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Performance Improvements of Cascade and Feed-Forward Control Schemes for Industrial Processes

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In this paper a comparison among different control schemes, frequently implemented in industrial processes is presented. These control structures range from simple configurations to complex structures, in order to reproduce problems to be faced in the choice of the more suitable control system of a real industrial process, a spray dryer, which presents significant problem of quality control in the presence of external disturbance. In particular, feedback control, feedback plus feedforward control, cascade control, cascade plus feedforward, and double-feedback plus feedforward are analyzed and compared. The control structures are evaluated on the basis of different criteria, including structural simplicity and ease of realization, performance indices (as maximum allowable deviations on controlled variable, integral of the absolute value of control error, total variation of control action) and constraints on manipulated variables. As final result, a cascade plus feedforward configuration is indicated as more suitable for the assigned specifications and its robustness to errors in the model and variations in the operating conditions is accounted for.

1. Introduction

Most of industrial plants are multivariable in their nature, as many input variables influence operations and many output variables can be chosen as performance indices. Very often they can be reduced to a series of simpler subsystems, where a main variable can be chosen to control product specifications: a necessary condition is real time measurability and sensitivity to most significant input perturbations. In these cases, conceptually simple control scheme, as cascade and feedforward control, combined with basic feedback control, can give acceptable performance in the desired operating conditions (Bacci di Capaci et al., 2015).
Basic aspects are treated in classical textbooks (Visioli, 2006), but the optimization of two degrees of freedom control structures still receive attention in specific literature (Kaya et al., 2007; Alfaro et al., 2009; Jesus et al., 2013; Marchetti et al., 2013). Very often, with the change in operating conditions (production rate, required quality of products), an optimal control system can become inadequate; the same problem may arise in the presence of specific perturbations affecting the process. In this situation the alternative is between a frequent design of control system components, for instance by adopting autotuning techniques (see for instance Leva and Marinelli, 2009), or a more general design able to maintain acceptable performance in the range of planned operating conditions. An example of this second approach is presented in this paper, referring to an industrial drying plant.

2. Modelling the process

The industrial process taken into consideration is a drying plant for the production of amorphous silica, which has several different commercial uses. Firstly, the raw material undergoes different chemical and physical processes to obtain a solution of sodium silicate. Then, through further chemical attacks, filtration and liquefaction of the solid material, a silica slurry is obtained, with water content higher than 70 % by weight, to be dried in a *spray dryer* unity. The drying section of the plant consists of three main elements (Figure 1, left): burner, dryer, and filter. The high-temperature gases from the burner meet and dry the slurry of silica in the spray dryer. Then, the flue gases and the dried product are separated trough a bag filter. Finally, the product is stored and packaged for transport, while flue gases are sent to chimney.

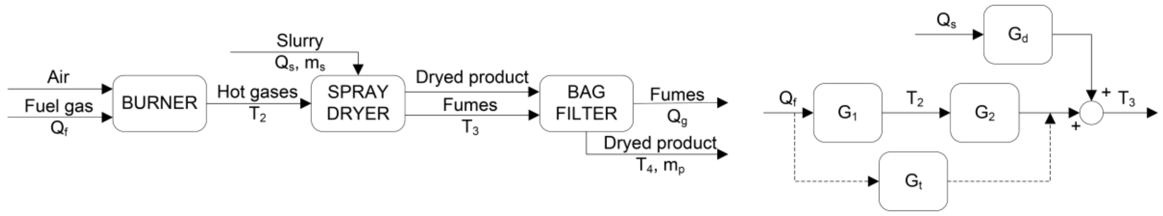

Figure 1: left) block diagram of the drying section; right) simplified diagram of the linear model dynamics.

All information about plant design and operations, main disturbances and acquired experience from operators are available. For confidentiality reasons only few details and numbers are given. Plant operations and the control problem are synthesized in the sequel.
The available measurements are the output temperature from the burner (T_2), the output temperature from the dryer (T_3), the output temperature from the filter (T_4), the input flow rate of fuel gas to the burner (Q_f, in Nm³/h), the input flow rate of slurry to the dryer (Q_s, in m³/h), and the output flow rate of flue gas form the filter (Q_g, in Nm³/h). The input air flow rate is not measured, while the initial water content of the slurry (m_s) and the residual moisture level of the dried product (m_p) are sampled off-line every 2/3 hours.
The burner heats air and fuel gas, basically methane, to suitable temperature for the dryer; air is in large excess in order to have not too high temperature and large gas flow rate for the drying process. A fan maintains light vacuum conditions in the plant and allows a global flow rate of fumes, almost constant. The silica slurry is shot in the dryer through a set of atomising lances, not always all simultaneously in operation, but activated depending on the production needs. The jet of the lances is directed upward, so that drops of water, as reached by the high-temperature hot gases, instantaneously evaporate, and the temperature of the gas mixture decreases ($T_3 < T_2$). The slurry has a solid component in suspension which cause a high abrasive action and a progressive increase of the nozzles' section. When the single flow rate is over 30% of the nominal flow, the replacement of the lance is required. This operation has to be seen as the most important and unavoidable perturbation affecting the dryer, several times every day. Other common perturbations are variations on fuel gas composition and on the input temperature to the burner.
The plant obtains various types of products for different end uses, which differ for the level of residual moisture and for the size of micro beads of silica. The specifications on the residual moisture affect the fuel gas flow rate and the water amount removed from the slurry, the fuel gas flow rate being proportional to the amount to evaporate. Apart from product properties correlated with upstream operations, the residual moisture of the silica is the variable to be controlled in this stage. An indirect moisture control is performed by means of the flue gas output temperature T_3: out-specs on m_p occur in case of persistent deviations, and variations of the set-point of T_3 are imposed to suppress them.
A first stage of process identification has involved simple linear models and it was based on a mixed approach, where some parameters are obtained from routine data and other via analytic approach, through material and energy balances. Each element of the drying process is described by a transfer function of first order plus time-delay (FOPTD), which includes only three parameters: static gain, time constant, and time-delay. Anyway, note that the actual process dynamics is highly nonlinear, since for different operation conditions, e.g., number of active atomising lances, and configuration of auxiliary flows, various model parameters can be identified. On the basis of the analytical modeling and some step-tests on the fuel gas flow rate, three main process dynamics (G_1, G_2, G_d) are established, as shown in Figure 1, right. The process G_2 from T_2 to T_3 could not be identified from routine data, and was estimated as the ratio G_t/G_1, where G_t is the dynamics of global process from Q_f to T_3 modelled as FOPTD, thus obtaining a lead/lag dynamics with one negative pole and one negative zero (see Table 1). The gain of G_d equal to -14 °C/(m³/h) proves that the variation of flow rate of slurry has the most significant effect on T_3, which makes Q_s the main process disturbance.

Table 1: Identified process parameters

Process	Input	Output	Gain K	Time constant τ [s]	Time-delay θ [s]
G_1	Q_f	T_2	0.5 °C/(Nm³/h)	25	5
G_t	Q_f	T_3	0.15 °C/(Nm³/h)	400	30
G_2 (= G_t/G_1)	T_2	T_3	0.3 °C/°C	25, 400	25
G_d	Q_s	T_3	-14 °C/(m³/h)	90	20

3. Control schemes

As illustrated before, acquired knowledge on the process shows that outlet temperature from the spray dryer is the main variable to control. Maintaining values of T_3 in a narrow range around set-point values (3°C) is assumed as necessary condition to guarantee the right content of moisture in the product m_p.
Figure 2 show the diagram for the process control. By sampling T_4 or by the off-line monitoring of m_p, the set-point of T_3 can be varied. As the set-point is increased, the set-point of T_2 is increased, thus increasing the flow rate of fuel gas.

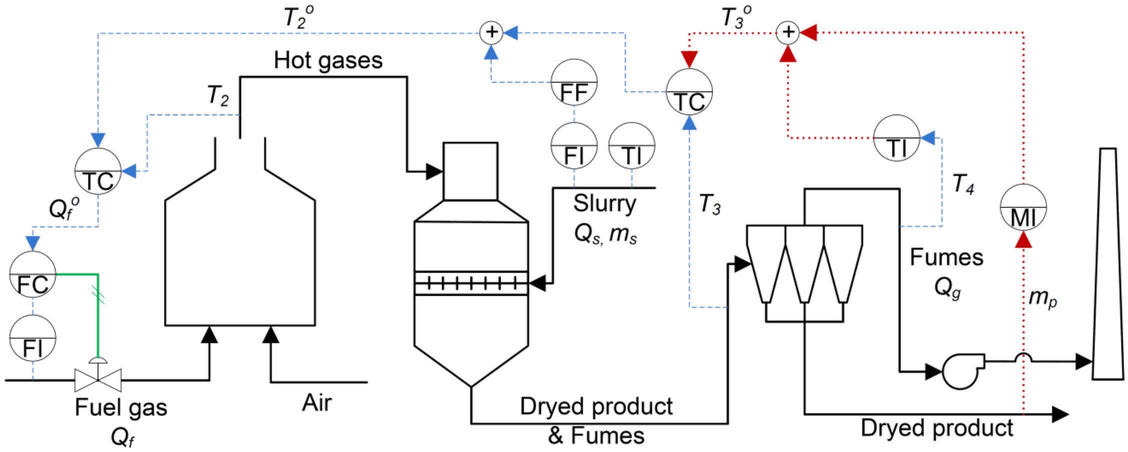

Figure 2: Representation of plant control schemes.

Five different control configurations are analyzed and compared: feedback control (FB), feedback plus feedforward control (FB + FF), cascade control (CC), cascade plus feedforward (CC + FF), and double-feedback plus feedforward (2FB + FF). Figure 3 shows the most complex configuration: cascade plus feed-forward. The other schemes can be obtained by opening or moving the various control loops. For example, in the case of 2FB + FF, the feedforward controller (C_{FF}) acts directly on the fuel gas (dashed line), and not between the two controllers (C_1 and C_2) as for the standard cascade. Also the fuel gas features a cascade control to get rid of disturbance affecting the feed line; for the sake of simplicity, due to its very fast dynamics, this internal control loop is neglected in the design of the control schemes. The installation of a online moisture sensor and a further controller acting on set-point of T_3 would constitute a sensible advantage; note that the control of T_3 will maintain a key role assuring product specifications. In addition, an online sensor of flue gas humidity (standard practice) would allow the development of a virtual sensor to use in conjunction with the off-line measurements of moisture in the slurry (m_s) and in the dried product (m_p).
Table 2 shows the algorithms adopted for the feedforward and feedback components of the considered control schemes. Note that in CC + FF and 2FB + FF schemes, design and algorithms of the FF component are slightly involved, by including internal loop elements (C_1 and G_1); thus, they have to be redesigned every time that CC is modified. Therefore, CC + FF^s and 2FB + FF^s are proposed as simplified versions of CC + FF and 2FB + FF, respectively. They have the same structure of the corresponding exact version, but the feedforward controller is designed as for the simple FF + FB configuration ($C_{FF}{}^0$). In the case of simple FB and FB + FF scheme, the single controller C (placed as C_2) has PI algorithm, tuned on the global process G_t, according to the Curve Response (CR) method (Brambilla et al., 1990). For the schemes with CC component, the internal controller of PI-type is tuned on the internal process $C_1 = f(G_1)$, according to CR technique, while the external controller, still of PI-type, is tuned on the dynamics of complex model G*, with the method of Ziegler-Nichols.

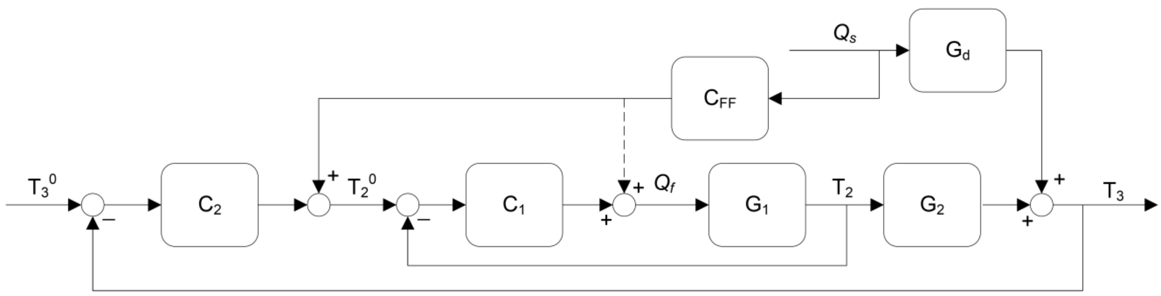

Figure 3: Cascade plus feed-forward configuration.

Table 2: Algorithms and design techniques for feedforward and feedback controllers

Control scheme	FF algorithm	C_1 algorithm	C_2 algorithm
FB	-	$C = f(G_t)$	
FB + FF	$C_{FF^o} = -\frac{G_d}{G_t} = -\frac{K_d}{K}\frac{\tau s+1}{\tau_d s+1}e^{-(\theta_d-\theta)}$	$C = f(G_t)$	
CC	-	$C_1 = f(G_1)$	$C_2 = f(G^*)$
CC + FF	$-\frac{G_d}{G^*} = -\frac{G_d}{\frac{C_1 \cdot G_1 G_2}{1+C_1 \cdot G_1}} = C_{FF^o}\cdot\frac{(1+C_1\cdot G_1)}{C_1}$	$C_1 = f(G_1)$	$C_2 = f(G^*)$
2FB + FF	$C_{FF^o}\cdot(1+C_1\cdot G_1)$	$C_1 = f(G_1)$	$C_2 = f(G^*)$
CC + FF^s	C_{FF^o}	$C_1 = f(G_1)$	$C_2 = f(G^*)$
2FB + FF^s	C_{FF^o}	$C_1 = f(G_1)$	$C_2 = f(G^*)$

Note that ideal formulations of feedforward components are physically realizable (PR) only when disturbance time-delay is larger than process time-delay, that is, when $\delta = \theta_d - \theta > 0$. Otherwise, when disturbance is faster ($\theta_d < \theta$), time-delay elements cannot be included, and only dynamic (or static) formulations can be designed. Note that this second scenario is closer to the plant reality, since θ_d is related to the fast propagation of the cooling effect inside the spray dryer because of the water evaporation, while θ depends on the hot gas flow rate and mostly on the pipe length between the burner and the spray dryer. Therefore, control system performance would benefit by a redesign of the pipe connecting the burner to the sprayer.

4. Simulation analysis

The additional opening of an atomizing lance in the spray dryer can be well simulated by a positive step change of 1 m³/h of the flow rate of slurry (Q_s). The performance of the different control schemes in rejecting such disturbance are analyzed below in the nominal situation and in robust conditions, that is, when uncertainties on the model of process and disturbance are present. The schemes are evaluated on the basis of different criteria. The structural simplicity is the first feature to be considered, then, some performance indices are computed. Moreover, the robust stability, and the physical realizability through some constraints on control variables are analyzed.
In the nominal situation, a unitary load disturbance on Q_s enters at time 20s, to be considered initial time of analysis of the variables as deviation from a zero stationary condition (see Figure 4). A specification on the deviation from T_3^o ($|\Delta T_3| < 3$ °C), and constraints on the deviations of T_2 ($\Delta T_2^{hl} = 250$ °C) and of the fuel gas flow rate ($\Delta Q_f^{hl} = 650$ Nm³/h) are imposed.

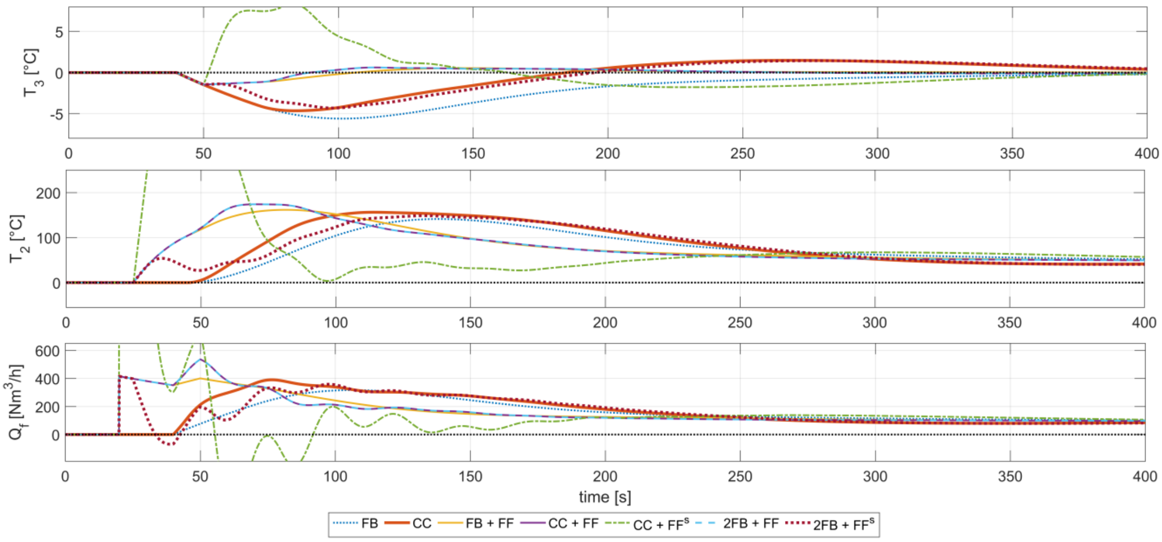

Figure 4: Time trends for rejection of disturbance Q_s in the nominal case.

It can be observed that FB and CC schemes are not sufficient to control T_3. Slow responses and excessive deviations (of about 5 °C) are recorded, thus the specification on ΔT_3 is not respected. All configurations with FF component allow a very good performance on T_3, and the best response is obtained by FB + FF. Also CC

+ FF and 2FB + FF are effective on T_3; to be noted that they show exactly the same behaviour: despite having different structures, they produce the same practical effect on control variables. Anyway, note also that they require a sharp peak of Q_f during transient response, which is close to the constraint on the fuel gas flow rate (ΔQ_f^{hl}). 2FB + FF^s allows a simplified design, and shows tolerable performance on T_3 (deviations of about 4 °C), but produces some oscillations on Q_f, and settling times similar to what obtained by CC scheme. Finally, performance of CC + FF^s are definitely not acceptable, since large deviations occur not only on T_3, but also on T_2 and Q_f; specification and constraints are not respected.
To be recalled that perfect rejections of the disturbance would be possible if ideal formulations of FF components were PR, that is, when disturbance is slower than process dynamics ($\theta_d > \theta$). In addition, it is to be stressed that improving controller tuning in simple schemes (FB and CC) does not help, while FF component is necessary to reject such type of external disturbance. CC does not significantly outperform FB, but is highly required, because internal disturbances, as changes in fuel gas composition or input temperature to the burner, must to be considered unavoidable.
Only control schemes showing acceptable results in the nominal case are further analysed to test robustness. Thus, CC + FF^s scheme is discarded due its very bad performance, while 2FB + FF is not considered since is proved to be totally equivalent to CC + FF scheme. Two scenarios, different from nominal conditions, are studied. In the first, only errors on process dynamics (G_t) are present, in the second, the errors are on the disturbance dynamics (G_d). In both cases, a multiplicative uncertainty is considered, so that the same relative error is present on the three parameters of the FOPTD models of Table 1. The actual parameters are:

$$\begin{array}{lll} K = K^o(1+\Delta) & \tau = \tau^o(1+\Delta) & \theta = \theta^o(1+\Delta) \\ K_d = K_d^o(1+\Delta_d) & \tau = \tau_d^o(1+\Delta_d) & \theta_d = \theta_d^o(1+\Delta_d) \end{array} \qquad (1)$$

where Δ and Δ_d are the levels of uncertainty, and the superscript “o” refers to nominal parameters. Note that such type of correlated and symmetric uncertainties is typical of variation of operating conditions due to variations in flow rate. Time trends for rejection of a unitary load disturbance on Q_s in the case of error on process model and disturbance model have been studied for increasing values of the uncertainty. In Figure 5, results are reported for a relative error of 75 % on G_t and G_d parameters, respectively.

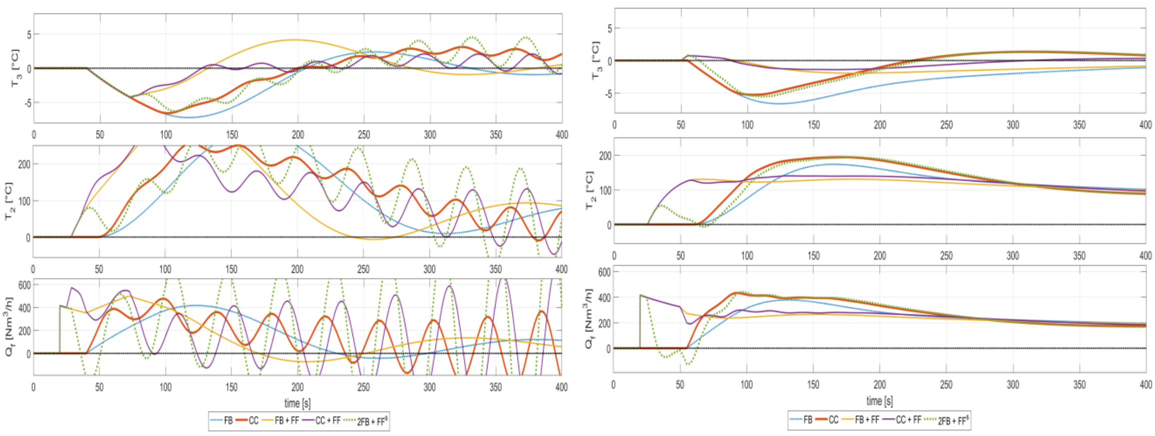

Figure 5: Time trends for rejection of disturbance Q_s in uncertain cases: left) error on process model (Δ = 0.75); right) error on disturbance model (Δ_d = 0.75).

It is well known that beyond some level of uncertainty, every control scheme may go into instability conditions. From Figure 5, it is to be noted that instability occurs only in the case of error on process model, and not for disturbance dynamics. This is because disturbance dynamics does not affect the characteristic equation of every control scheme, which instead includes only process dynamics (G_t, or combinations of G_1 and G_2). Therefore, only the feedback component is responsible of instability in closed-loop mode, while FF elements do not play any additional negative role and are inherently robust (Lewin and Scali, 1988).
Table 3 shows the interval of values for some performance indices in the case of errors on process dynamics, ranging from the nominal case to 75 % of uncertainty, which causes unstable conditions. The integral of the absolute value of control error on T_3 (IAE), the total variation of the output temperature from burner (ΔT_2), the total variation of fuel gas flow rate (ΔQ_f) to be correlated with the total variation of control action, and the maximum values of deviations for T_3, T_2, and Q_f ($|\Delta T_3|^{max}$, ΔT_2^{max}, ΔQ_f^{max}) are computed.

FB scheme produces the largest value of IAE in the nominal case, but it also shows good robustness to uncertainty. CC slightly outperforms FB, but can show oscillations in uncertain cases, and it is unstable for Δ = 0.75, as indicated by large values of ΔT_2 and ΔQ_f. FB + FF is more robust than CC + FF, but tends to present slow trends and unacceptable deviations on T_3. CC + FF allows the lower values of IAE at the increase of the uncertainty, but produces larger oscillations on T_2 and Q_f, which fall into instability for Δ = 0.75. The simplified scheme 2FB + FF^s, with almost acceptable performance in the nominal case (see Figure 5), now reveals high sensitivity to uncertainty and quickly reaches unstable conditions, thus proving to be a non practical solution. Overall, CC + FF scheme proves to be the best solution, with an acceptable tradeoff between performance and robustness. For example, for Δ = 0.35, the specification on T_3 is guaranteed ($|\Delta T_3|^{max}$ = 2.85) with a tight control (IAE = 220), and by respecting constraints on manipulated variables (ΔT_2^{max} = 229, ΔQ_f^{max} = 552).

Table 3: Interval of values for performance indices in the case of error on process model (Δ = 0 – 0.75)

Control scheme	IAE	ΔT_2	ΔQ_f	$|\Delta T_3|^{max}$ [°C]	ΔT_2^{max} [°C]	ΔQ_f^{max} [Nm³/h]
FB	[668 – 966]	[230 – 641]	[532 – 1043]	[5.6 – 7.2]	[141 – 292]	[317 – 416]
CC	[585 – 936]	[272 – 1008]	[704 – 6958]	[4.7 – 6.6]	[156 – 255]	[390 – 475]
FB + FF	[109 – 677]	[274 – 757]	[831 – 1485]	[1.5 – 4.2]	[162 – 322]	[415 – 499]
CC + FF	[118 – 416]	[298 – 1886]	[1133 – 14350]	[1.5 – 4.2]	[173 – 288]	[536 – 794]
2FB + FF^s	[571 – 937]	[311 – 3119]	[1885 – 25227]	[4.3 – 6.3]	[148 – 290]	[415 – 1319]

5. Conclusions

The problem of the control of a complex plant has been represented by a series of linear models which can be considered sufficiently reliable and valid to evaluate the effects of various parameters and disturbances on the main controlled variables. A change in the flow rate of silica slurry is the most important perturbation affecting the product quality (moisture) and requires a tight control of the dryer output temperature.

This specification can be obtained only adopting feedforward control (FF), even though a perfect neutralization is not possible also in the nominal case, owing to faster dynamics of perturbation with respect to the manipulated variable (fuel gas). A redesign of the plant would permit to get closer to this ideal situation. Cascade control does not outperform simple feedback, but is required by the unavoidable presence of other internal disturbances, so the proposed complete scheme includes both features. The design and algorithm of the FF component may seem a bit involved, but approximate structures prove not to be able to give acceptable performance and exceed constraints on manipulated variables.

Being based on a linear model, results of the different control schemes have been validated for parameter variations, directly correlated with variations of gas flow rates and then of production of the plant. Therefore, the cascade plus feedforward package can be considered robust in a wide range of operating conditions.

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