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IMPLEMENTATION OF CONSTRAINED PREDICTIVE OUTER-LOOP CONTROLLERS: APPLICATION TO A BOILER CONTROL SYSTEM

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Abstract: This paper addresses the problem of implementing predictive controllers for supervisory level control systems. In this configuration the manipulated variables calculated by the Predictive Controller are used as command signals for the Distributed Control Systems, which provide references to the operator-tuned local PID controllers that act on the physical system. This structure introduces the problem of loosing of performance if the inner-loop controllers are re-tuned. The paper discusses the solution to this problem based on the use of a two-degrees-of-freedom structure in the inner loop, that separates open and closed-loop properties. Both design guidelines and robustness issues are discussed. *Copyright © 2002 IFAC*

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1. INTRODUCTION

For multivariable process control problems with strong interactions between the controlled and manipulated variables and strict constraints, the conventional multiloop PID control configuration may not provide adequate control performance (Seborg, 1994). Model-Based Predictive Control (MBPC) solves these problems by predicting future process behaviour and calculating control variables taking into account the process constraints (Clarke et al. 1987). These techniques have been applied very successfully to different Process Control Problems (Camacho and Bordons 1995, Froisy 1994).

A common structure for a MBPC in industrial process control problems is shown in figure 1. The MBPC calculates the future control signals based on the measured variables. These control signals (manipulated variables), are sent to the Distributed Control System as command signals for the actuators (such as valve positioning commands). Local PID controllers act on the physical system to obtain the desired manipulated variable, which is fed-back to both the local PID and the predictive controller. This is one of the structures discussed by Lee et al. (1997), where it is called a "Cascade control – series connected system". It was also studied by Saez et al. (2000), where it was proved that under certain conditions the master controller could be selected to make control characteristics independent of the slave controller. However the solution proposed by Saez et al. (2000) is not used in this paper, as it is based on perfect knowledge of the slave-loop controller (which is not always possible), and generates a pole/zero cancellation between slave and master controllers (which in certain cases is undesirable).

It must be pointed out that direct control of the plant by the predictive controller is also frequent in practical implementations (an example for a Steam Generator is presented in Khotare et al., 2000). However, plant operators in industry are not usually ready to permit direct control of the plant by they predictive controllers, unless they are already very familiar with predictive control. The implementation discussed in this paper makes possible to prove the improvement in performance of predictive controllers, which could later be upgraded to direct control, if desired.

In most industrial implementations these PID loops in the slave loop are tuned by the operators of the plants, based on their knowledge of the local process. The Predictive Controller includes, in the models used for prediction, the PID, actuator and measurement filter dynamics. That means that if any of the inner-loop controllers, which will be referred to as *slave* controllers, is re-tuned the real system differs from the model used for prediction. This difference causes

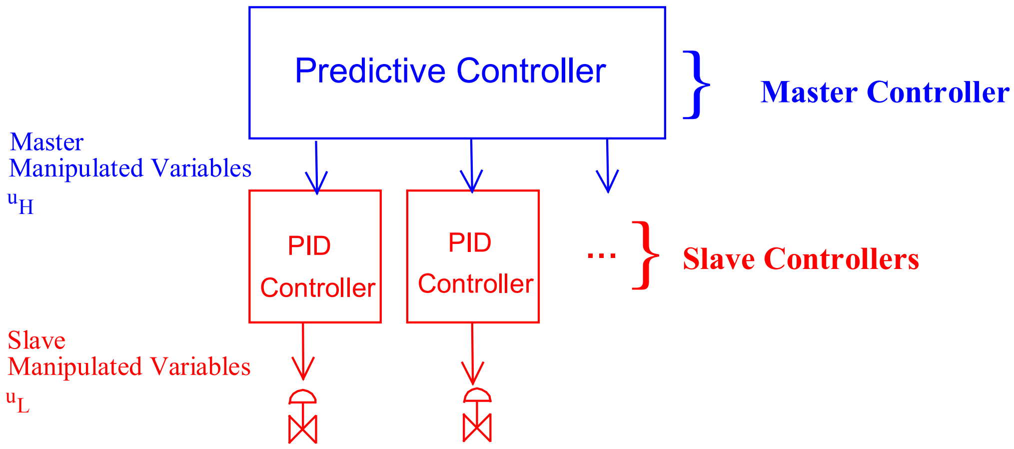

Figure 1: Control System Structure

a worsening of performance of the overall system, even making the system unstable. In Saez et al. (2000) it was shown that if the supervisory level control was an unconstrained GPC and the plant controller model were perfect the master controller would not depend on the slave controller.

This paper discusses a method of solving this problem for a general master controller in the presence of constraints, based on the augmentation of the slave controller with a Two-Degrees-of-Freedom (2DoF) structure, that includes Prefilter and Feedforward Compensators. In Tadeo and Alvarez (1998) it was shown that by using this structure the dynamics "seen" by the Predictive Controller are not affected when the slave-controller parameters change. This paper discusses the selection of these blocks, presenting a particular selection that simplifies greatly the problem of transforming constraints on the manipulated variables from the slave to the master controller. The effect of uncertainty in the plant and controller is also studied. The application of this method to a simulated industrial boiler will show that the good performance obtained with the control structure discussed in this paper is maintained even when retuning the slave controllers.

2. INNER-LOOP CONTROL STRUCTURE

To implement the slave controller a two-degrees-of-freedom (2DoF) structure is proposed that separates open-loop properties from closed-loop properties (Pernebo, 1981). 2DoF structures provide to the designer the option of separating the achievement of desired *regulating properties* (robust stability, disturbance rejection and measurement noise attenuation, also known as *closed-loop* or *feedback properties*) and *servo properties* (command tracking with reduced control effort, also known as *open-loop properties*). The use of a 2DoF structure has long been common practice in industry. Notwithstanding this, it is only recently that control theorists have understood the advantages of designing separately for open-loop and closed-loop properties (Vidyasagar, 1985, Wolovich, 1995). In the literature different 2DoF structures have been proposed and applied to solve different engineering problems (Grimble, 1998, Tadeo and Holohan, 1998, Yaesh and Shaked, 1991, Youla and Bongiorno, 1985 and the references therein).

From all the available 2DoF structures the one shown in Figure 2 is selected in this paper because it is readily available in most industrial control systems. Its application to Model Reference Adaptive Controller is shown in (Vilanova, 1996). By an adequate selection of the Prefilter and Feedforward Compensators, it is will be proved that with this structure the closed-loop properties can be designed independently of the Feedback Compensators. This is known as a type of separation principle, which is proved in the next section.

3. SEPARATION PRINCIPLE

Properties of the proposed control structure are now discussed. As the slave controllers are normally SISO, these properties are presented only for the SISO case. However they also hold in the MIMO case, with a more cumbersome notation.

Supposing that the plant to be controlled is linear and time-invariant, it is possible to factorize its transfer function as the product of a stable transfer function and the inverse of another stable transfer function: $G = ND^{-1}$ (See Vidyasagar, 1985, for a extended discussion). This factorization is called a *stable coprime factorization* if N and D are stable and have no common unstable zeros. Given such stable coprime factorization it is always possible to find X, Y stable and coprime (without any common unstable zero) such that the Bezout identity holds:

$$DX+NY=I$$

Then (Youla, 1976) the set of controllers which stabilize the feedback system is given by:

$$K = (Y - QD)(X - NQ)^{-1}$$

where Q is any stable transfer function. In other words, a transfer function K stabilizes the system if and only if it can be written in this form for some

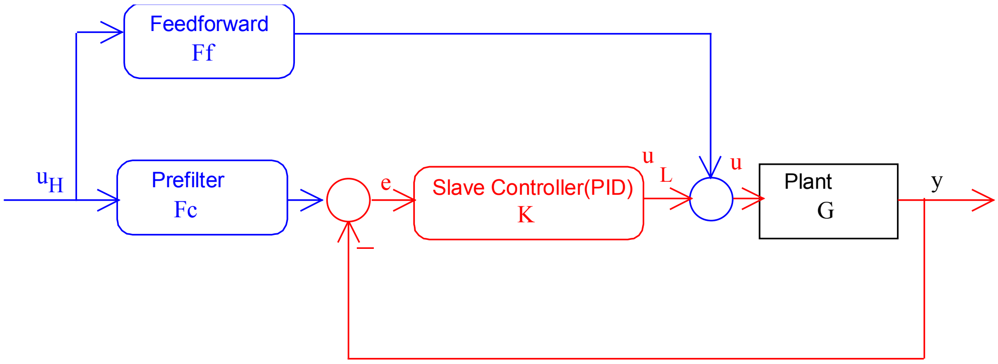

Figure 2: Proposed Control Structure

stable Q. (From now on T_{wz} denote the transfer function between the signal w and the signal z).

Lemma 1 (Separation Principle) (Tadeo and Alvarez, 1998)
- Let G be a linear and time-invariant system, expressed as a stable coprime factorization $G = ND^{-1}$.
- Let K be any stabilizing compensator.
- If the control signal is calculated using the 2DoF structure shown in Figure 2:
$$u = F_f u_H + K(F_c u_H - y).$$
- If $F_c = NR$ and $F_f = DR$, where R is any stable transfer function.

Then
The open-loop properties are independent of the feedback controller K. Moreover:
$$T_{u_H y} = F_f$$
$$T_{u_H u} = F_c$$

Lemma 2 (Separation Principle in the presence of uncertainty)
- Let G be the LTI nominal model of the plant, with stable coprime factorization $G = ND^{-1}$
- Let the uncertainty in the plant model described by an inverse multiplicative uncertainty: $\tilde{G} \equiv G(I + \Delta)^{-1}$ be the set of plants
- Let K be any stabilizing compensator for the set of plants $\tilde{G}$ (This implies Robust Stability).
- Let $S \equiv (I + KG)^{-1}$
- If the control signal is calculated using the 2DoF structure shown in Figure 2:
$$u = F_f u_H + K(F_c u_H - y).$$
- If $F_c = NR$ and $F_f = DR$, where R is any stable transfer function.

Then
$$T_{u_H y} = F_f (I + S\Delta)^{-1}$$
$$T_{u_H u} = F_c (I + \Delta)(I + S\Delta)^{-1}$$

Proof: straightforward calculations following the proof of Lemma 1 in Tadeo and Alvarez (1998) and using the fact that $S = DX$ (X from the Bezout Identity).

Remarks:
- This result implies that in the presence of uncertainty the open-loop properties depend on the feedback controller K. However, observe that, if the slave loop is correctly tuned, in steady-state the transfer function $T_{u_M y}$ does not change in the presence of uncertainty: $S(z=1) = 0$. Then, the steady-state model does not change in the presence of uncertainty.
- Equivalent results are obtained using other descriptions of the uncertainty. However, when

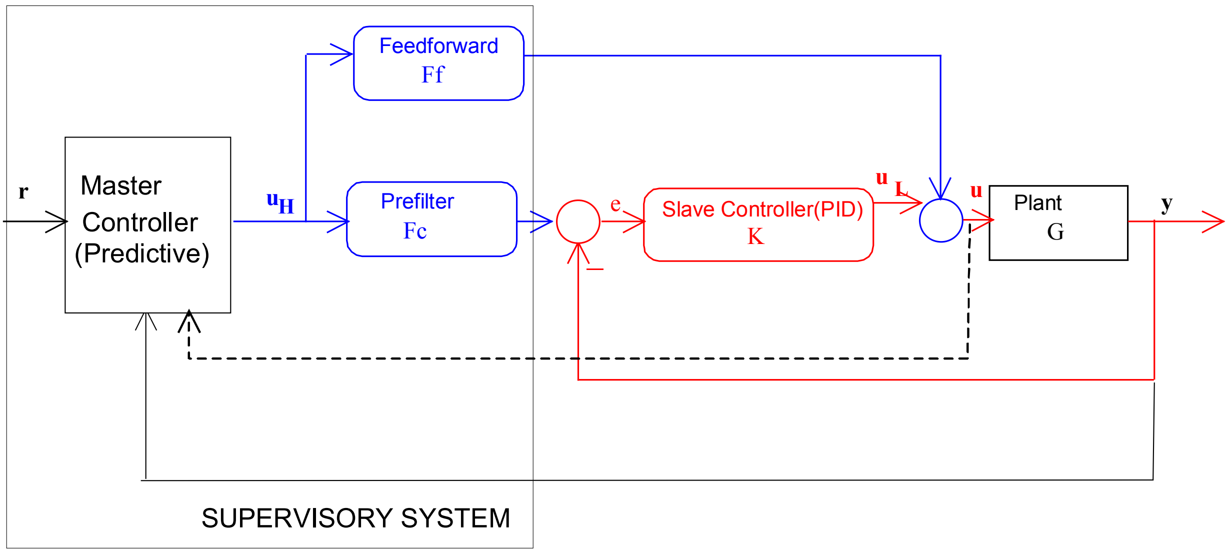

Figure 3: Complete control structure

considering an inverse multiplicative uncertainty, as it is proposed in this paper, there is a simple dependence between $T_{u_H y}$ and Δ, which simplifies the robustness analysis (or synthesis) of the master controller.

- Observe that although in the expressions of F_c and F_f appear the coprime factorization of the plant, this expressions are only used to prove the separation principle. In practice, as it is discussed later, to select F_c and F_f it is not necessary to calculate the coprime factorization: the only properties needed to select F_c and F_f are the following: F_c should be a stable transfer function containing as zeros the unstable zeros of the plant, and F_f should be a stable transfer function containing as zeros the unstable poles of the plant.

4. PREDICTIVE CONTROLLER IMPLEMENTATION

So far, the implementation of the second degree-of-freedom in the slave loop has been presented. The effect of these additional blocks in the MBPC that acts as supervisory system is now discussed.

The complete control structure studied in this paper is shown in Figure 3. To make this structure compatible with most distributed control systems, the MBPC can be augmented with the Feedforward and Prefilter Compensators of the slave loops, as shown in the boxed area of Figure 3. Observe that in many supervisory control systems the output signals fed to the slave controller and the master control could be different (as it is usual in a valve positioning system, where the output controlled by the slave loop (the valve position) is different from the output considered by the predictive controller (temperature, pressure, etc). This paper concerns the simplified structure in Figure 3. Without loss of generality, results in this paper can be easily extended to other problems, as it will be discussed later.

As proved in Lemma 1, the transfer functions between each manipulated variable u_H and the corresponding plant control signal u is given by $T_{uu_H} = F_c = NR$. That means that (in the nominal case) the plant seen by the MBPC is just $T_{u_H y} = NR$ (See figure 3). This is the model that the MBPC should consider for prediction. Moreover, the effect of the manipulated variable u_H over the signal sent to the actuator (u) is given by $T_{uu_H} = F_f = DR$. That means that physical constraints can be readily transformed to constraints on the manipulated variables, which are the optimization variables. This will be discussed later.

In the control valve positioning problem, the only change is that the model considered in the master MBPC is $T_{u_H y} = PNR$, where P is the valve transfer function.

For comparison purposes, the plant seen by the MBPC, without the second degree-of-freedom ($F_c = 1, F_f = 0$) is the complementary sensitivity of the slave loop: $T_{yu_H} = GK(I+GK)^{-1}$. This transfer function depends on the tuning of the slave controller (it depends on K). However, including the second degree-of-freedom as needed in lemma 1 makes it independent (it does not depend on K). The same can be said of the control signal, which transfer function is the control sensitivity: $T_{uu_H} = K(I+GK)^{-1}$. In contrast with the contrary of the 2DoF structure, this transfer function depends on K, giving as a result a lack of robustness of the MBPC to changes in K.

5. CONSTRAINED MULTIVARIABLE PREDICTIVE CONTROLLER

MBPC is a control strategy based on the explicit use of a model to predict the process output over a long period of time. From the proposed MBPC the Generalized Predictive Controller (Clarke et al., 1987, Camacho and Bordons, 1995, Maciejowski, 2002) was selected, since it gives understandable and intuitive solutions, taking into account process and operating constraints.

The implementation of a MBPC, with on-line control system reconfiguration characteristics, is now discussed. This controller is based on the one presented in (Alvarez and Prada, 1997), where input and output constraints were considered, and a constraint handling procedure was applied when the optimization problem had no feasible solution.

The implementation of this MBPC with the control structure depicted in figure 4 is now discussed: The bounds on the slave manipulated variable u must be transformed to bounds on the manipulated variable u_H minimized in (1). By the result in Lemma 1:

$$u \equiv T_{uu_H} \cdot u_H$$

that is, bounds on where . means convolution. Thus, the bounds on u can be transformed to bounds on u_H by deconvolution. As pointed out by Lee et al. (1998) this is not trivial, since the relationship among the variables are usually not static. However, with the proposed augmentation of the slave controller, it is possible to simplify this transformation, using the facts that there is an additional degree of freedom in the selection of R, and it is possible to transform input constraints to output constraints.

The proposed selection of Prefilter and Feedforward Transfer Functions is now discussed separately for stable and unstable plants.

Stable plant:
If G is stable then there is a simple solution: it is possible to select

$$F_f \equiv \alpha \equiv 1/\beta$$

$$F_c \equiv G/\beta$$

with this selection, constraints on u are transformed on constraints on u_H by dividing with a constant, which is the plant gain:

$$u_H \equiv u/\beta$$

Unstable plant:
If G is unstable, the unstable poles of the plant must be zeros of F_f the simplest solution is to select

$$F_f \equiv \frac{1}{\beta \prod_{k=1}^{n}(1-a_k)} \prod_{k=1}^{n}\left(1-a_k z^{-1}\right)$$

$$F_c \equiv G // F_f$$

where $\{a_k\}_{k=1}^{n}$ are the poles on or outside the unit disk (which are supposed to be simple, without loss of generality), and $G//F_f$ corresponds to G/F_f after cancellation of the $\{a_k\}_{k=1}^{n}$ poles and zeros.

Observe that now instead of using deconvolution to transform constraints on u to constraints on u_H it is proposed to include these constraints on u as output constraints in the supervisory control, considering u as an additional output, with the transfer function:

$$y_2 \equiv u = F_f u_H$$

In most predictive control implementation, this transformation from input constraints to output constraints usually increases the number of constraints in the optimization problem, as the output horizon is usually longer than the control horizon. This might increase the infeasibility problems, so an infeasibility handler (Teresa et al., 1997) might be needed.

6. APPLICATION TO A BOILER CONTROL SYSTEM

The application of the control technique presented in this paper to a simulated industrial control problem is now presented. The control problem involves the servocontrol of a boiler, which is part of a sugar manufacturing factory in Cuba (Tadeo et al., 1996). A model of the plant was identified from operating data. The controller design is based on a simplified first-order plus dead-time model. Denoting by $p(s)$ the steam pressure (controlled variable), $F_c(s)$ the inlet fuel flow (manipulated variable) and $F_v(s)$ the steam flow demand (measured disturbance), the approximated model is:

$$p(s) = \frac{k_0 e^{-\nu s}}{\tau s + 1} F_c(s) + \frac{k_v}{\tau_v s + 1} F_v(s)$$

where the normalized nominal parameters identified in the real plant are

$$K_0 = 1.73, \nu = 3.0, \tau = 10.3, K_v = 0.1, \tau_v = 1.0$$

The inlet fuel flow is controlled by a local discrete PI controller that regulates the opening of a set of valves, with nominal tuning parameters K_p=0.1 and T_i=20. This PID controller is frequently re-tuned by the operators of the plants due to changes in the process disturbances. The valves dynamics can be roughly approximated by the linear-model:

$$F_c(s) = \frac{k_v e^{-\delta s}}{\tau_v s + 1} u_S(s) + \frac{k_f}{\tau_f s + 1} P_f(s)$$

where P_f is a disturbance acting on the fuel flow, and the nominal parameters are:

$$k_v = 1.0, \tau_v = 1.0, \delta = 3, k_f = 0.1 \text{ and } \tau_f = 0.1$$

The implementation of a MBPC following the ideas of Alvarez and Prada (1997) was presented by Tadeo et al. (1996). Although improved performance and important energy savings were obtained, these improvements were lost whenever the valve-positioning PID controller was re-tuned. The step response of the nominal system controlled by the MBPC with the 1DoF structure and nominal PID tuning parameters is depicted in Figure 4. It can be seen that, in the nominal case, the control system performs adequately, improving the regulation characteristics.

The effect of changing the PID tuning is shown in Figure 4, where the step response is simulated when the PID gain (K_p) is increased from 0.1 to 0.2 and when it is decreased to 0.05. It can be seen that the performance is worsened when changing the PID gain, even making the system unstable for gains greater than 0.2. A similar result is obtained when varying the integral time. This fact prompted the augmentation of the control structure to the 2DoF

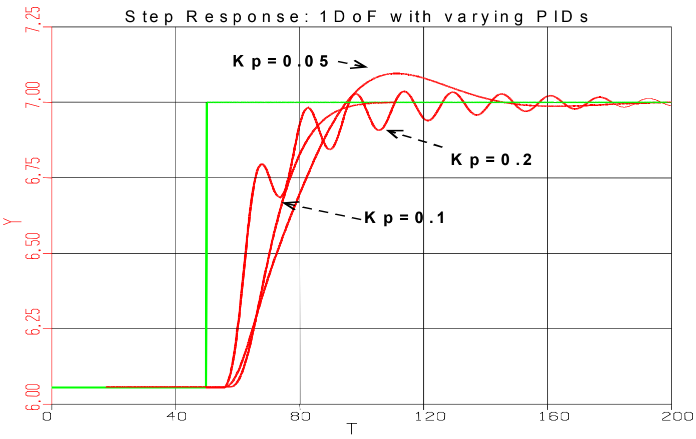

Figure 4: Step Response with 1DoF Control Structure and Different Tuning Parameters

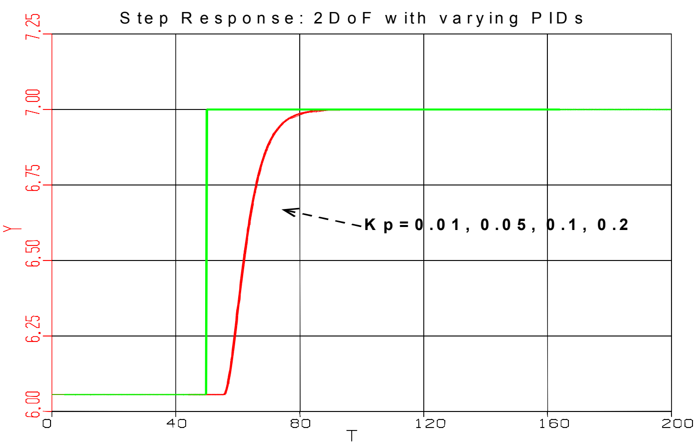

Figure 5: Step Response with the Proposed Control Structure and Different Tuning Parameters

structure discussed previously in this paper.

To augment the MBPC the discrete-time equivalent of the plant model (with sampling time *Tsamp=1.5 min*) was calculated. As it is an stable system, following the ideas proposed in this paper, the following filters are selected:

$$F_f = \frac{1}{k_v} \qquad F_c = \frac{2z^{-2}}{(1+\tau^*)+\left(1-\tau^*\right)z^{-1}}$$

(where $\tau^* = \dfrac{2\tau}{Tsamp}$ and *Tsamp* the sampling time)

The MBPC was augmented as depicted in Figure 3, and its closed-loop response studied for different controller parameters. The simulations corresponding to K_p=1.0, 0.2, 0.05 and 0.01 are shown in Figure 5. It is possible to check that the closed-loop behavior is maintained despite these important changes in the slave controller parameters, improving the robustness of the complete control system.

7. CONCLUSIONS

This paper has addressed the problem of implementing Model Based Predictive Controllers as supervisory control systems in a cascade structure. In this configuration, common in process control problems, the manipulated variables, calculated by the Predictive Controller, are used as command signals for the actuators (such as valve positioning commands) by the Distributed Control System. Then the Distributed Control System uses local PID controllers to act on the physical system. The main problem of this configuration is that the system is not robust against slave controller variations: performance of the overall system can worsen when any of the slave controllers is re-tuned; a situation that happens often during the normal operation of an industrial plant.

In this paper, the properties of a solution to this problem has been studied, and design guidelines for stable and unstable plants have been presented. The structure is based on augmenting the slave with a two-degrees-of-freedom structure, which command tracking properties independent of the feedback controller tuning. The paper has shown how it is possible to augment the predictive controller with this two-degrees-of-freedom structure, and that an adequate selection of the filters makes possible to transform constraints on the input to constraints on the input in the supervisory level (stable case) or constraints on a secondary output variable (unstable).

This solution has been applied to an industrial boiler control problem, where there is one inner loop controlling fuel flow, which can be re-tuned due to changes in the process. By using simulation it has been shown that using the technique presented it is possible to maintain the performance of the predictive control system despite re-tuning of the low level controller. This has been compared with the situation where the slave is not augmented by a Two-Degrees-of-Freedom structure, showing that changes in the parameters of the slave controller affect the performance, whereas adding the Prefilter and Feedforward Compensators maintains the nominal performance for a wide range of tuning parameters.

It should be noted that, although the proposed control structure has been presented in the context of MBPC Control, the techniques discussed in this paper can be applied to other supervisory control schemes.

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