# Predictive control of thermal Power Plants

C. Aurora[1], L. Magni[1,*,†], R. Scattolini[1], P. Colombo[2], F. Pretolani[2] and G. Villa[2]

[1] *Dipartimento di Informatica e Sistemistica, Università di Pavia, via Ferrata 1, 27100 Pavia, Italy*
[2] *CESI, via Rubattino, 54, 20134 Milano, Italy*

## SUMMARY

This work presents the results of a project aimed at verifying the applicability of industrial model predictive control (MPC) to thermal Power Plants. The research is motivated by the need to improve the efficiency of power plants so as to cope with the high levels of competition induced by the deregulation of the energy market. A detailed plant simulator, already used for operators training and controllers tuning, is coupled to an industrial software package implementing the dynamic matrix control algorithm. The achieved results witness the great potentialities of MPC, with respect to classical decentralized schemes, in terms of economical savings, reduction of pollutants, improved flexibility, easier tuning and better documentation. Copyright © 2004 John Wiley & Sons, Ltd.

KEY WORDS: model predictive control; thermal Power Plants

## 1. INTRODUCTION

The deregulation of the energy market in western Europe has dramatically increased the competition among the energy producers. This calls for higher and higher levels of efficiency in the management of the operating units in order to fulfill a number of requirements. Among them, the most important are: (i) the optimization of the steady state operating conditions according to precisely quantified economic criteria, (ii) the possibility to operate with flexibility over the whole operating range with high rates of load variations, (iii) efficient dynamic control strategies explicitly coping with the constraints imposed by technological limits and by environmental restrictions, (iv) systematic procedures for the design and documentation of the control system.

Due to its almost unique features, which fully comply with all these requirements, model predictive control (MPC), see e.g. the survey papers [1,2] and the books [3,4], has extensively been used during the past twenty years mostly in the chemical and the petrochemical industry, where nowadays it is unanimously considered as the proper approach to the control design, see Reference [5]. In spite of the many similarities between the problems of managing and controlling (petro)-chemical units and power generation plants, the application of MPC in the

---

*Correspondence to: Professor Lalo Magni, Dipartimento di Informatica e Sistemistica, Università di Pavia, via Ferrata 1, 27100 Pavia, Italy
†E-mail: lalo.magni@unipv.it

power industry has not yet been exploited in depth. Notable exceptions are the works reported in [6–9], where different control structures based on *ad hoc* MPC implementations have been considered to control a reduced order nonlinear plant model and in Reference [10], where a real-time laboratory simulator has been used in the testing phase. Experimental results concerning the application of MPC for control of the temperature and the pressure of the steam at the superheater outlet in a large coal-fired power plant have been reported in Reference [11].

This paper presents the results of a detailed feasibility study on the practical applicability of MPC to thermal power plants. The plant is simulated with ALTERLEGO [12,13], a simulation environment developed by CESI, Centro Elettrotecnico Sperimentale Italiano, and validated over the years with real plant data. In ALTERLEGO also the classical co-ordinated scheme used in real plants is duplicated; it is composed by a number of PI regulators, by nonlinear static elements and by a number of logic blocks used during critical operating conditions, such as the plant start-up, and for safety reasons. The simulator is coupled with DMCplus by Aspentech, one of the most widely used commercial software environments for MPC implementing the popular dynamic matrix control (DMC) algorithm [14]. The MPC algorithm is applied to substitute in the co-ordinated scheme a number of single-input, single-output PI regulators controlling the steam pressure, the superheater and reheater steam temperatures and the opening of the reheat attemperator valve. Conversely, the generated power is controlled as in the classical decentralized approach in view of its fast dynamics and for safety reasons. Then, after the introduction of MPC, the resulting control scheme can be viewed as a 'boiler following'. A number of results is reported to witness the many potential benefits of the proposed solution.

The paper is organized as follows. Section 2 is devoted to present the main characteristics of the plant under control, of the traditional co-ordinated scheme and the control structure considered in the MPC implementation. In Section 3 the DMC algorithm is briefly summarized together with the main characteristics of the DMCplus package. Section 4 deals with the identification of the linear model used in the synthesis of the MPC controller. The main tuning guidelines of the MPC regulator and the simulation results are presented and discussed in Section 5, while the conclusions of this research activity are drawn in Section 6.

## 2. PLANT MODEL AND REGULATION SCHEME

The modelled plant is a conventional once-through 140–320 MW multi-fuel power plant, reproducing one of the units of Rossano Calabro in Italy. A schematic diagram of the plant is shown in Figure 1. The main pressure must be maintained at 170 bar by acting on the fuel flow valve (FFV), while the air flow valve (AFV) regulates the percentage of oxygen in the exhausts. The feed water flow is driven by the pump (FW) into multiple heat-exchangers stages: the economizer (ECO), located in the last part of the boiler, is intended to improve the global plant efficiency by pre-heating the water flow coming from the condenser (COND); the evaporator (EVAP), situated in the flame zone, causes water to turn into steam by radiation; the superheater, composed by a radiation–convection (SH2) plus a convection (SH1) stage, feeds the high-pressure stage of the turbine (HP), superheating the saturated steam to 540°C; the turbine governor valve (TGV), which plays the role of the main actuator in the closed-loop speed regulation of the generator (GEN), controls the steam flow to the HP turbine; the reheater
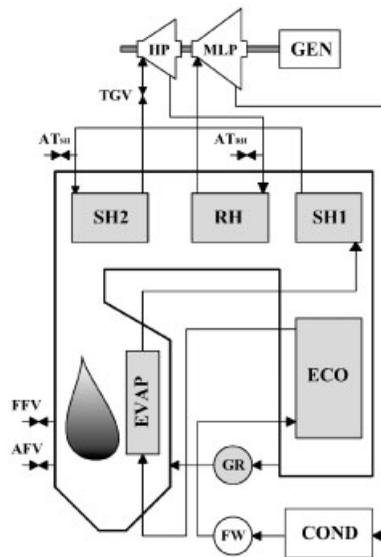
Figure 1. A schematic power plant diagram.

(RH) takes the exhaust steam from the HP stage back to $540°C$, and feeds the medium–low pressure turbine stage (MLP).

Two attemperators are provided to regulate the steam temperature: the first one (ATSH), intended to compensate the slow fuel dynamics by means of a faster regulation during transients, is maintained at a 50% opening valve value in steady state operations; the reheater attemperator (ATRH), instead, should be sparingly used, due to its negative influence on the boiler efficiency. The reheater regulation is also performed by means of the gas re-circulation (GR) damper, which modifies the combustion properties, thus allowing a balance action between radiation and convection.

The plant simulator is built with ALTERLEGO, a modular simulation tool based on module libraries of the main plant components and on an efficient numeric solver. ALTERLEGO has been widely used in many applications for the design and validation of the control schemes subsequently applied to the real plants as well as for the operators' training. In particular, the model adopted in this study has been tuned with data collected at the oil-fired unit of Rossano Calabro, in the south of Italy.

The 'traditional' regulation implemented in the simulator and used as a standard in many applications is the classical co-ordinated control scheme schematically depicted in Figure 2, where at the right-hand side of the dotted vertical line there is the power regulation scheme, not detailed here for simplicity. The signals involved are listed in Table I, while the meaning of the blocks in Figure 2 is explained in Table II.

The feedback regulators used in the scheme are mostly PI. Static nonlinear elements and logic switches are also used for performance enhancement and safety requirements. Finally, many loops are implemented according to a cascade structure. In particular, the PI regulator used to control the reheater steam temperature $T_{RH}$ produces the reference signal for the inner loop closed on the reheater attemperator.
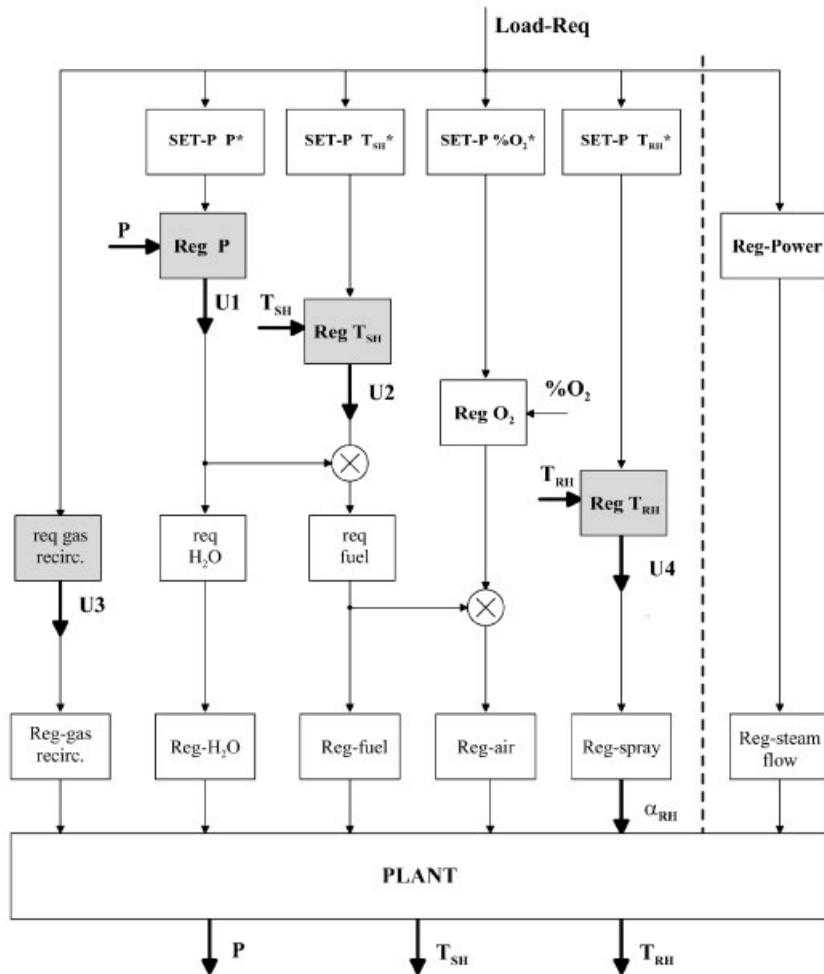
Figure 2. The traditional co-ordinated regulation.

In order to get a benefit from the application of MPC to the power plant control, reducing at the same time the impact of its introduction on the plant operators, in this study it has been decided to substitute only a limited number of traditional regulation blocks, taking into account the following considerations.

1. In thermal power plants, the main problems are related to the control of the boiler, due to the strong couplings between the manipulated ($U2$, $U3$, $U4$ in Figure 2) and the controlled ($T_{SH}$, $T_{RH}$, $\alpha_{RH}$) variables. For this reason, a multivariable centralized control scheme is advisable. Conversely, the steam turbine, and the related power regulation, is a simpler SISO system, characterized by a faster dynamics, where the reference is the load request, and the control variable is the steam flow rate; in this case it is not necessary to replace the traditional local PID regulator;

Table I. Variables list.

| Variable | Description |
|----------|-------------|
| $P$ | Main steam pressure |
| $T_{SH}$ | Superheater outlet steam temperature |
| $T_{RH}$ | Reheater outlet steam temperature |
| $\alpha_{RH}$ | Opening of the spray valve at the reheater |
| $\%O_2$ | Oxygen percentage in the exhausts |

Table II. Co-ordinated regulation blocks.

| Block | Description |
|-------|-------------|
| $SP - X$ | Set-point generation for variable $X$ |
| $Reg - X$ | Feedback regulator of variable $X$ |
| $req - X$ | Request generation for variable $X$ |

2. A decentralized control scheme is not the best choice to include the overall efficiency improvement in the control objectives. In particular, the use of the gas recirculation as control variable in a decentralized feedback structure is usually prevented by the typical non-minimal phase behaviour (inverse response) in the transfer function between the gas recirculation and the pressure ($P$). Then, only a feed-forward action is usually performed. This strategy is non-efficient and often causes an excessive use of the reheater attemperator, which negatively affects the boiler efficiency.

For all the above reasons, it has been a priori decided to substitute the PI's regulating the pressure $P$ and the temperature $T_{SH}$ (grey boxes in Figure 2) with the MPC control algorithm, which must also compute the two additional signals $U3$ and $U4$ acting on the gas recirculation request and on the reference signal for the reheater attemperator respectively. The goal is to achieve a tighter control action of $P$, $T_{SH}$ and $T_{RH}$ during dynamic transients caused by load variations and to compute the steady-state values, in particular the value of $\alpha_{RH}$, in order to enhance the plant efficiency and the life duration of the plant components. Note that in this new multivariable scheme, schematically represented in Figure 3 (where $P^*$, $T_{SH}^*$ and $T_{RH}^*$ are the set-point signals) the gas recirculation request acts in closed-loop. The resulting overall control scheme can be viewed as a 'boiler following' one, which can cause as a drawback fast and large pressure variations due to load changes, see Reference [10]. Then, one of the goals of this study is also to check whether the MPC approach can efficiently cope with this potential drawback.

## 3. MPC ALGORITHM AND TUNING GUIDELINES

The adopted algorithm is DMC (dynamic matrix control) [14], one of the most popular and widely used MPC methods, with hundreds of applications in the process industry, (see e.g. Reference [5]). For readers convenience, the main characteristics of algorithm are briefly
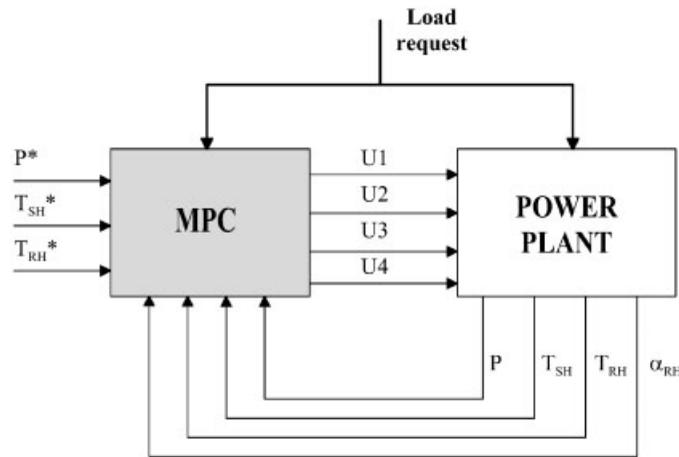
Figure 3. The multivariable control scheme.

reported in the sequel, with reference to SISO (single input, single output) systems. For a more detailed presentation see Reference [14].

In DMC, the plant under control, with input $u$ and output $y$, is asymptotically stable and described by the truncated finite impulse response (FIR) model

$$y(t) = \sum_{i=1}^{n} h_i u(t - i) + d(t)$$

where $t$ is the discrete-time index, $h_i$, $i = 1, \ldots, n$ is the $i$th impulse response coefficient, and $d(t)$ represents a disturbance signal. Letting $s_i$ be the $i$th step response coefficient ($h_i = s_i - s_{i-1}$) with $s_0 = 0$ and $\delta u(t) = u(t) - u(t - 1)$ be the control increment, the model can also be written as

$$y(t + k) = \sum_{i=1}^{k} s_i \delta u(t + k - i) + d(t + k) + \sum_{i=k+1}^{n-1} s_i \delta u(t + k - i) + s_n u(t + k - n)$$

At time $t$, the future value of the disturbance is assumed to be given by

$$d(t + k) = d(t) = y(t) - \sum_{i=1}^{n-1} s_i \delta u(t - i) - s_n u(t - n)$$

Then, the $k$ step ahead prediction of the output $y$ is

$$y(t + k) = \sum_{i=1}^{k} s_i \delta u(t + k - i) + f_k(t)$$

where the sum is made over the sequence of future control moves

$$\Delta U(t) = [\delta u(t)' \ \delta u(t + 1)' \ \ldots \ \delta u(t + M - 1)']$$

to be properly selected by the MPC algorithm along the control horizon $M > 0$, while the term

$$f_k(t) = y(t) + \sum_{i=k+1}^{n-1} s_i \delta u(t+k-i) + s_n u(t+k-n) - \sum_{i=1}^{n-1} s_i \delta u(t-i) - s_n u(t-n)$$

is known at time $t$.

The DMC control law is derived by minimizing, at any time $t$ and with respect to $\Delta U(t)$, the cost function

$$J = \sum_{i=t+1}^{t+N} (w(i) - y(i))' Q(w(i) - y(i)) + \delta u(i-1)' R \delta u(i-1) \tag{1}$$

where $N > 0$ represents the prediction horizon, while $Q$ and $R$ are positive definite matrices and $w$ is the reference signal. The minimization is performed under the following constraints

$$|u(t+i)| \leqslant u_m \quad i = 0, \ldots, M-1 \text{ (hard constraints)} \tag{2}$$

$$\delta u(t+M+i) = 0 \quad i \geqslant 0 \text{ (hard constraints)} \tag{3}$$

$$|w(t+i) - y(t+i)| \leqslant \varepsilon_m \quad i = 1, \ldots, N \text{ (soft constraints)} \tag{4}$$

Once the optimization problem (1)–(4) has been solved, and according to the Receding Horizon principle, only the first term $\delta u^0(t)$ of the computed optimal sequence $\Delta U^0(t)$ is applied, while the whole procedure is repeated at the next sampling time $t + 1$.

Besides implementing the above dynamic control algorithm also for multi-input, multi-output plants, the software package DMCplus used in this study allows to estimate the step response coefficients $s_i$. The identified models are preliminarily used for the tuning and the testing of the MPC regulator. Once a good controller tuning has been achieved, the resulting regulator is subsequently applied in real-time operations.

Among the main features of the method which have been proven to be crucial in this study, the following ones have to be recalled.

1. The definition of priorities in the constraints fulfillment, by ranking the manipulated and controlled variables and, in case of unfeasibility, by violating the 'lower priority' constraints. This feature has been used to specify the following priorities: $\alpha_{RH}$, $P$, $T_{SH}$, $T_{RH}$ for the lower limits, $P$, $T_{SH}$, $T_{RH}$, $\alpha_{RH}$ for the upper limits (the higher priority has been assigned to the lower limit of $\alpha_{RH}$, which constitutes a true 'hard constraint'; the pressure $P$ follows, by virtue of its direct link with the generated power, with a working range of $\pm 5$ bar; finally, the operating limits $\pm 15^\circ$C have been set for the two temperatures $T_{SH}$ and $T_{RH}$, to prevent excessive stress of the steam turbine, without penalizing the performances during transients).

2. In the computation of the dynamic control action, the penalties (matrices $Q$ and $R$ in the performance index (1)) on the future error variables and control increments can be made to explicitly depend on the sign and amplitude of the error components. In the tuning procedure [4], the diagonal values of matrix $Q$ have been preliminary set to the inverse of the squared nominal values of the corresponding controlled variables. Then an asymmetric weaker weight to the reheater attemperator $\alpha_{RH}$ has been used to allow for the full usability

of the spray valve opening during transients, while it has been strongly penalized in steady-state optimization as specified in the following. As for the elements of matrix $R$, the main temperature control variable, i.e. the fuel pilot, has been slightly more penalized to prevent too high temperature gradients.

3. The possibility to specify different modes of operation, i.e. the tracking of a fixed constant set-point, or the fulfillment of oscillation bands. Specifically, the variables $P$, $T_{SH}$ and $T_{RH}$ have been controlled in tracking with set-point values $P^0 = 170$ bar, $T_{SH}^0 = 540°C$ and $T_{RH}^0 = 540°C$; while $\alpha_{RH}$ has been forced to remain in a prescribed band with a minimum value of $3°C$.

4. The use of feed-forward and disturbance signals, which has been useful to force an instantaneous control movement at any new load variation.

5. The possibility to include explicit constraints on the future variations of the control variables. This has been crucial to limit the variations of $U3$, so as to reduce the oscillations induced by an excessive use of the gas recirculation.

6. The identified model can be also used to compute the optimal steady state working point by solving a linear programming problem defined according to economic and safety criteria specified by the user, and under constraints on the values of the manipulated and controlled variables. Specifically, this feature has been considered to include in the optimization problem the goal to maintain, in steady-state conditions, the reheater spray valve as close as possible.

## 4. IDENTIFICATION PROCEDURE

It is well known that thermal power units are highly nonlinear plants, with gains and time constants which monotonically depend on the load, see Prasad *et al.* [8]. In order to assess the extent of the plant nonlinearities, to obtain an initial rough estimation of the dominant time constants involved, and to examine the plant cross couplings, a preliminary set of open-loop experiments has been performed. Specifically, the PI regulators Reg $- P$, Reg $- T_{SH}$ and Reg $- T_{RH}$, and the gas recirculation open-loop request in Figure 2 have been disconnected and step changes have been imposed to the manipulated variables $U1$–$U4$ starting from a steady state condition corresponding to the operating point of 240 MW.

Some of the results achieved are reported in Figure 4. This figure shows that the single transfer functions exhibit a quite complex behavior, with time constants of different orders of magnitude, and with some inverse responses which make the control problem difficult. From these results, it has been computed the following plant relative gain array (RGA) matrix [15].

The RGA matrix shows that there are strong coupling between the control variables $U2$, $U3$, $U4$ and the controlled variables $T_{SH}$, $T_{RH}$ and $\alpha_{RH}$. Moreover, the use of $U2$ for control of $T_{SH}$ as in the decentralized scheme of Figure 2 causes significant problems (the corresponding element in RGA is negative). On the contrary, the pressure $P$ in steady state conditions mainly depends on the control variable $U1$, however it has been decided to include $P$ between the variables controlled by MPC, so as to regulate the main plant variables according to a 'boiler following' strategy.

The extent of the plant nonlinearity can be appreciated with Figures 5 and 6, which show the transients of the controlled variables $P$, $T_{SH}$ and $T_{RH}$ in front of a step variation of amplitude 1% of the control variable $U2$ starting from different operating points. From this figure it is
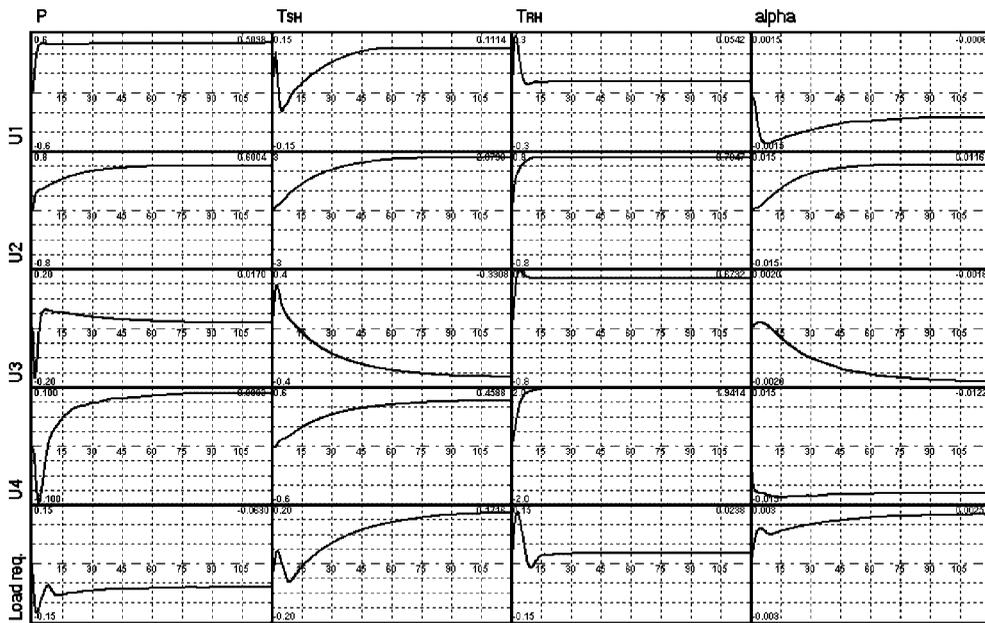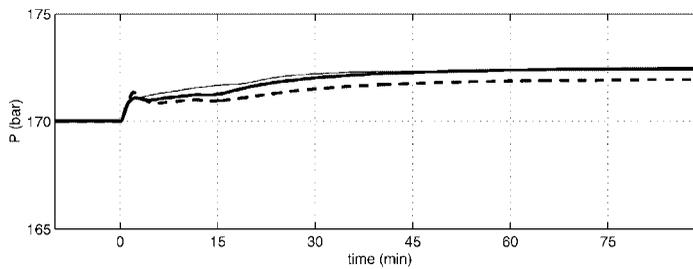
Figure 4. Plant step responses.



Figure 5. Response of $P$ to a step variation in $U2$. Operating points: thin line = 300 MW, bold line = 220 MW, dashed line = 140 MW.

apparent that the plant is much slower at low loads than at full load. Moreover, the plant gains vary significantly in the operating range. The same conclusion could be drawn by examining the other links between the control and controlled variables, and are in perfect agreement with the results reported in Reference [8].

From Figure 4, it can be seen that at 240 MW, the dominant time constants range from a minimum of about 90 s for the pressure to a maximum of about 1250 s for the superheat steam temperature. In view of the 'fast' pressure dynamics, the sampling time used in identification and control has been set equal to 15 s and the identified step response, single-input, single-output,
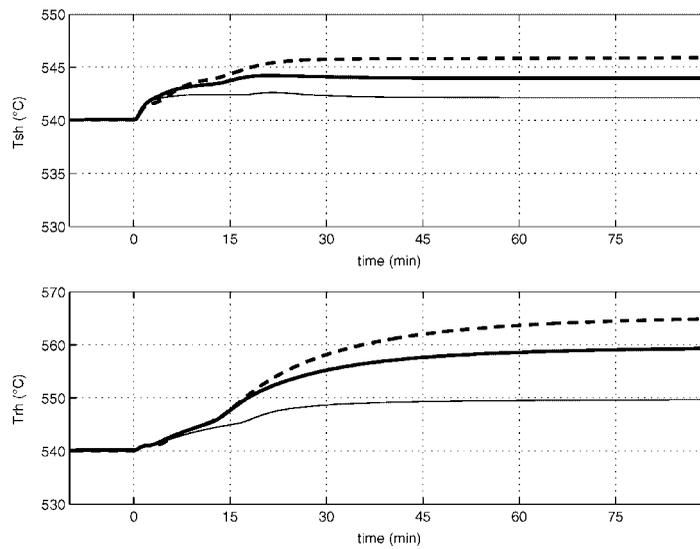
Figure 6. Response of superheater and reheater temperatures $T_{SH}$ and $T_{RH}$ to a step variation in $U2$. Operating points: thin line = 300 MW, bold line = 220 MW, dashed line = 140 MW.

FIR models have been composed by 480 coefficients, which is sufficient to consider the settling time of all the controlled variables. These models have been identified by using as input variables the signals $U1$–$U4$ obtained combining step variations, for a good identification of the static gains, and PRBS (pseudo random binary signals), necessary to excite the system in the frequency range of interest. The amplitude of the PRBS has been limited to 5–10% of the nominal values of the corresponding signals, which represents a feasible perturbation also for real plants. Other characteristics have been selected according to the guidelines given in Reference [16].

## 5. EXPERIMENTAL RESULTS

Many simulation experiments have been performed to assess the performance of the MPC algorithm. Among them, in the sequel only three sets are reported for conciseness. The first one refers to ramp variations from 140 to 320 MW, and it is here discussed to illustrate the ability of the MPC algorithm to cover all the operating range in spite of the plant nonlinearities. The second set of experiments shows the performance of the predictive controller in front of a 40 MW step-like variation of the load, which represents a critical test for the control algorithm. Indeed, it is important to check the performance of MPC with respect to these perturbations in view of the higher and higher rates of load variations required by the market. Finally, the third set of simulations presents the results achieved by forcing a profile of load variations recorded on a real plant during a period of 4 days. This is useful to draw some conclusions on the 'average' performance of MPC in standard

operating conditions and to compare them to that provided by the classical co-ordinated control scheme.

### 5.1. Experiment 1

Starting from the operating regime corresponding to the minimum load (140 MW), the load request is brought to the maximum value of 320 MW at a rate of 3.2 MW/min. Correspondingly, the transients of the main control and controlled variables are shown in Figures 7–9.

From these figures, it is apparent that MPC produces a much tighter control action than the co-ordinated scheme, as also witnessed by the value of the root mean square (RMS) error of $P$, $T_{SH}$, $T_{RH}$ computed for the two algorithms and reported in Table III. This Table also shows the mean value of $\alpha_{RH}$ and of the gas recirculation request $U3$. It is apparent (see also Figure 9) that in steady state conditions MPC maintains more closed the spray attemperator valve by means of a better use of the gas recirculation (by virtue of the closed loop regulation of this variable, not implemented in the traditional control scheme), so increasing the plant thermal efficiency. On the contrary, during the transient caused by a load variation, the valve is used to achieve a better and faster control action.

### 5.2. Experiment 2

A load step variation of 40 MW has been imposed starting again from the operating point of 240 MW. The obtained results are reported in Figures 10–12, while the RMS and mean values of the main variables are summarized in Table IV.
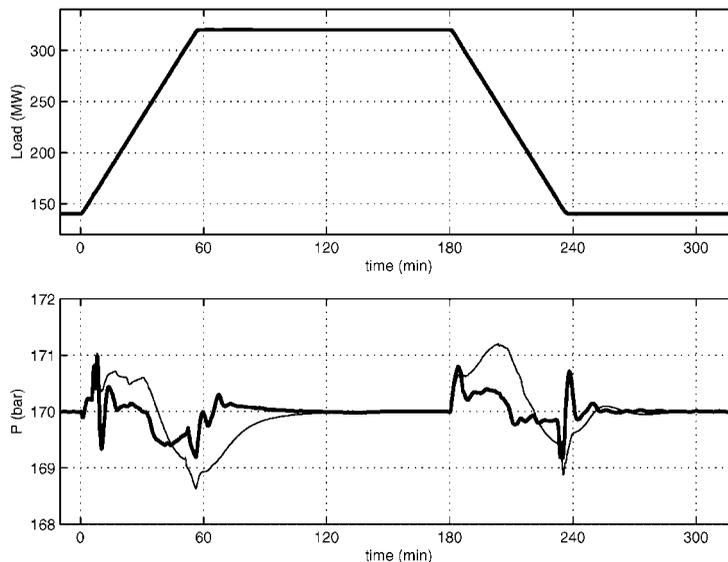


Figure 7. Experiment 1: load request and steam main pressure ($P$) with the traditional co-ordinated control scheme (thin line) and the MPC regulator (bold line).
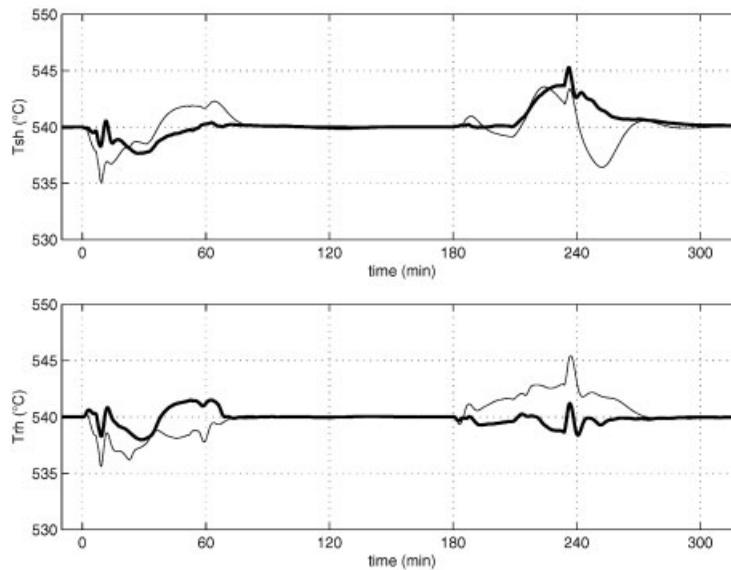
Figure 8. Experiment 1: superheater ($T_{SH}$) and reheater ($T_{RH}$) temperatures with the traditional co-ordinated control scheme (thin line) and the MPC regulator (bold line).
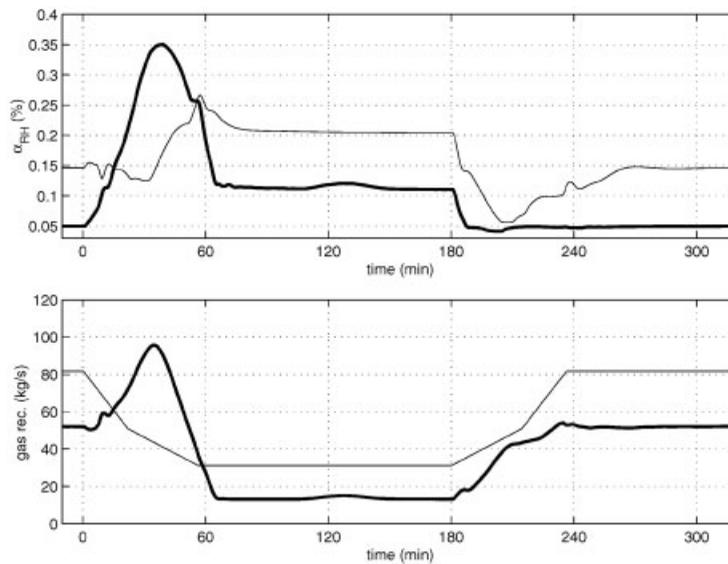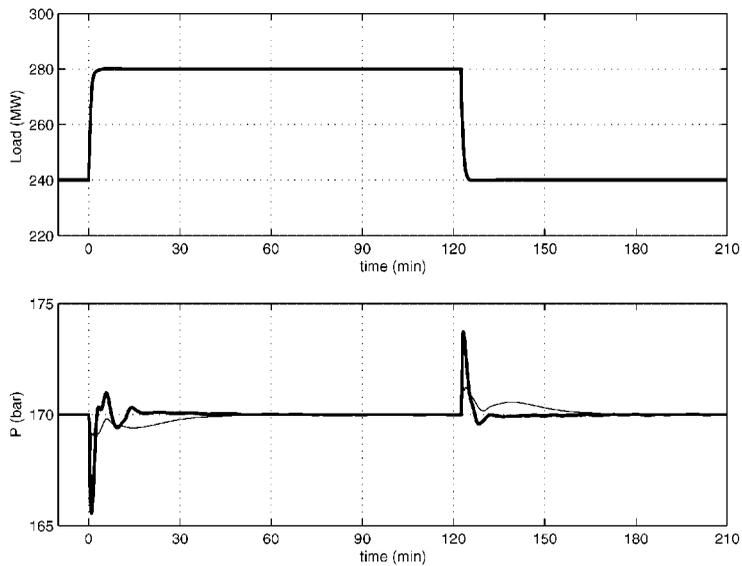


Figure 9. Experiment 1: reheater spray valve opening ($\alpha_{RH}$) and gas recirculation flow with the traditional co-ordinated control scheme (thin line) and the MPC regulator (bold line).

Table III. Performance of the controllers in Experiment 1.

| | RMS | | Mean value | |
|---|---|---|---|---|
| | MPC | Co-ordinated | MPC | Co-ordinated |
| $P$ | 0.222 bar | 0.452 bar | | |
| $T_{SH}$ | 1.136°C | 1.442°C | | |
| $T_{RH}$ | 0.597°C | 1.520°C | | |
| $\alpha_{RH}$ | | | 10.1% | 16.2% |
| Flow gas recirculation | | | 39.281 kg/s | 54.745 kg/s |



Figure 10. Experiment 2: load and steam main pressure ($P$) with the traditional co-ordinated control scheme (thin line) and the MPC regulator (bold line).

In this case, it can be noted that the pressure $P$ exhibits a larger undershoot with MPC then with the classical control scheme of Figure 2, as expected in view of the 'boiler following' nature of the MPC implementation. However, its variation of about 3 bar is largely inside the prescribed band of $\pm 5$ bar. Conversely, the regulation of the temperatures is significantly improved as well as the use of the reheater spray valve and of the gas recirculation.

### 5.3. Experiment 3

The recorded transients of the load request registered along a period of 4 days in a real plant have been used to test the performance of MPC in real operating conditions. The results reported in Figures 13–18 and in Table V confirm the conclusions drawn from Experiments 1
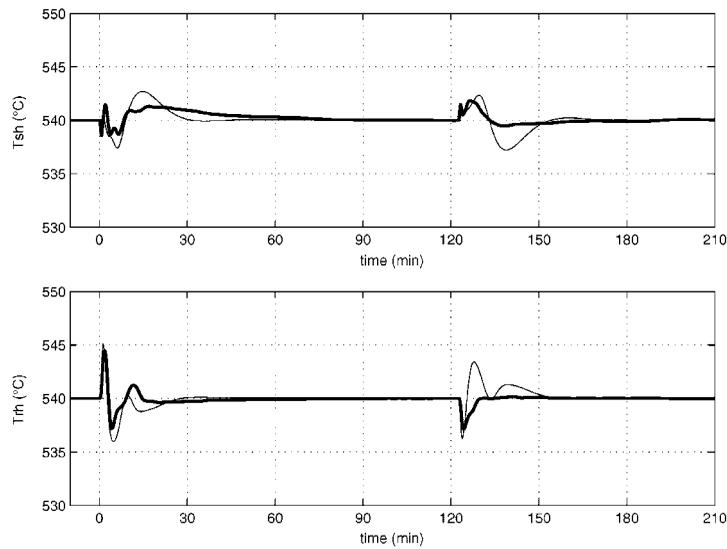
Figure 11. Experiment 2: superheater ($T_{SH}$) and reheater ($T_{RH}$) temperatures with the traditional co-ordinated control scheme (thin line) and the MPC regulator (bold line).
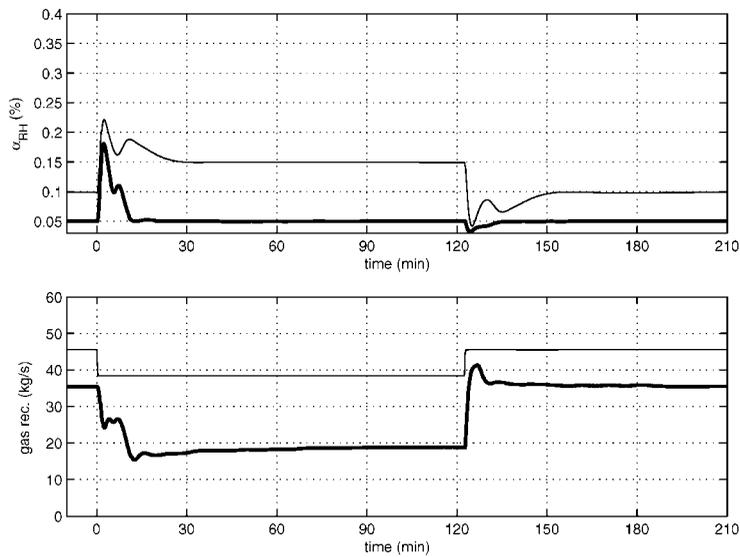


Figure 12. Experiment 2: reheater spray valve opening ($\alpha_{RH}$) and recircle gas flow with the traditional co-ordinated control scheme (thin line) and the MPC regulator (bold line).

Table IV. Performance of the controllers in Experiment 2.

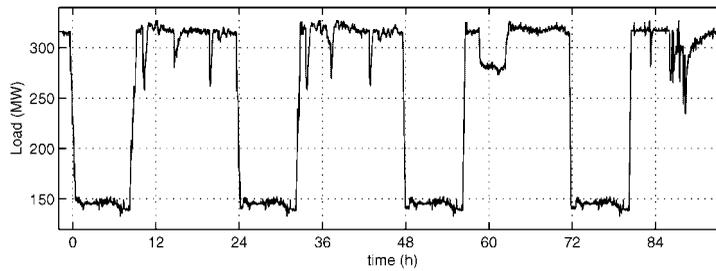| | RMS | | Mean value | |
|---|---|---|---|---|
| | MPC | Co-ordinated | MPC | Co-ordinated |
| $P$ | 0.477 bar | 0.319 bar | | |
| $T_{SH}$ | 0.589°C | 1.066°C | | |
| $T_{RH}$ | 0.618°C | 0.924°C | | |
| $\alpha_{RH}$ | | | 5.50% | 14.2% |
| Flow gas recirculation | | | 23.031 kg/s | 40.174 kg/s |



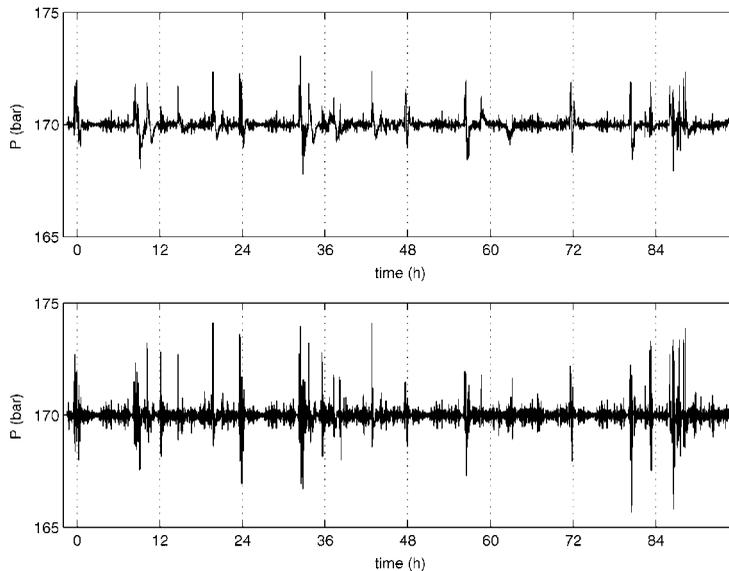Figure 13. Experiment 3: load during the long-period test.



Figure 14. Experiment 3: main pressure ($P$) with the traditional co-ordinated control scheme (upper graph) and the MPC regulator (lower graph).
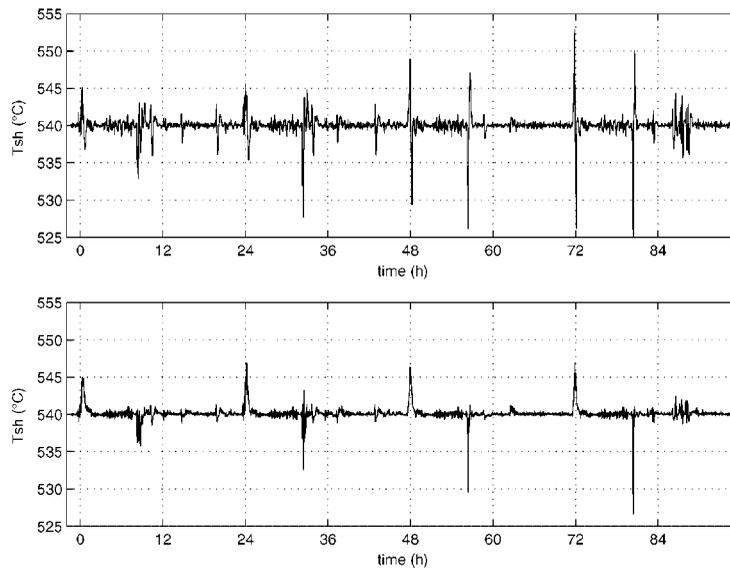
Figure 15. Experiment 3: superheater ($T_{SH}$) temperature with the traditional co-ordinated control scheme (upper graph) and the MPC regulator (lower graph).

Table V. Performance of the controllers in Experiment 3.

|  | RMS | | Mean value | |
|---|---|---|---|---|
|  | MPC | Co-ordinated | MPC | Co-ordinated |
| $P$ | 0.420 bar | 0.316 bar | | |
| $T_{SH}$ | 0.790°C | 1.448°C | | |
| $T_{RH}$ | 0.802°C | 1.285°C | | |
| $\alpha_{RH}$ | | | 8.79% | 17.20% |
| Flow gas recirculation | | | 29.058 kg/s | 48.796 kg/s |

and 2. Specifically, the regulation of the pressure provided by MPC is worse than that of the traditional decentralized scheme, although the variations with respect to the set-point value are largely inside the prescribed band. Conversely, the steam temperatures $T_{SH}$ and $T_{RH}$ are controlled much better and the use of the reheater spray valve and of the gas recirculation is optimized.

In this experiment, the specific fuel consumption has also been computed, it is 14.2864 kg/s for the co-ordinated scheme and of 14.2530 kg/s for the MPC regulator; this reduction is equivalent to a considerable saving of money. Moreover, on the basis of a qualitative analysis, it is possible to appreciate a pollutants decrease, directly consequent on the fuel consumption reduction (Figures 16–18).
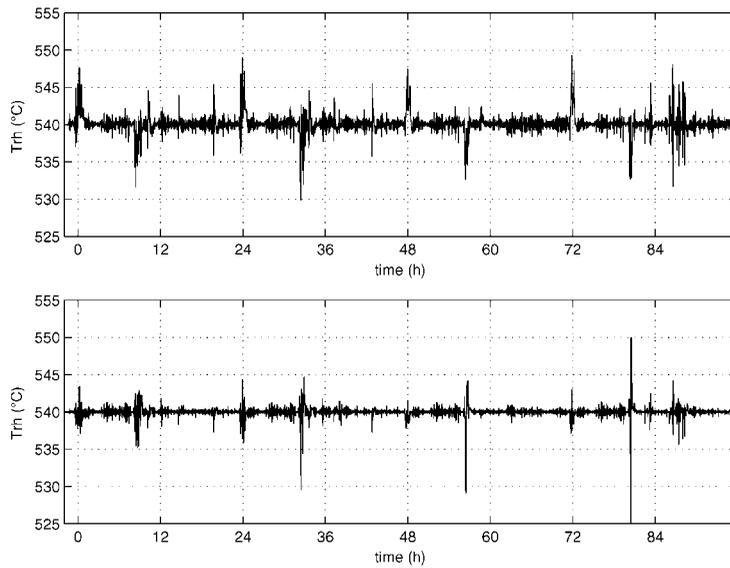
Figure 16. Experiment 3: reheater ($T_{RH}$) temperature with the traditional co-ordinated control scheme (upper graph) and the MPC regulator (lower graph).
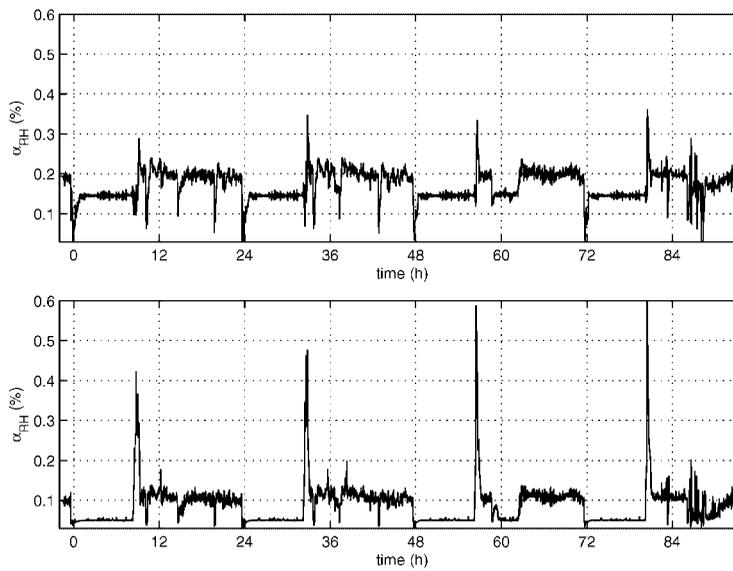


Figure 17. Experiment 3: reheater spray valve opening ($\alpha_{RH}$) with the traditional co-ordinated control scheme (upper graph) and the MPC regulator (lower graph).
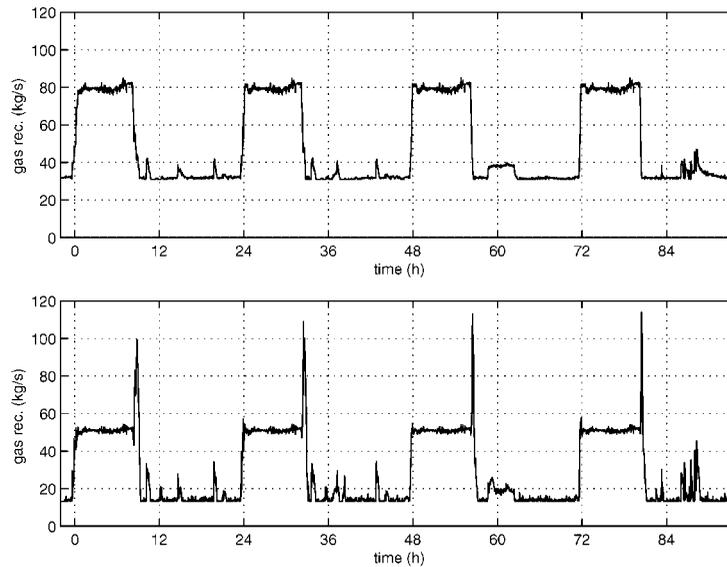
Figure 18. Experiment 3: recircle gas flow with the traditional co-ordinated control scheme (upper graph) and the MPC regulator (lower graph).

## 6. BENEFITS FOR INDUSTRIAL APPLICATIONS AND CONCLUSIONS

This study allows to draw the following conclusions on the use of MPC for control of thermal power plants.

1. MPC produces a significant increase of the overall thermal efficiency. This is achieved with a more efficient management of the reheater spray valve and of the gas recirculation. In turn, this improved efficiency leads to a reduction in the fuel consumption and to significant economic savings. Moreover, a reduced fuel consumption also means less pollutants in the atmosphere: this is of major importance in view of the more and more stringent legal requirements on the pollutants emissions;
2. The improved regulation of the steam temperatures $T_{SH}$ and $T_{RH}$ produces minor thermal stresses of the materials, which means reduced costs for the maintenance and supervision of the plant. Moreover, this tighter control action could allow for higher set-points for $T_{SH}$, $T_{RH}$, with an improvement of the plant thermal efficiency;
3. Thanks to the better overall regulation, it is possible to work on a wider operating range and to considerably increase the flexibility in the management of the production units; in turn, this allows to quickly change the operating conditions and, consequently, to have access to dynamic services;
4. The 'optimal' regime can be computed by minimizing an economic criterion which can be frequently modified to cope with changing external conditions;
5. The synthesis procedure is easier and better documented than in the case of co-ordinated decentralized control schemes. This has a great importance in the management of power plant units.

Finally, it has to be noted that the MPC algorithm has been used to substitute only a part of the classical co-ordinated control scheme, so that further advantages could be achieved by using it as the unique high level controller generating the reference signals for lower level traditional single-input single-output regulators.

## ACKNOWLEDGEMENTS

## REFERENCES

1. Garcia CE, Prett DM, Morari M. Model predictive control: theory and practice—a survey. *Automatica* 1989; **25**:335–348.
2. Mayne DQ, Rawlings JB, Rao CV, Scokaert POM. Constrained model predictive control: Stability and optimality. *Automatica* 2000; **36**:789–814.
3. Camacho EF, Bordons C. *Model Predictive Control in The Process Industry*. Springer: Berlin, 1999.
4. Maciejowski J. *Predictive Control with Constraints*. Prentice-Hall: Englewood Cliffs, NJ, 2001.
5. Qin SJ, Badgwell TA. An overview of industrial model predictive control technology. *Fifth International Conference on Chemical Process Control*, Kantor JC, Garcia CE, Carnahan B (eds). 1996; 232–256.
6. Rovnak JA, Corlis R. Dynamic matrix control for fossil power plants. *IEEE Transactions on Energy Conversion* 1991; **6**(2):320–326.
7. Prasad G, Swidenbank E, Hogg BW. A neural net model-based multivariable long-range predictive control strategy applied in thermal power plant control. *IEEE Transactions on Energy Conversion* 1997; **13**(2):176–182.
8. Prasad G, Swidenbank E, Hogg BW. A local model networks based multivariable long-range predictive control strategy for thermal power plants. *Automatica* 1998; **34**(10):1185–1204.
9. Prasad G, Irwin GW, Swidenbank E, Hogg BW. Plant-wide predictive control for a thermal power plant based on a physical plant model. *IEEE Proceedings—Control Theory Applications* 2000; **147**(5):523–537.
10. Lu S, Hogg BW. Predictive co-ordinated control for power-plant steam pressure and power output. *Control Engineering Practice* 1997; **5**(1):79–84.
11. Oluwande G, Boucher AR. Implementation of a multivariable model-based predictive controller for superheater steam temperature and pressure control on a large coal-fired power plant. *European Control Conference*, Karlsruhe, 1999.
12. Castiglioni L, De Chirico M, Pretolani F, Spelta S. A real time simulator implementation in the unix environment. *BIAS 25th edition*, Milano, 1993.
13. Cori R, Spelta S, Guagliardi G, Pretolani F, Maltagliati P, Persico F, Sommani M. The legocad system: a computer aided power plant modelling environment (in italian). *Proceedings of XC National Symposium AEI*, Lecce, 1989.
14. Cutler CR, Ramaker BC. Dynamic matrix control—a computer control algorithm. *Proceedings Joint Automatic Control Conference*, San Francisco, CA, 1980.
15. Bristol EH. On a new measure of interactions for multivariable process control. *IEEE Transactions on Automatic Control* 1966; **11**:161–166.
16. Rivera DE, Jun KS. An integrated identification and control design methodology for multivariable process system application. *IEEE Control Systems Magazine* 2000; **20**:25–37.