Ann. Inst. Statist. Math.
Vol. 40, No. 3, 419–438 (1988)

APPLICATION OF TIME SERIES ANALYSIS AND MODERN CONTROL THEORY TO THE CEMENT PLANT

S. HAGIMURA[1], T. SAITOH[2] AND Y. YAGIHARA[3]

[1]*Chichibu Cement Co., Ltd., 794-6, Chichibu-city, Saitama 368, Japan*
[2]*Chichibu Cement Co., Ltd., 5310, Mikajiri, Kumagaya-city, Saitama 360, Japan*
[3]*System Sougoh Kaihatsu Co., Ltd., 1-28-23, Hongou, Bunkyo-ku, Tokyo 113, Japan*

(Received July 28, 1988; revised August 17, 1988)

Abstract. Cement plant process is full of internal noise sources and feedback loops. Therefore, statistical approach is required to understand its dynamic characteristics. Time series analysis has been applied to some important subprocesses of a cement plant process. These are the vertical mill process, calcining process and clinker cooling process. Based on the AR models of these, a set of optimum controllers have been designed by modern control theory. Successful results of application are reported in this paper. A method of determining optimal production level is also discussed.

Key words and phrases: Time series analysis, identification, control, cement plant process.

1. Introduction

Cement plant process constitutes a multiple inputs and outputs system with many internal noise sources and feedback loops. To control such a system successfully, a model of the actual process is required. In the cement process the most important one is the kiln process. This process is a distributed parameter system that has an extension of its behavior over space and time. In the initial stage when first process computer was installed at Kumagaya plant of Chichibu Cement Co. Ltd. for process control, a mathematical model derived by the equation of the heat and material balance within the kiln was used as a process model for control. This model was consisted of simultaneous partial differential equations of gas and material temperatures. However, this model could not take account of the noise sources existing in the actual process. In fact, control using this model was not quite successful.

On the other hand, by aggregating a kiln process into a lumped parameter system, conventional local control of PID type with single input and output was also tried. The resulting control was robust, but the result was not always satisfactory because it could not take account of the interactions among the process variables.

Since the introduction of the second process computer at Kumagaya plant in 1967, raw material mill process and cement rotary kiln process have been analyzed with time series analysis method. AR models of cement kiln processes were estimated by using FPE criterion, and the computer control systems were realized based on these models. Successful results have been obtained by this approach (Otomo *et al.* (1969)). This is an important contribution in the field of control, such that a systematic procedure for model identification and control system design for noisy industrial processes is established for the first time. In 1983, we applied time series analysis to a new type of cement kiln process, called NSP (New Suspension Preheater) kiln process and obtained practically useful AR models of calcining and clinker cooling process. Since then an on-line control based on these models has been running up to the present. In this paper, we will discuss our experience on this application and a further advanced system design.

2. Explanation of cement plant

In 1970, Chichibu Cement company succeeded in controlling the wet type kiln, that uses the raw material in the form of wet slurry, by an optimum control, called optimum regulator control. Early report on this work was submitted at the IFAC 4th World Congress at Warsaw in 1969 (Otomo *et al.* (1969)) and detailed reports are given in Otomo *et al.* (1972) and Akaike and Nakagawa (1973).

About fifteen years ago, the type of cement kiln was changed from wet to dry type, that used powdered dry material, for better productivity and energy saving. In 1983, we performed an analysis of a new type of kiln, called SF (Suspension preheater with Flash furnace) kiln by using a micro computer system SILTAC (Self-Instructive, Learning and Tutorial system for statistical Analysis and Control of dynamic systems). SILTAC is an interactive system developed and commercialized by System Sougoh Kaihatsu Company based on the TIMSAC program package given in Akaike and Nakagawa (1973).

We fitted AR models for the calcining and clinker cooling process of the SF kiln, and optimal controls were developed for these processes. Since then these on-line controls have been running as described in Hagimura *et al.* (1986). In 1985, the same type of controller were implemented at Chichibu No. 2 plant. Some other new applications of SILTAC have been developed in succession at Kumagaya plant, such as raw material blending mill (vertical mill) process control, prediction system of cement quality, etc.

Here we will explain briefly the cement production process for readers who are unfamiliar with the process. The cement production plant is composed of raw material process, NSP (New Suspension Preheater) kiln process, finishing process and shipping process as shown in Fig. 1.

At raw material process, four kinds of raw material are weighed, ground and blended. Homogenized raw mixture is stocked in storage silo.

NSP kiln process is divided into the following three subprocesses;

(1) the cyclone suspension preheater tower (P.H. tower) process where the calcining of raw mixture is performed,

(2) the rotary kiln (kiln) process where clinker, being half-finished product, is burned,

(3) the clinker cooling (cooler) process where clinker is cooled quickly.

Among the several cement production processes, the kiln process is most important in keeping the cement quality. It is required to perform stable calcining of raw mixture, steady burning of clinker and steady cooling of clinker. These subprocesses have various internal disturbances and feedback loops. Therefore, it is very difficult to control these processes. Here a practical procedure of model identification and control system design is badly needed.

In the finishing process, the clinker is ground together with gypsum ($CaSO_3 \cdot 2H_2O$) in a ball mill to produce the finished product, i.e., cement.

3. Identification and control of the kiln process

3.1 *Calcining control*

The calcining process exists in FF (Flash Furnace), as shown in Fig. 2, where the following chemical reaction takes place at about 850°C;

$$CaCO_3 \rightarrow CaO + CO_2\uparrow .$$

The above reaction absorbs heat of about 420 KCal/kg and the necessary heat is supplied by the FF coal feed, exhaust gas from rotary kiln and secondary air from clinker cooler. Fluctuations of the heat content of the inlet gas from the kiln and cooler must be regarded as uncontrollable disturbances of the calcining process. Accordingly, for the purpose of the stationary reaction the outlet gas temperature T_1 of C_4 cyclone must be kept within a narrow allowance of the constant set value by manipulating FF coal feed. Moreover, the feedback from the secondary air temperature T_3 to the outlet gas temperature T_1 (see Fig. 2) has time lag of about 25 min. We were concerned with the possible occurrence of the lag oscillation caused by this time lag under particular type of disturbances. Figure 3 illustrates that our anxiety was in fact real. Conventional PID controller was not only unable to reduce the oscillation, but often induced it, because of the lack of the consideration of the interaction between T_1 and T_3.

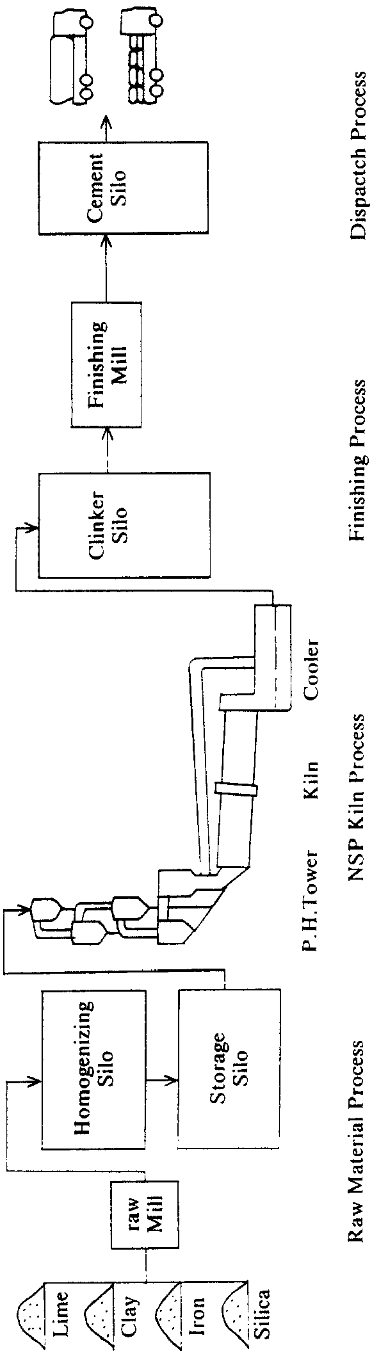

Fig. 1. Cement plant general flow.

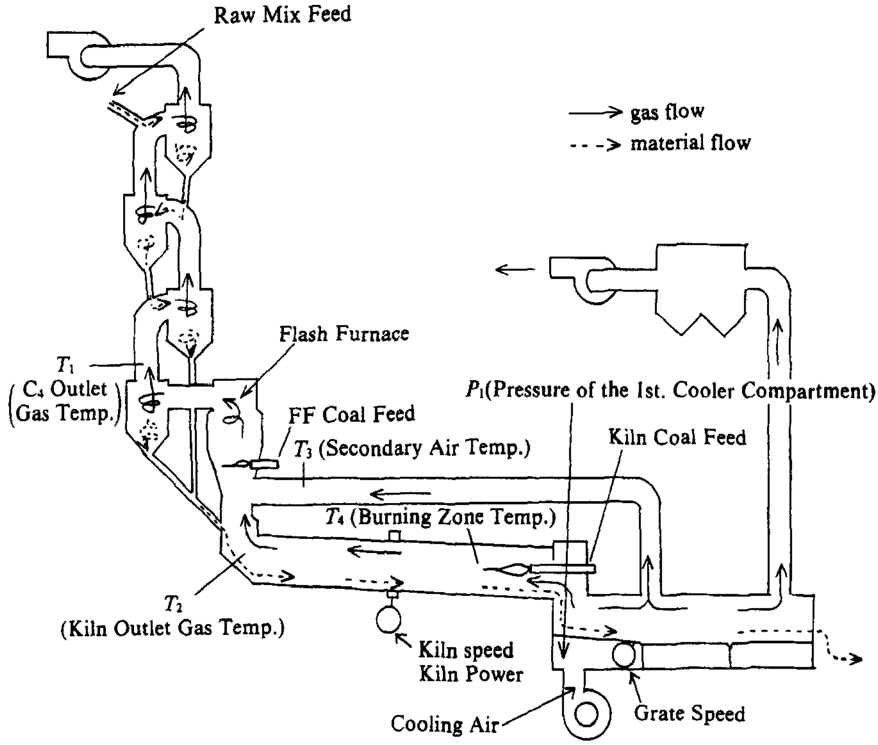

Fig. 2. General flow of NSP kiln.

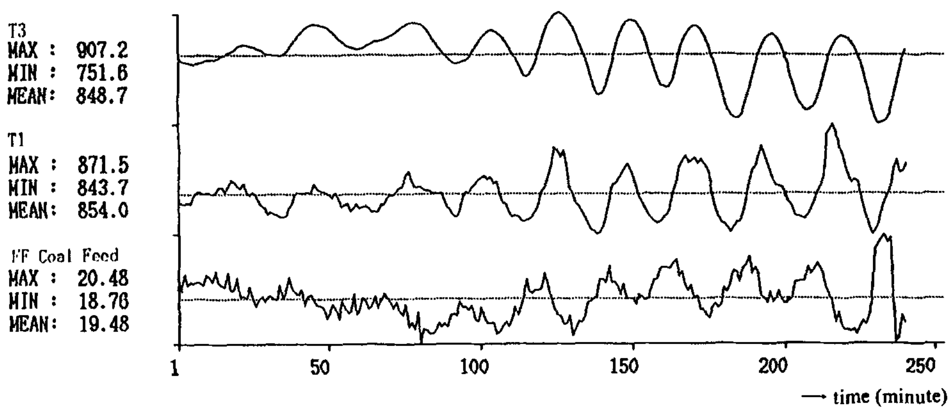

Fig. 3. PID control of T_1 and lag oscillation between T_1 and T_3.

Our purpose of control is to stabilize T_1 by regulating FF coal feed. We analyzed the behavior of the calcining process by selecting six variables, kiln coal feed, FF coal feed and raw mix feed as inputs and T_1, T_2 and T_3 as outputs in Fig. 2. Firstly, we adopted the following multiple AR model,

$$X(n) = \sum_{m=1}^{M} A_m X(n-m) + U(n) ,$$

where $X(n)$ is a six-dimensional process variable vector, A_m's are coefficient matrices and $U(n)$ is a white noise vector. For easy reference of the model identification and controller design procedures to be used in the present study, see Nakamura and Toyota (1988). Then the minimum AIC, or minimum FPE, procedure was applied to a data set of length 150 and produced the model order $M = 2$.

The relative power contributions (Akaike (1968), Akaike and Nakagawa (1973)) of the variables to the variation of T_1 are shown in Fig. 4. Kiln coal feed and T_2 contribute to T_1 only in the frequency domain near direct current (DC), and raw mix feed, FF coal feed and T_3 contribute significantly to T_1 in the medium frequency band (time period $30 \sim 60$ min.). The large contributions of kiln coal feed and T_2 to T_1 suggest that these variables control the basic behavior of heat to keep the whole process of kiln at an appropriate heat environment. This is in good agreement with our experiences.

To control this system we developed the following model of the process with raw mix feed and FF coal feed as inputs and T_1, T_3 as outputs,

$$X(n) = \sum_{m=1}^{M} a_m X(n-m) + \sum_{m=1}^{M} b_m Y(n-m) + W(n) ,$$

where $X(n)$ is controlled variable vector (output), $Y(n)$ is manipulated variable vector (input) and $W(n)$ is a white noise vector about $X(n)$. a_m's and b_m's are coefficient matrices. The behavior of the value of FPEC

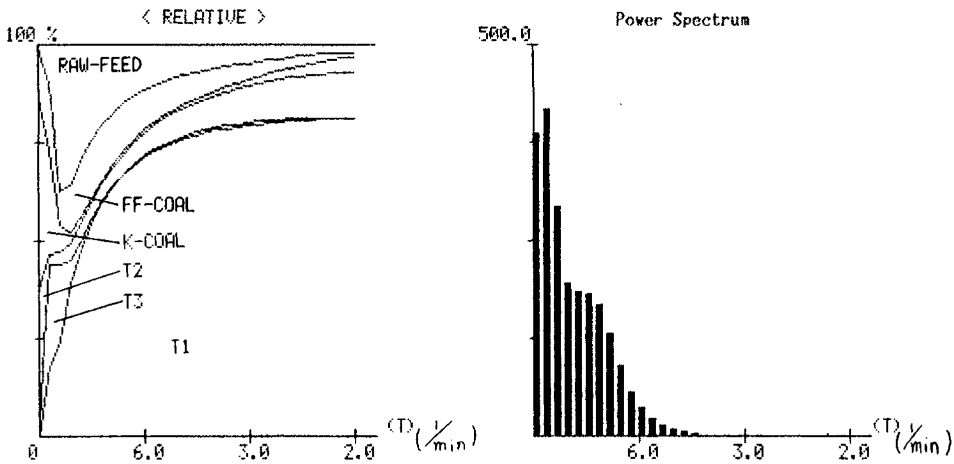

Fig. 4. Relative power contribution of T_1 and power spectrum.

criterion vs model order was checked and the third order model was selected. The covariances of innovations of the AR model of order 3 was close to diagonal form (Table 1). The contributions to T_1 from raw mix feed, FF coal feed and T_3 were similar to Fig. 4. Our experience of the process suggests that this simple model (T_1, T_3; raw mix feed, FF coal feed) is best for the present purpose of control.

However, amount of production, being directly proportional to raw mix feed, is subject to the variation of raw mixture feed. When the raw mixture feed is frequently manipulated by control, the amount of production does not accord well with the decision by the management. Therefore, in the actual control we adopted the model (T_1, T_3; FF coal feed) without raw mix feed, that is model (4) in Table 2.

The process model was transformed into the following state space model in order to design the optimum controller,

$$Z(n) = \Phi Z(n-1) + \Gamma Y(n-1)\,,$$

$$X(n) = HZ(n)\,,$$

where $Z(n)$ is state variable vector and Φ, Γ, H are coefficient matrices defined as follows.

Table 1. Normalized covariance matrix of innovation.

	1	2	3	4
1	1.00	−0.05	0.03	0.02
2	−0.05	1.00	−0.07	−0.12
3	0.03	−0.07	1.00	−0.06
4	0.02	−0.12	0.06	1.00

[System output]: 1 = T_3, 2 = T_1. [System input]: 3 = FF-coal, 4 = Raw mix feed.

Table 2. Comparison with AIC.*

Variable name	Model(1) $M=2$(1)	Model(2) $M=3$	Model(3) $M=3$	Model(4) $M=3$	Model(5) $M=3$
T_2	O	O	O		
T_3	O	O	O	O	O
T_1	O	O	O	O	O
Kiln coal	O				
FF-coal	O	O	O	O	O
Raw mix feed	O	O			O
Min. AIC(2)	1520.4	1530.5	1539.4	924.9	914.5

(1) M = model order minimizing AIC(M).
(2) AIC(M) = $N \log_e$ FPEC(M).
* Data length $N = 150$, Sampling interval $\Delta t = 1$ min.

$$\Phi = \begin{pmatrix} a_1 & 1 & 0 & \cdots & 0 \\ a_2 & 0 & 1 & \cdots & 0 \\ \vdots & \vdots & \vdots & \ddots & \vdots \\ a_{M-1} & 0 & 0 & \cdots & 1 \\ a_M & 0 & 0 & \cdots & 0 \end{pmatrix}, \quad \Gamma = \begin{pmatrix} b_1 \\ b_2 \\ \vdots \\ b_{M-1} \\ b_M \end{pmatrix}, \quad H = [\ 1 \quad 0 \quad \cdots \quad 0\ ] .$$

Then the optimum controller was designed to minimize the following loss function by manipulating $Y(s-1)$,

$$J_I = E\{K_I\} ,$$

$$K_I = \sum_{s=1}^{I} \{Z(s-1)^T Q Z(s-1) + Y(s-1)^T R Y(s-1)\} ,$$

where Q is a nonnegative definite weighting matrix, R is a positive definite matrix and I was put equal to 10. Dynamic programming method was applied to compute optimal control gain G, and the control was realized by

$$Y(n) = GZ(n) .$$

The control gains were tuned by assigning various values to the diagonal elements of the weighting matrices Q and R. Usually, the inverses of the one-step prediction variances of controlled variables and variances of manipulated variables were assigned to the corresponding diagonal elements of Q and R, respectively. This choice of Q was often too strict and the actual control showed hunting behavior, because the actual process was under the influences of the other variables which the model did not include. Therefore, Q and R were tuned empirically. The diagram of control system is shown in Fig. 5.

The results of actual control by both conventional PID and optimum control, hereafter called the SILTAC control, is shown in Fig. 6. The PID control did not use the information of T_3, in determining the amount of FF coal feed. In the SILTAC control T_3 was successfully used for the prediction of T_1. As it is impossible to control T_3 directly with FF coal feed, the

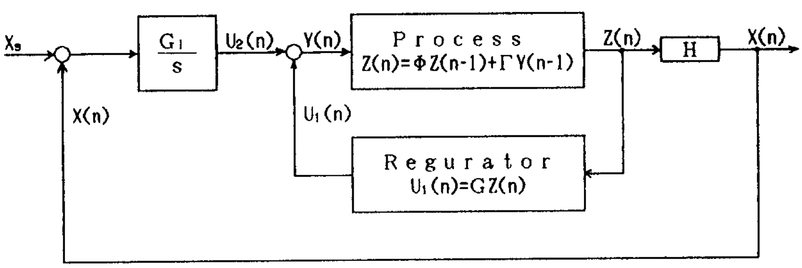

Fig. 5. Block diagram of control system.

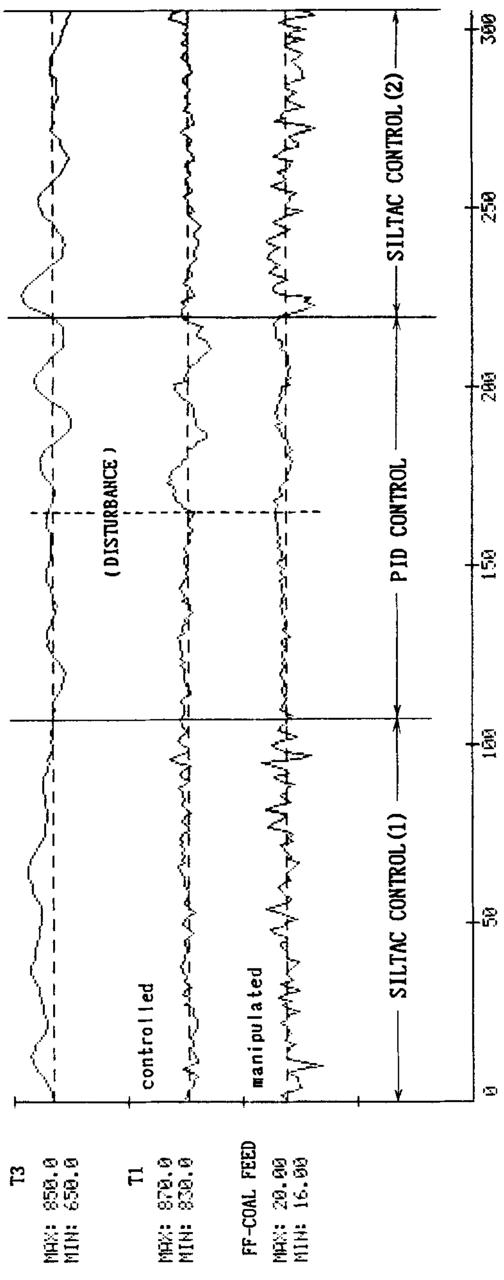

Fig. 6. Result of PID control and SILTAC control.

control gains from T_3 were put nearly equal to zero through tuning Q. We can see from Fig. 6 that the process was very stable under the SILTAC control(1), but unstable under the PID control and the lag oscillation was started by the mudring failure, caused by the dropping of the clinker coating of the kiln wall. This lag oscillation in T_1 was dumped promptly by the SILTAC control(2). Table 3 and Figure 7 show comparison of average, variance and power spectrum under both controls. We can see by Table 3 and Fig. 7 that while T_3 is kept free, T_1 is controlled within a narrower range around the set point (850°C) by the SILTAC control than the PID control, and also the fluctuation nearly around the DC component is suppressed well by the SILTAC control. Pay attention to difference of the each scaling factors in Fig. 7. In Fig. 7, the PID control appears better than the SILTAC control nearly around time period five minutes, but the reasons were that the response of T_1 from FF-coal has time delay about five minutes and the integral action of the PID control was too large because of imperfect tuning of reset time.

3.2 *Clinker cooling control*

The clinker cooling process is controlled by quenching with a cold air. The heat exchanged air through the hot clinker, called the secondary air, is at about 750 ~ 850°C and used for the kiln and FF coal firing. Large fluctuations of this heat exchange disturb the combustion of fuel in the kiln and flash furnace, and this causes the fluctuations of temperatures in the kiln, especially the burning zone temperature T_4 and outlet gas temperature T_1 from C_4 cyclone. This leads to the inefficient use of fuel and poor quality of clinker. The prime cause of the change of heat is the variation of the thickness of clinker layer on the grate, and we can measure the thickness indirectly through measuring the pressure in the compartment under the grate. In conventional control, cooler grate speed was manipu-

Table 3. Average, variance, Std.Dev. under PID/SILTAC control.

		SILTAC(1)	PID	SILTAC(2)
$T_3^{(1)}$	Average	805.50	784.90	794.20
	Variance	169.00	234.60	1152.00
	Std.Dev.	13.00	15.32	33.94
$T_1^{(2)}$	Average	849.90	849.30	849.10
	Variance	3.52	9.98	3.85
	Std.Dev.	1.88	3.16	1.96
FF-Coal	Average	18.53	18.67	18.45
	Variance	0.105	0.026	0.175
	Std.Dev.	0.324	0.16	0.418

(1) Set point was free.
(2) Set point was 850.0°C.

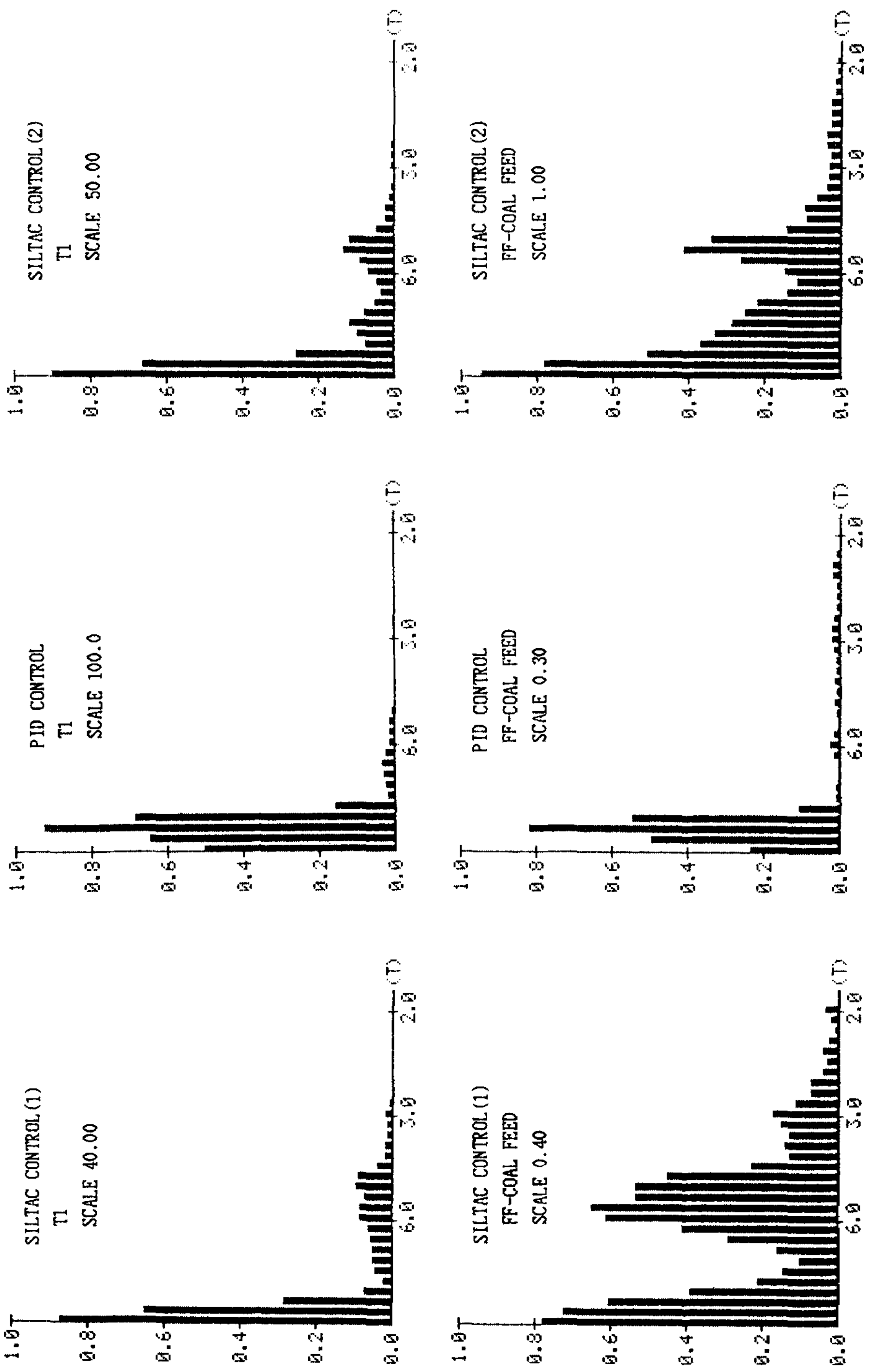

Fig. 7. Power spectra of T_1 under PID/SILTAC control.

lated to keep the pressure constant by PID controller. However, the secondary air temperature was not always kept constant by keeping the pressure constant. For the purpose of stabilizing the secondary air temperature, we considered the system with five variables shown in Fig. 8, and identified a model through these data. The sampling interval was a half minute, and data length was 375.

According to Fig. 8, wattage of the kiln driving motor (kiln power), burning zone temperature (T_4) and secondary air temperature (T_3) have larger time constants compared with those of the cooling process. The identification of a model for the process which contains variables having extremely different time constants each other is difficult, and the frequency property of variables has to be taken into account in the selection of variables. For this reason, we considered the model without kiln power and T_4. The results by the PID and SILTAC control are shown in Fig. 9. As the process under both these controls was stationary and stable, both controls achieved nearly similar results.

3.3 *Raw mill control*

Vertical mill (Duda (1985)), of which the principle of grinding is the same as that of stone mill, is superior to ball mill in power consumption. Two types of control of vertical mill, the mill pressure control and the bucket elevator power control, have been developed for power energy saving and stable operation. The mill pressure control keeps the difference of pressure between the inlet and outlet constant by manipulating raw material feed, and consequently the quantity of raw material on the table feeder in the mill is kept constant. The bucket elevator power control keeps the electric power of bucket elevator constant by manipulating raw material feed, and consequently the quantity of raw material fed into the mill is kept constant.

A single input and single output PID controller has been used for the mill pressure control, but we can not evaluate which is superior, the mill pressure control or the bucket elevator power control, viewing from the point of energy saving. We applied the SILTAC control with the AR model which consists of mill pressure, bucket elevator power and raw material feed. Only the mill pressure could be stabilized by the PID control as shown in Fig. 10, but both the mill pressure and bucket elevator power were stabilized by the SILTAC control as shown in Fig. 11.

3.4 *Identification under the on-line control*

Dynamic properties of a process may deviate gradually from the originally identified model. The phenomena can be monitored by observing sequentially one-step ahead prediction errors of the model. If the drift of sequential errors are observed, the process model must be re-identified. In such a situation it is desirable that the identification is performed under

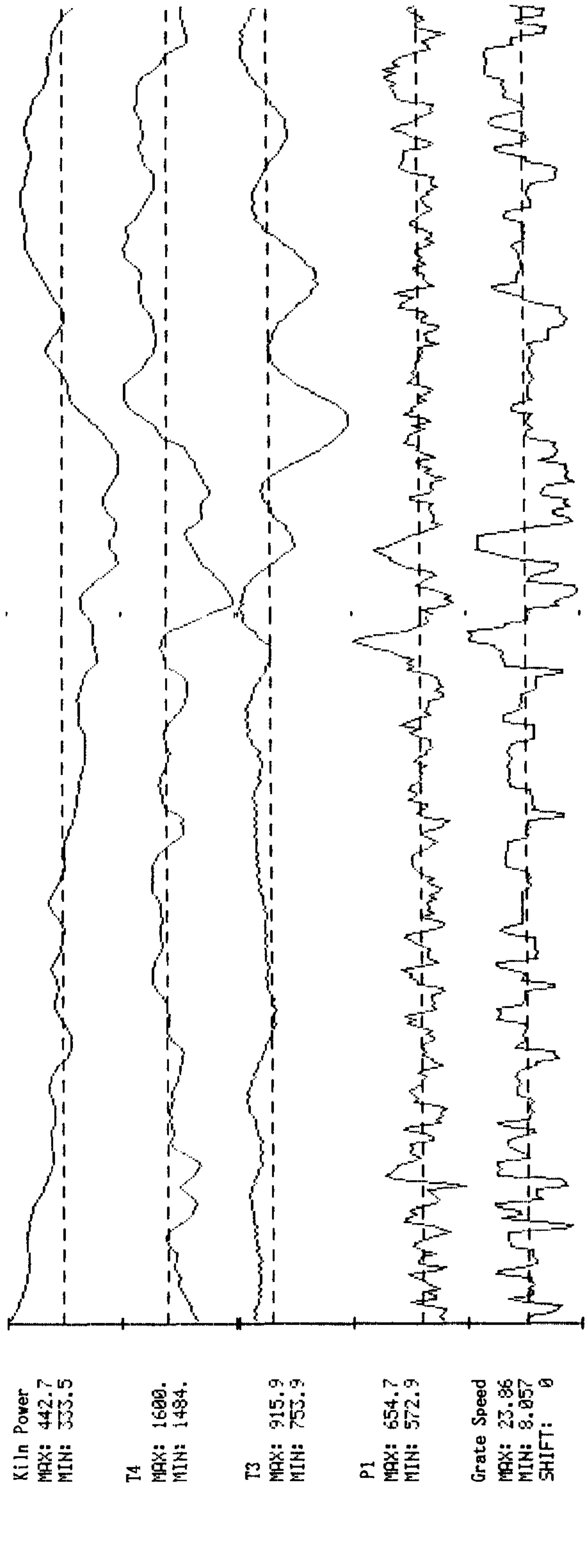

Fig. 8. Data of cooling process.

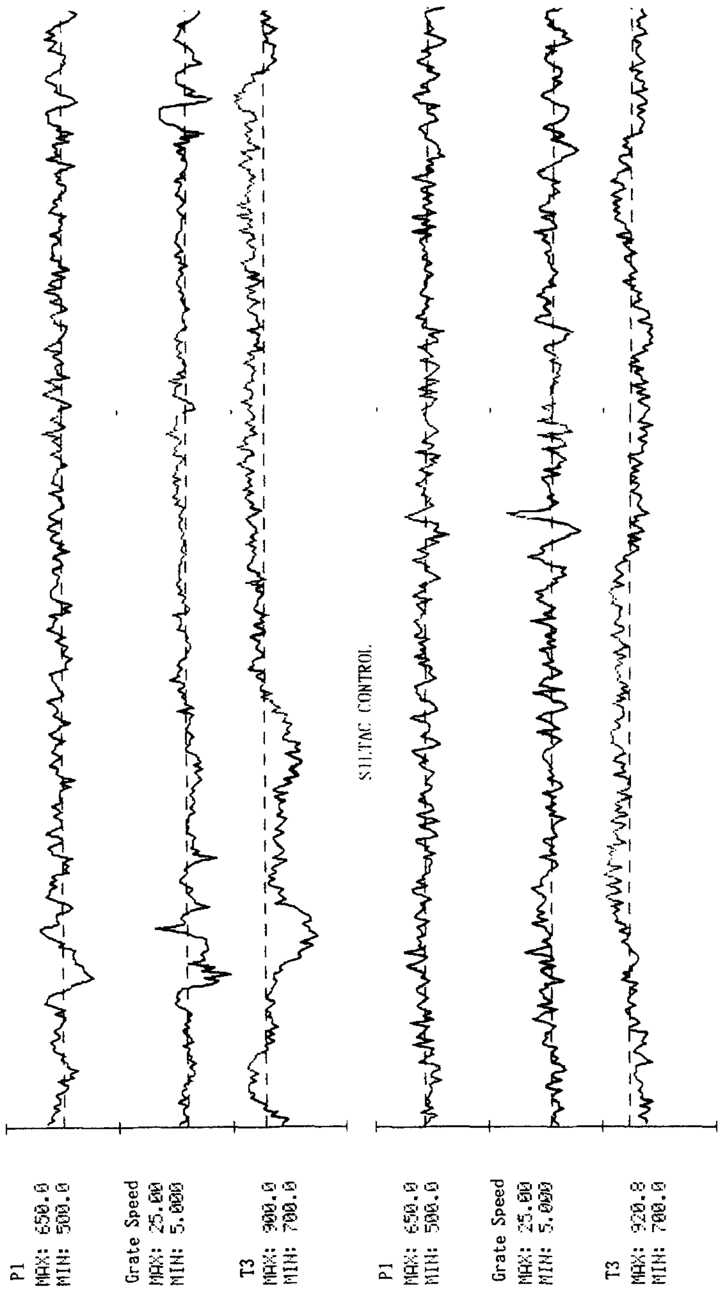

Fig. 9. Result of PID control and SILTAC control.

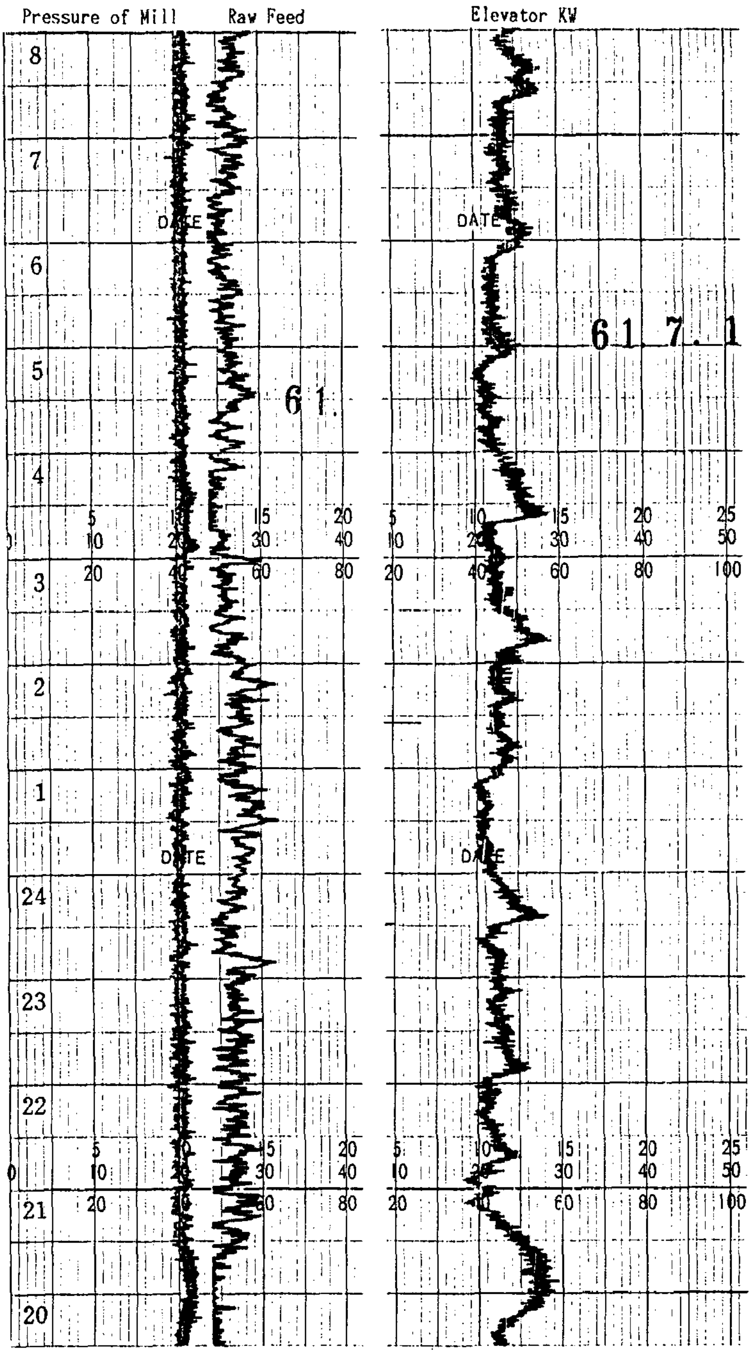

Fig. 10. Result of PID control (vertical mill process).

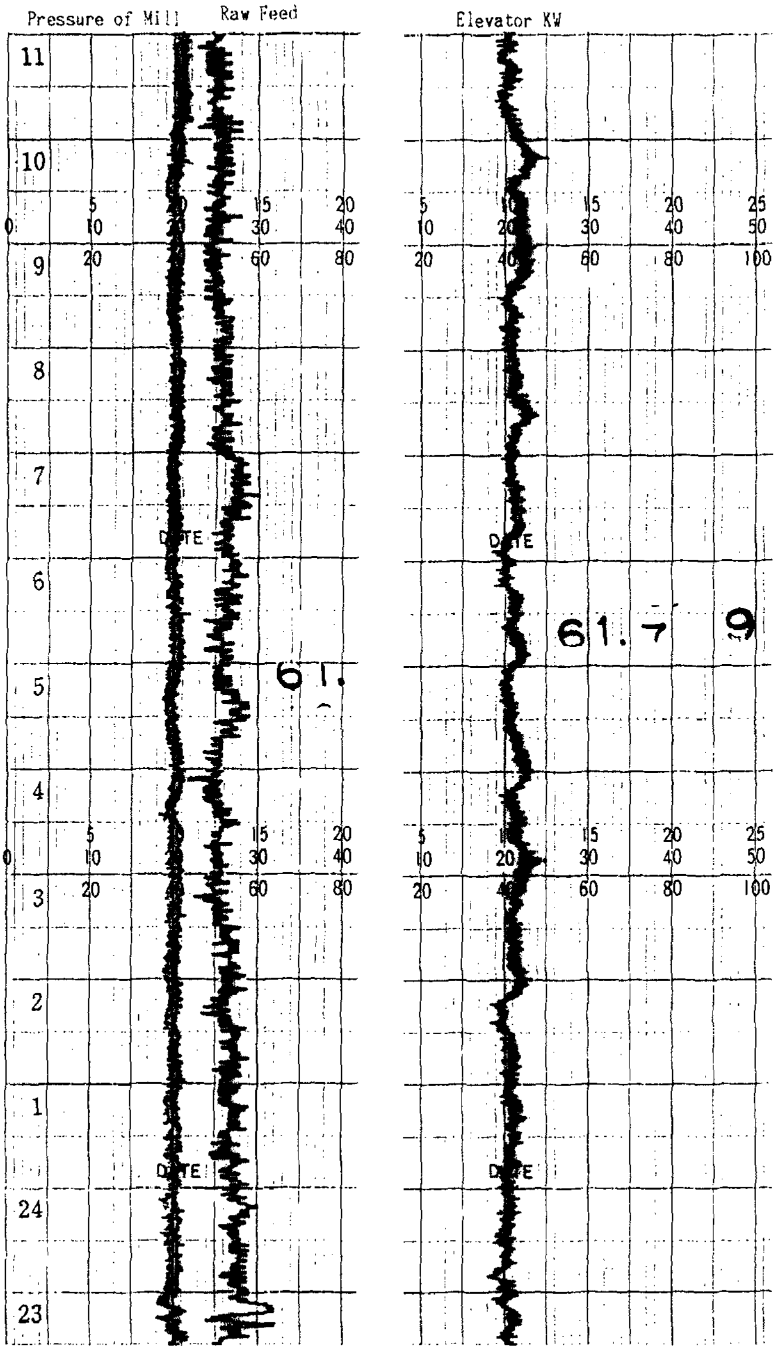

Fig. 11. Result of SILTAC control (vertical mill process).

on-line control. We collected the data for the identification by adding white noises with adequate variances to the control signals of manipulated variables under on-line control. According to our experience, the identification by this method is successful, if the white noises are mutually independent and the magnitudes of the signals are properly chosen. The adequate magnitudes were two or three times as big as standard deviation of raw data. At present, an on-line algorithm of this type is under test at Kumagaya plant.

3.5 *Optimum production level and pursuit control*

A production level is decided by the plant manager. Based on the decided production level, the field operator has to decide a set of the balanced set point values of controlled variables in control systems. Then he tries to minimize the production cost and to keep high quality of products. The decision is very difficult but important. The multiple AR model for the decision of the set points is identified through process data with longer sampling intervals, such as eight hours, than the stability control. The decision problem of the set points is solved by linear programming (LP method) subject to the steady state model (Nakagawa and Yagihara (1985), Hagimura *et al.* (1986)). Suppose the dynamic model is as follows,

$$X(n) = \sum_{m=1}^{M} a_m X(n-m) + \sum_{m=1}^{M} b_m Y(n-m) .$$

Then, the following steady state model is obtained (Jury (1958)).

$$X_s = K_p Y_s ,$$

$$K_p = \left[I - \sum_{m=1}^{M} a_m \right]^{-1} \left(\sum_{m=1}^{M} b_m \right),$$

where X_s and Y_s are a set of balanced set point vectors and K_p is the steady state gain matrix of the process. The formulation of LP problem is as follows.

$$J = \alpha^T X_s + \beta^T Y_s \rightarrow \text{minimizing or maximizing} .$$

Subject to;

$$X_s - K_p Y_s = E_s(m_x, m_y) ,$$

$$X_s \leq X_u - X_L ,$$

$$Y_s \leq Y_u - Y_L ,$$

$$X_s,\ Y_s \geq 0\ ,$$

where $E_s(m_x, m_y) = (m_x - X_L) - K_p(m_y - Y_L)$ and m_x, m_y are mean value vectors of $X(n)$ and $Y(n)$ used for identification, X_u, Y_u are upper limit vectors and X_L, Y_L are lower limit vectors. The object function is chosen by the policy of the plant manager. The concrete object function J is composed by using principal component analysis or canonical correlation analysis in order to realize energy cost saving and/or high quality keeping. Figure 12 shows data used for identification, and the results of simulation

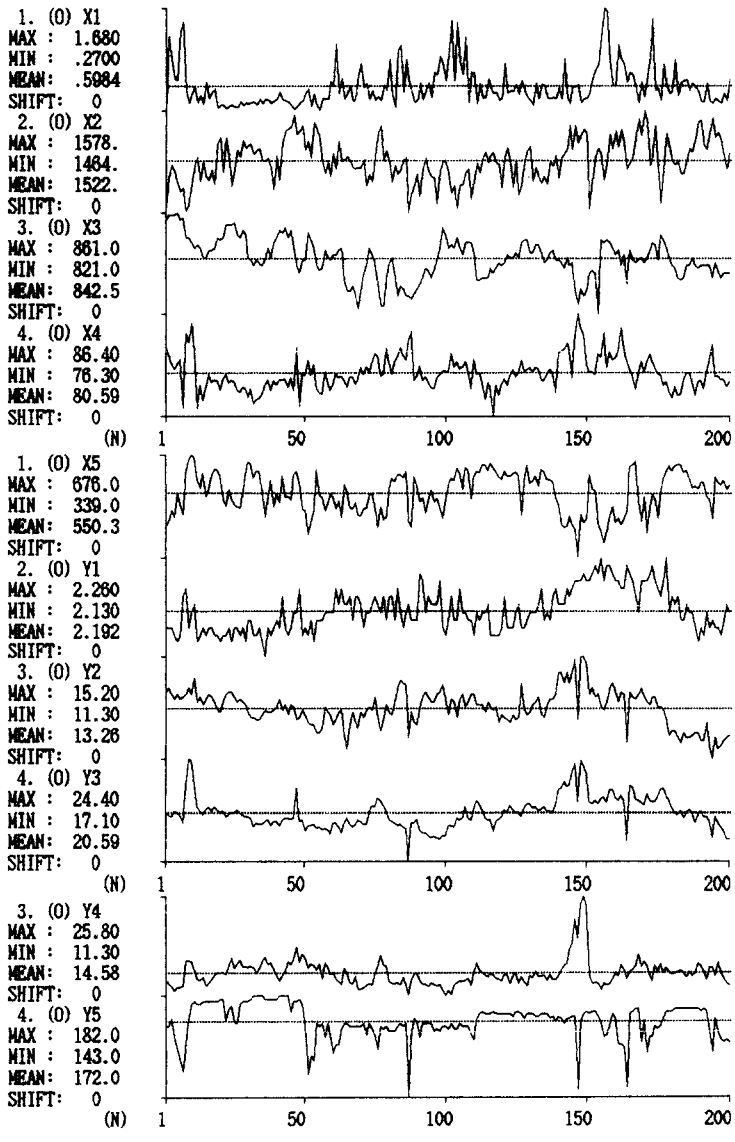

Fig. 12. Data for identification of production level model.

by LP are given in Table 4.

When a set of the optimum set points is given, the present state of process is guided to these new set points by pursuit control with many variables. The pursuit control system is composed of a state feedback (feedback compensator) and an integral action (serial compensator) as shown Fig. 5, and transfer gradually and smoothly the current state of process to the new set points.

By realizing the above-mentioned optimum production level dynamically, we can step up to a dynamic production management from a traditional static one.

4. Conclusion

We have applied time series analysis method and modern control theory to raw material blending process (raw mill control), calcining process (calcining control) and clinker cooling process (clinker cooling control) in cement plant process. In there, the multiple AR models and the optimum regulators with the state feedback and integral action were implemented. These on-line controls were so useful in stable operation and energy saving as a human operator hesitated to stop these on-line controls. However, because of the gradual deviation of dynamic properties of a process apart from the originally identified model, the process model needs to be re-identified under on-line control so as to continue the on-line control for a long time. The development of the procedures for the auto-

Table 4. Optimum production level by LP.*

Variable name	Mean value[1] (A) ±Std. Dev.	LP solution[2] (B)	Difference (B)−(A)
X_1; fcao[3]	0.5984± 0.2758	0.3636	−0.2384
X_2; T201	1521.80±23.658	1513.41	−7.89
X_3; T32	842.52± 7.9095	840.42	−2.10
X_4; oil-l/t[4]	80.59± 1.6934	79.11	−1.48
X_5; W200	550.34±67.480	617.63	67.29
Y_1; kiln-HM	2.1919± 0.0267	2.1651	−0.0268
Y_2; W252	13.26± 0.6924	12.57	−0.69
Y_3; W262	20.59± 1.1242	19.47	−1.12
Y_4; N215-1	14.58± 1.9776	16.55	1.97
Y_5; N200[5]	172.02± 6.7435	170.28	−1.74

(1) Mean values of data used for identification (m_x, m_y).

(2) Optimum set point values solved by linear programming (X_s, Y_s). Upper and lower limit of variables; σ_x, σ_y = Std. Dev. of $X(n)$ and $Y(n)$

$$(m_x, m_y) - (\sigma_x, \sigma_y) \leq (X_s,\ Y_s) \leq (m_x, m_y) + (\sigma_x, \sigma_y).$$

(3) fcao means residual lime (Cao) in clinker, and is concerned with quality of cement.

(4) oil-l/t means productivity of fuel (liter per ton of clinker).

(5) N200, kiln revolution speed, is concerned with amount of production.

* The object function was composed by using the eigen vector of first principal component.

testing of the reliability of the re-identified model and for the auto-tuning of the control gains seems very important.

The state of actual process may drift into a nonlinear domain under a special type of disturbances. In such a situation it seems that we have to consider the application of AI (Artificial Intelligence) technique which uses qualitative knowledge about the handling of the nonlinearities retained by a skillful human operator.

The decision of production level leads us to a choice of balanced set points which optimize cost productivity and quality of products. In conventional control a human operator must decide the set points of controlled variables subjectively. We proposed the procedure determining objectively the balanced set points of multi-variate system with LP method.

Acknowledgements

The successful application of time series analysis to cement plant processes was realized by the initiative of Dr. T. Nakagawa, President of System Sougoh Kaihatsu Company. We would like to express our gratitude to Professor H. Akaike of the Institute of Statistical Mathematics and Dr. T. Nakagawa, and finally to Chichibu Cement Company for permitting the publication of the material discussed in the paper.

REFERENCES

Akaike, H. (1968). On the use of a linear model for the identification of feedback systems, *Ann. Inst. Statist. Math.*, **20**, 425–439.

Akaike, H. and Nakagawa, T. (1973). *Statistical Analysis and Control of Dynamic Systems*, Saiensusha (in Japanese) (to be published in English).

Duda, D.-L. W. H. (1985). *Cement Data-Book*, 3rd ed., 206–209, Bauverlag GmbH. Wiesbaden und Berlin.

Hagimura, S., Saitoh, T., Yagihara, Y. and Kominami, T. (1986). The hierarchical control with stability and production level control of cement NSP kilns, preprint, *IFAC 5th MMM Congress, Tokyo*, 77–78.

Jury, E. I. (1958). *Sampled-Data Control Systems*, **26**, Wiley, New York.

Nakagawa, T. and Yagihara, Y. (1985). The approach to the design of optimum production level and pursuit control in industrial process, *J. SICE* (in Japanese), **24**, 77–83.

Nakamura, H. and Toyota, T. (1988). Statistical identifications and optimal control of thermal power plants, *Ann. Inst. Statist. Math.*, **40**, 1–28.

Otomo, T., Nakagawa, T. and Akaike, H. (1969). Implementation of computer control of a cement rotary kiln through data analysis, preprint, *Tech. Session 66, IFAC 4th World Congress, Warsaw*, 115–140.

Otomo, T., Nakagawa, T. and Akaike, H. (1972). Statistical approach to computer control of cement rotary kilns, *Automatica—J. IFAC*, **8**, 35–48.