum quality. The LP-DMC-PI version can deal with unstable processes or processes with integrators whereas the LP-DMC can deal only with stable processes and processes with integrators. However in the LP-DMC case the level in tank C1 is only maintained in the operating point (due to the use of slope variables) whereas in the LP-DMC-PI that level can be moved (depending on the Linear Program). In this sense LP-DMC-PI is more flexible than LP-DMC.

The sampling time is another important factor. In the LP-DMC case only one sampling time is needed (depending on the fastest loop it can be calculated) whereas in the LP-DMC-PI case two sampling times are needed and this is not straightforward. If the sampling time of the PI controller is chosen based only on the loop that controls (the level loop in this work) this can lead to a poor performance. As the DMC sampling time is larger than the one in the PI controller this can produced bad sampling of the fastest loops. For example, if the PI sampling time is chosen only taken into account the level loop then the DMC sampling time must be larger and therefore it may be not the right one for the flow loop (the fastest loop in the process). This can be checked comparing the flow responses in figures 4 and 5.

Finally, the LP module does not consider error in the models. This could lead to a poor operating point. An application taking into account these errors could improve the performance of the current algorithm.

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