Design and Analysis of Boiler–Turbine–Generator Controls Using Optimal Linear Regulator Theory

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Abstract—The demand for improved dynamic response of fossil-fired power plants has motivated a comprehensive program of control system design and analysis. Previous papers have reported the development of a nonlinear mathematical model of a drum-type, twin furnace, reheat boiler–turbine–generator (RBTG) system which is suitable for control system analysis and has been extensively verified by field test. On the basis of this model, local stability, observability, and controllability have been examined over the load range, using linearization and modal analysis. An approach to control system design has been developed based on optimal linear regulator theory and which recognizes the limitation of an imperfect model. This approach produces "integral-type" action which guarantees zero steady-state errors. The controller does not require complete state feedback. Improved performance has been demonstrated by comparison with the existing control structure through simulation using the nonlinear process model.

I. Introduction

THE CONTINUING increase in demand for electric power, unanticipated delays in new generating capacity additions, and the trend toward larger generating stations and larger interconnections are among the many factors which have magnified the importance of individual unit response capability to the power system operating objective of providing reliable and efficient electric service. During normal operation, good unit response capability is essential for stable implementation of the megawatt dispatch system load control concept [1]. In emergency situations, responsive generation can be coordinated for load pickup or rejection in order to avoid or minimize cascading of system disturbances [2]. It is essential that generating units have a sufficiently high degree of stability to be able to stick with the system through an emergency situation without unreasonable risk. Should isolation become necessary, the unit must be capable of controlled rejection of generation without complete shutdown in order to service its local loads and to be available for system restoration [3].

These system operating requirements conflict with the obvious desire to maximize the life and to avoid damage of enormously expensive and complex primary equipment. This is a particularly important concern at the present time when replacement generation is frequently not available or at best involves extremely high operating costs. In recent years, new information concerning turbine metal fatigue due to cyclic thermal stress [4], [5] has made this a major consideration in operating generating stations. Nevertheless, numerous reports [6]–[8] suggest, on the basis of test experience, that the primary equipment itself does not impose a serious inherent limitation on load change capability and that, with suitably designed automatic controls, the objectives of system operation can be met—consistent with unit safety and life requirements.

In [8], Durrant and Vollmer suggest a variety of alternative operating and control strategies for boiler–turbine–generator systems to meet different system operating objectives. Among these are nonstandard automatic control procedures such as using attemperating sprays to generate steam in the superheater for assisting load pickup, manipulating gas flow for control of temperatures, incorporating variable steam pressure operation to regulate turbine rotor temperature variations, and relaxing throttle temperature tolerances, also to obtain better control of rotor temperature. They note that the operating objectives are frequently conflicting with respect to a given procedure and suggest further investigation to clarify the implications of these alternatives for specific applications.

Optimization and simulation provide a framework particularly well suited to the identification and evaluation of alternative control strategies. There have been some previous attempts to apply optimal control theory to the control of a power boiler. Notable among these are the works of Nicholson [9]–[11] and Anderson [12]. Nicholson's use of an oversimplified boiler model has made his positive results essentially meaningless for large power boiler applications. Anderson's work, on the other hand, followed an extensive effort of model development [13]. In [12] Anderson concludes that integrated optimal control schemes do not significantly improve the performance of the unit considered.

Anderson's conclusions are contrary to the optimism generated for coordinated control schemes by test experience and are also subject to question on the basis that the model used is still not an adequate characterization of a typical power boiler. In the work reported herein, every effort has been made to avoid such criticism. The model used in these studies has been used to simulate the Philadelphia Electric Company's Cromby Number 2 unit and has been subjected to extensive comparisons with closed-loop steady-state and open-loop transient field tests [14], [15]. The model is nonlinear, and all manipulated variables normally considered for automated operation are included. Several rather subtle details which have been previously overlooked but which are critical to wide-range unit operation have been represented, such as multiple

Manuscript received September 1, 1972; revised January 26, 1973. Paper recommended by J. Peschon, Past Chairman of the IEEE S-CS Utility Systems Committee.

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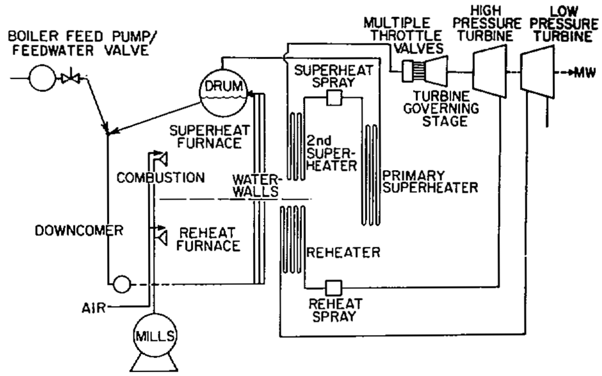

Fig. 1. Schematic diagram of Cromby Number 2 unit.

regulating valves, burner positions, and multiple feed-pumps.

From the outset of the study, the objective has been to go beyond identification and evaluation of alternative control strategies and to provide a feedback controller design suitable for implementation should it be warranted on the basis of simulation results. To achieve this end, the controller design methodology described in detail in [16] has been utilized. This procedure is based on optimal linear regulator theory but circumvents the practical deficiencies of standard results. In particular, the design incorporates a practical method of state reconstruction when there are a limited number of essentially noise-free outputs, retains the advantage of classical proportional integral (PI) controllers that steady-state accuracy is guaranteed even in the presence of immeasurable constant disturbances, and is not dependent upon unreasonable model precision.

In Section II, the plant, its nonlinear mathematical model, and some results of local linear analysis are discussed. Formulation of the overall control problem and a description of the current control system are included in Section III. In Section IV, the design algorithm is described in the context of the present application, and computer simulation results are discussed in Section V.

II. Mathematical Model

The aforementioned Cromby Number 2 unit is typical of a large class of power generating stations and has been used as the object of analysis for the studies reported in this paper. Cromby Number 2 is a 200-MW boiler–turbine–generator system which includes a pulverized coal-fired, twin furnace, drum-type, controlled circulation, single reheat boiler. In [15], the system (shown schematically in Fig. 1) was partitioned into subsections, to each of which were applied the requisite laws governing the transfer of energy and mass and the equations of state describing material properties. The resulting mathematical model consists of 14 first-order nonlinear differential equations and 70 nonlinear algebraic equations describing the variables of interest, many of which may be suppressed if desired. All plant parameters used in the model were obtained from physical data or calculated from acceptance test data.

TABLE I
State Variables

1) Superheat furnace metal temperature.
2) Reheat furnace metal temperature.
3) Drum water volume.
4) Drum steam density.
5) Primary superheater steam density.
6) Secondary superheater steam density.
7) Reheater steam density.
8) Primary superheater enthalpy.
9) Average secondary superheater enthalpy.
10) Secondary superheater outlet enthalpy.
11) Average reheater enthalpy.
12) Reheater outlet enthalpy.
13) Mass of coal in crusher zone of mill.
14) Fraction of total mill volume occupied by coal.

TABLE II
Control Inputs

1) Feedwater valve area.[a]
2) Governing valve area.[a]
3) Mill feeder stroke.[a]
4) Superheat spray flow.
5) Reheat spray flow.
6) Air flow.[a]
7) Superheat furnace burner tilts.[a]
8) Reheat furnace burner tilts.[a]

[a] Used by existing control system.

TABLE III
Output Variables (Partial List)

1) Generation.[a]
2) Throttle flow.[a]
3) Throttle pressure.[a]
4) Throttle temperature.[a]
5) Reheater outlet flow.[a]
6) Reheater outlet pressure.
7) Reheater outlet temperature.[a]
8) Drum pressure.
9) Gas flow.
10) Impulse chamber pressure.
11) Impulse chamber temperature.
12) Primary superheater outlet flow.
13) Primary superheater outlet pressure.
14) Primary superheater outlet temperature.
15) Reheater inlet (cold reheat) pressure.
16) Reheater inlet (cold reheat) temperature.
17) Coal flow rate.
18) Drum level.[a]
19) Feedwater flow.[a]
20) Average secondary superheater temperature.
21) Average reheater temperature.

[a] Used by existing control system.

The mathematical model is described by the equations

$$\dot{x} = f(x,u,y) \qquad 0 = g(x,u,y) \tag{1}$$

where x is the 14-dimensional state vector, u is the 8-dimensional control vector, and y is the 70-dimensional output vector as defined in Tables I–III.

Local properties of the nonlinear system have been examined over the process load range by generating approximate linear models at the desired steady-state operating points in the following form:

$$\dot{x} = Ax + Bu \qquad y = Cx + Du \tag{2}$$

where (x,y,u) in (2) represent deviations from the steady-state operating values (x_0,y_0,u_0).

The variation of the linear model eigenvalues with load along a typical steady-state operating profile is shown in Fig. 2. Illustrated are 13 of the 14 eigenvalues. The remaining eigenvalue is always zero. The arrow points in the direction of decreasing load. The seemingly erratic behavior of the eigenvalues is a result of the highly nonlinear valve characteristic. Examination of the eigenvectors leads to a few general observations. The zero eigenvalue is associated with the drum water volume. Drum water volume is affected by every mode which is consistent with the knowledge that water level requires tight regulation. There is generally a very high degree of coupling between the state variables. Two modes are clearly identifiable with the mill dynamics.

Local observability and controllability have been examined. If feeder stroke is not available as a control input, then the two mill modes are uncontrollable. The system is controllable, however, even when superheat and reheat sprays are not used as control inputs. Drum level must be measured in order to have an observable situation. Otherwise, almost any selection of outputs will suffice.

III. Operating Objectives and Existing Controls

The principal operating objectives can be summarized in the following statement:

> The control system should provide for maximum rate of change of generator output from the initial state to an assigned target state without exceeding specified limits on process variables.

The limits currently specified for Cromby Number 2 on the key output variables are given in Table IV.

There is a serious need for a basic evaluation of what constitutes tolerable variations of these process variables. If the constraints are set too loose, then the risk of equipment damage is great; and, if they are set too tight, a low value of the maximum rate of change of generation will be determined for the unit. It is obvious that excessively high pressures constitute a safety hazard and that excessive or widely varying steam temperatures should be avoided because of the close clearances in the turbine and the possibility of metal fatigue due to cyclic thermal stress.

However, the precise specification of acceptable limits is perhaps somewhat arbitrary. For example, there has been considerable discussion within the industry of variable pressure operation [17] in which case operating pressures as much as 400 psia below nominal are advocated, and the plant is normally operated at low load with as much as a 100°F drop in throttle temperature. Such practices contradict the limits of −50 psia on throttle pressure and −10°F on throttle temperature. The emergency lower limit on temperature (−200°F) is more realistic. However, these constraints have been set and, until a convincing evaluation is made which shows justification for changing them, they must be adhered to in any control design.

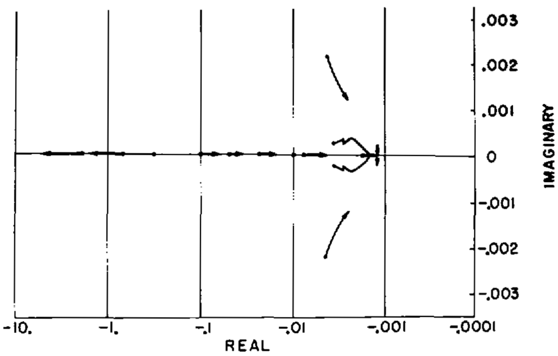

Fig. 2. Root locus for RBTG system. Dot represents high load; arrow represents low load.

TABLE IV
Cromby Number 2 Unit Output Variable Constraints

	Nominal Values	Deviations	
		Normal	Emergency
Throttle pressure	1825 psia	±50 psia	±75 psia
Throttle temperature	1000°F	±10°F	+20°F −200°F
Reheat temperature	1000°F	±10°F	+20°F −200°F
Excess oxygen	4.4%	2–5%	1.5–6%
Drum level	0 in	±3 in	±4 in

TABLE V
Cromby Number 2 Unit Control Input Constraints

	Minimum Value	Maximum Value
Feedwater valve normalized area	0	1
Governing valve normalized area	0	8
Normalized feeder stroke	0	1
Superheat spray flow	0	60 klb/h
Reheat spray flow	0	60 klb/h
Air flow	180 lb/s	600 lb/s
Superheat burner tilts	−30°	+30°
Reheat burner tilts	−30°	+30°

The control constraints are tabulated in Table V. These constraints are based on physical limits of travel for valve actuators or on maximum equipment capacity. The minimum value for air flow insures safe furnace conditions at low load.

Both sprays and tilts are provided for temperature control. Sprays are used to supplement the tilts which normally are the primary means of temperature control. This additional capability to prevent temperatures from becoming too high is consistent with the concern over excessive thermal stress and close turbine clearances.

The existing Cromby Number 2 unit control system is composed of five distinct control loops. These include: the power generation control loop, Fig. 3; the fuel control loop, Fig. 4; the air control loop, Fig. 5; the steam temperature control loop, Fig. 6 (there are actually two identical controllers—one for superheat, the other for reheat temperatures); and the drum level control loop, Fig. 7. The mathematical characterizations shown are, of course,

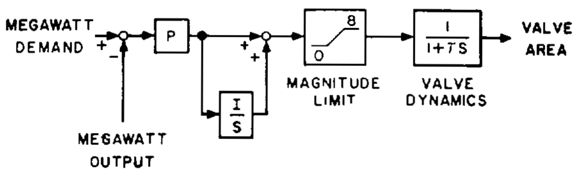

Fig. 3. Power generation control loop model, Cromby Number 2 unit. $P = 0.03$; $I = 0.4$; and $\tau = 10$.

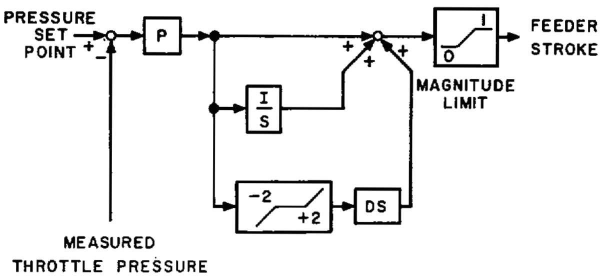

Fig. 4. Fuel control loop model, Cromby Number 2 unit. $P = 0.001$; $I = 0.003$; $D = 120$.

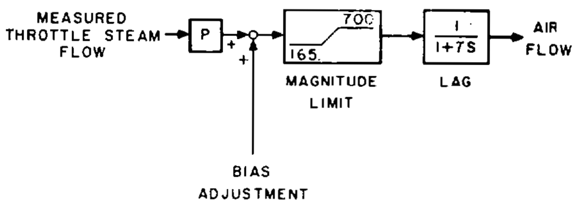

Fig. 5. Air flow control loop model, Cromby Number 2 unit. $P = 1238.26$; $\tau = 10$.

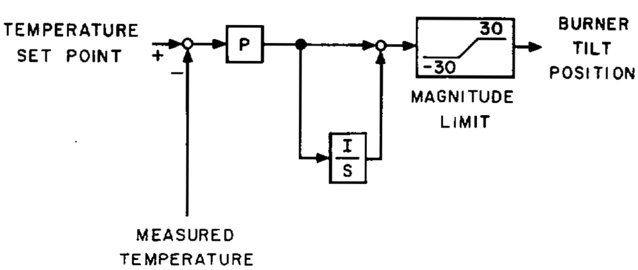

Fig. 6. Steam temperature control loop model, Cromby Number 2 unit. $P = 1.5$; $I = 0.01$.

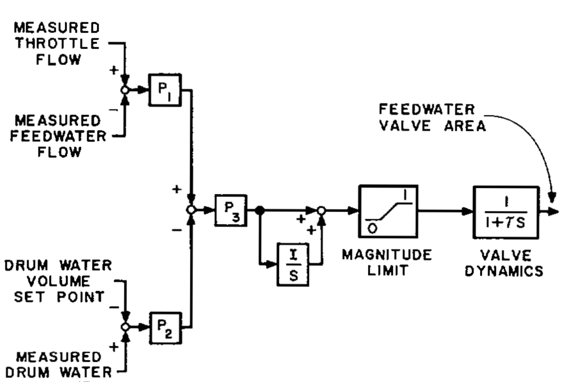

Fig. 7. Three-element feedwater control loop model, Cromby Number 2 unit. $P_1 = 10$; $P_2 = 0.007$; $P_3 = 0.01323$; $I = 5$; $\tau = 10$.

idealizations of the actual controls. Closed-loop simulation shows, however, that the system is a very good representation of the actual unit.

IV. Controller Design

Conventional power generation control systems have evolved with time and have independent feedback control of key process variables. These regulators were designed to hold the process variables at a fixed desired value. Today it is recognized that such control systems are inadequate for load tracking, and manufacturers are now including feedforward features which will change set points as a function of demand input. Even the most advanced of these conventionally designed control systems cannot handle this highly interactive process under rapid load change in a satisfactory manner.

Modern control theory provides the techniques for the design of dynamically integrated control systems for multivariable processes. The methodology proposed in [16] has been utilized to design a control system for the boiler–turbine–generator system of interest. The design process is briefly described below in somewhat less general terms than provided in [16]. An important feature of the procedure is that, by including a simple characterization of model error, the resultant controller retains the steady-state accuracy of classical PI controllers.

The feedback controller is designed on the basis of an approximate linear model obtained at a preselected steady-state operating point. The model used for design takes the following form:

$$\begin{aligned} \dot{x} &= Ax + Bu \\ \dot{w} &= \nu \\ y &= Cx + Du + w \end{aligned} \tag{3}$$

where x is an n-dimensional state vector, y is a p-dimensional output vector, u is an m-dimensional input vector, and w is a p-dimensional bias vector. The bias noise ν is a white-noise process having zero-mean and covariance $V_\nu \delta(t)$.

The random bias vector w has been specifically introduced to represent model inaccuracies. As the objective is the synthesis of an optimal deterministic controller, the limiting form as V_ν vanishes is of particular interest. In this case, the bias vector becomes a constant, but *a priori* unknown, bias.

The objective is to steer the system so that y tracks a constant desired value $\bar{y}$ while u varies moderately about some nominal value. To obtain an appropriate cost functional, consider w to be a constant. In this case, the following steady-state conditions on (x,u) with $y = \bar{y}$ are obtained from (3) with $\dot{x} = 0$:

$$0 = A\bar{x} + B\bar{u}, \qquad \bar{y} = C\bar{x} + D\bar{u} + w. \tag{4}$$

In the regulator problem there exists at least one solution $(\bar{x},\bar{u})$ for each $(\bar{y},w)$. In this case, (4) must consist of no more than $n + m$ independent equations. Equations (4) can be arranged in the form

$$\begin{bmatrix} A & B \\ C & D \end{bmatrix}\begin{bmatrix} \bar{x} \\ \bar{u} \end{bmatrix} = \begin{bmatrix} 0 \\ \bar{y} - w \end{bmatrix}.$$

If, in addition, the coefficient matrix is of full rank (which, in this case, guarantees that the number of rows are not greater than the number of columns), then a solution for $(\bar{x},\bar{u})$ is

$$\begin{bmatrix} \bar{x} \\ \bar{u} \end{bmatrix} = \begin{bmatrix} A & B \\ C & D \end{bmatrix}' \left\{\begin{bmatrix} A & B \\ C & D \end{bmatrix}\begin{bmatrix} A & B \\ C & D \end{bmatrix}'\right\}^{-1}\begin{bmatrix} 0 \\ \bar{y} - w \end{bmatrix}. \quad (5)$$

Upon partitioning the solution, $\bar{x}$ and $\bar{u}$ are obtained in the form

$$\bar{u} = U(\bar{y} - w) \qquad \bar{x} = X(\bar{y} - w). \quad (6)$$

A quadratic cost functional can be defined as

$$J_T = (y - \bar{y})'Q_0(y - \bar{y}) + \int_0^T \{(y - \bar{y})'Q(y - \bar{y}) + (u - \bar{u})'R(u - \bar{u})\}dt \quad (7)$$

where Q_0, Q are nonnegative definite and R is positive definite. It is desired to minimize

$$\lim_{T\to\infty} \frac{1}{T} E\{J_T\}.$$

In composite form the system (3) can be written as

$$\dot{x}_1 = A_1x_1 + B_1u_1 + G\nu$$
$$y_1 = Hx_1 + Du_1 \quad (8)$$

where

$$x_1 = \begin{bmatrix} x - X\bar{y} \\ w \end{bmatrix}, \qquad u_1 = [u - U\bar{y}], \qquad y_1 = y - \bar{y}$$
$$A_1 = \begin{bmatrix} A & 0 \\ 0 & 0 \end{bmatrix}, \qquad B_1 = \begin{bmatrix} B \\ 0 \end{bmatrix}, \qquad G = \begin{bmatrix} 0 \\ I_m \end{bmatrix}$$
$$H = [C \quad I_m]. \quad (9)$$

The optimal controller as obtained in [16] is

$$u_1^* = -M\hat{x}_1, \qquad M = [K \,\vdots\, U + KX] \quad (10)$$

where

$$K = (D'QD + R)^{-1}(B'S + D'QC) \quad (11)$$

and S satisfies the Riccati equation

$$0 = \{A - B(D'QD + R)^{-1}D'QC\}'S + S\{A - B(D'QD + R)^{-1}D'QC\} - SB(D'QD + R)^{-1}B'S + \{C'QC - C'QD(D'QD + R)^{-1}D'QC\}. \quad (12)$$

The state estimate x_1 is obtained from

$$\hat{x}_1 = H^*(y_1 - Du_1) + \Theta_2\xi$$
$$\dot{\xi} = \Gamma_4\xi + \Phi_2(y_1 - Du_1) \quad (13)$$

where the parameters are defined below.

The matrix H^* is given by

$$H^* = (GV_\nu G' + P_0A_1')V_\nu^{-1} \quad (14)$$

where P_0 satisfies the Riccati equation

$$P_0A_1'[I - H'G'] + [I - GH]A_1P_0 - P_0A_1'H'V_\nu^{-1}HA_1P_0 = 0. \quad (15)$$

The matrix Θ_2 is given by

$$\Theta_2 = \begin{bmatrix} I_n \\ -C \end{bmatrix}. \quad (16)$$

If Λ_n denotes a matrix whose rows are a set of n linearly independent rows of $I - H^*H$ and $\tilde{A}_1$ denotes the matrix

$$\tilde{A}_1 = A_1 - B_1M, \quad (17)$$

then Γ_4 and Φ_2 are given by

$$\Gamma_4 = \Lambda_n\tilde{A}_1\Theta_2, \qquad \Phi_2 = \Lambda_n\tilde{A}_1H^*. \quad (18)$$

From the results of [16] it is known that the $2n + p$ eigenvalues of the closed-loop linear system include the p zero eigenvalues corresponding to the bias variables w, the n stable eigenvalues corresponding to the closed-loop system matrix $\tilde{A} = A - BK$, and the n stable eigenvalues of the matrix $A_nA_1\Theta_2$ which are associated with the observer. Moreover, a prescribed degree of stability α for the observer eigenvalues is attained by replacing A_1 in (14), (15) by $A_1 + \alpha I$. If α is zero and the system matrix A is stable (which is the case for the reheat boiler–turbine–generator (RBTG) system), then $P_0 = 0$ and H^* and Λ_n specialize to

$$H^* = \begin{bmatrix} 0 \\ I_p \end{bmatrix}, \qquad \Lambda_n = [I_n \quad 0]. \quad (19)$$

Note that an alternate form of (13) which has certain advantages for application is

$$\dot{\xi} = \Lambda_nA_1\Theta_2\xi + \Lambda_nB_1u_1 + \Lambda_nA_1H^*(y_1 - Du_1). \quad (20)$$

This allows the estimator to use the actual applied control inputs, which is particularly important when the system controls saturate. For the case $P_0 = 0$, (20) specializes to

$$\dot{\xi} = A\xi + Bu_1. \quad (21)$$

To apply this controller to the actual nonlinear process, the steady-state process outputs and controls are characterized as functions of the megawatt demand (MWD). Then, the linear feedforward elements can be replaced by the actual nonlinear relationships

$$\bar{u} = g(\text{MWD}) \qquad \bar{y} = f(\text{MWD}). \quad (22)$$

The resultant controller is illustrated in Fig. 8.

V. Computer Simulation Results

The control design methodology described above has been developed into a convenient and flexible digital computer program. This program is used in conjunction with the RBTG linear analysis program and the RBTG nonlinear simulation program to design and analyze RBTG controls. A typical case study can include the following steps.

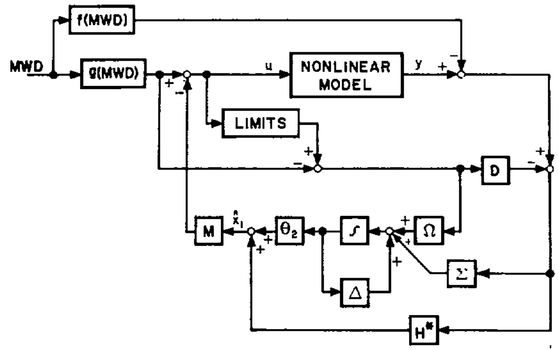

Fig. 8. Block diagram of closed-loop nonlinear system with state estimator. $\Delta = \Lambda_n A_1 \Theta_2$; $\Sigma = \Lambda_n A_1 H^*$; $\Omega = \Lambda_n B_1$.

Step 1: Specify a desired steady-state profile. (This can be done with the aid of the simulation program.)

Step 2: Specify a nominal operating load level for controller design. Use the linear analysis program to obtain the A, B, C, and D matrices.

Step 3: Specify which controls and outputs are to be used.

Step 4: Specify cost functional weighting matrices and observer degree of stability.

Step 5: Execute controller design program.

Step 6: Execute simulation program.

Current plant operating practice is to use the steady-state profile reported in [14]. However, superheat and reheat burner tilts are positioned at their positive limit above approximately 200 MW, and consequently it is not possible to regulate temperature. This difficulty can be circumvented by adjusting the excess air flow. This has been done to develop a steady-state operating profile for the load range 130–230 MW, which provides for superheat and reheat steam temperatures of 1000°F with burner tilt positions suitably interior to their constraints. Of particular note is the nonlinear characteristic of the governing valves as shown in Fig. 9.

The control inputs and process outputs used in the existing control system were used in the design of the optimal control system. The control inputs are: 1) feedwater valve area; 2) governing valve area; 3) mill feeder stroke; 4) superheat furnace burner tilts; 5) reheat furnace burner tilts; and 6) air flow. The process outputs used for control are: 1) generation; 2) throttle flow minus feedwater flow minus (coefficient) drum level error; 3) throttle pressure; 4) throttle temperature; and 5) reheater outlet temperature.

The first case study was made with all weightings set to unity, and the weightings were then adjusted by observing the ability of the control system to keep the process outputs within the specified constraints. The optimal control system which is used in the following comparison with the existing control system is defined by the set of weightings given in Table VI.

The conventional control system, the state variable feedback system, and the state estimator system were simulated. The state variable feedback system was investigated to establish the ultimate potential for improvement.

Figs. 10–12 compare the response of the three control systems to a 10-MW step decrease in the megawatt demand. Fig. 13 compares the response for a 25-MW step decrease.

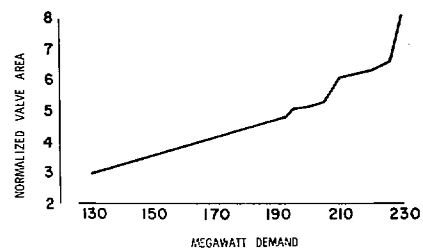

Fig. 9. Cromby Number 2 unit steady-state profile for governing valves.

TABLE VI

OPTIMAL CONTROL SYSTEM WEIGHTINGS

Variable	Weighting Value
Feedwater valve area	0.2
Governing valve area	4×10^{-10}
Mill feeder stroke	0.15
Superheat burner tilts	0.2
Reheat burner tilts	0.02
Air flow	20
Generation	4000
Throttle flow minus feedwater flow minus (0.28) drum level error	1
Throttle pressure	6000
Throttle temperature	400
Reheater outlet temperature	100

VI. CONCLUSIONS

This paper reports the development of a methodology for the design and analysis of multivariable process controls and its application to the control of conventional, drum-type, fossil-fired, single reheat steam power plants. The design methodology is based on optimal linear control theory, incorporates feedforward, provides a method of state reconstruction, retains the steady-state accuracy advantage of classical PI controllers, and is not dependent upon unreasonable model precision.

The application to the design and analysis of RBTG system control has been successful, and the question as to whether or not the application of modern control theory can improve the control of fossil-fired RBTG systems can

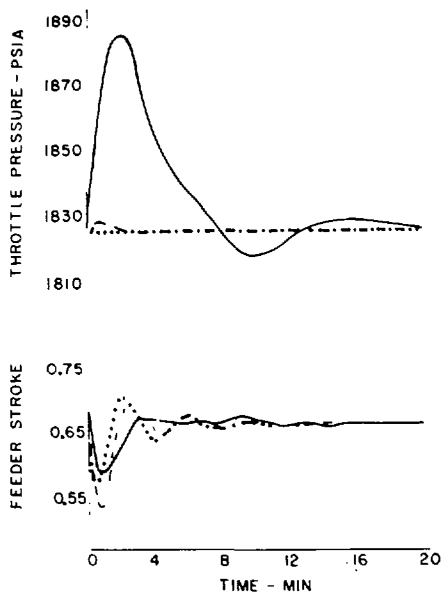

Fig. 10. Throttle pressure and feeder stroke response to a 10-MW step decrease. Solid line denotes conventional; broken line denotes state estimator; dotted line denotes state feedback.

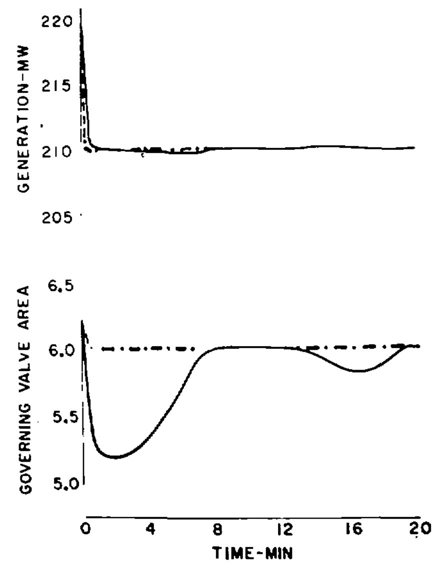

Fig. 11. Generation and governing valve response to a 10-MW step decrease. Solid line denotes conventional; broken line denotes state estimator; dotted line denotes state feedback.

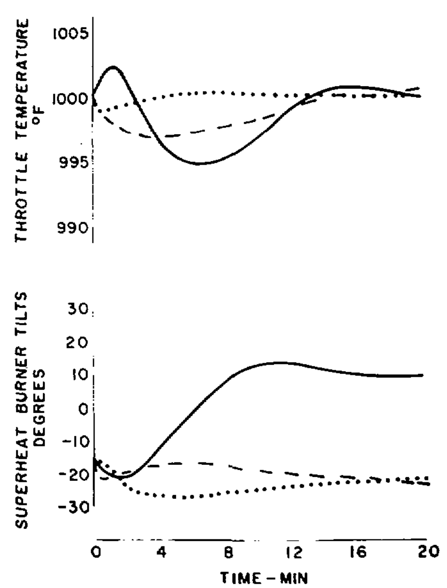

Fig. 12. Throttle temperature and superheat burner tilt response to a 10-MW step decrease. Solid line denotes conventional; broken line denotes state estimator; dotted line denotes state feedback.

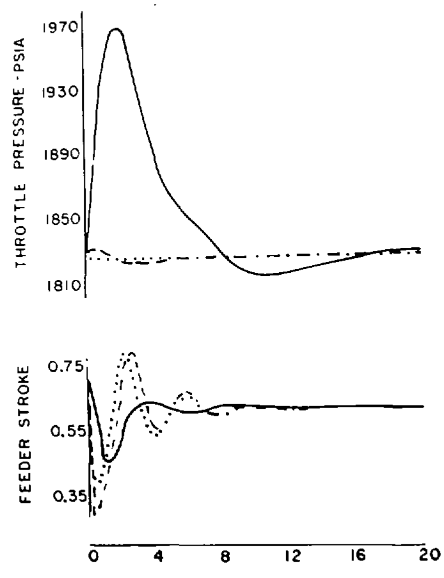

Fig. 13. Throttle pressure and feeder stroke response to a 25-MW step decrease. Solid line denotes conventional; broken line denotes state estimator; dotted line denotes state feedback.

be answered in the affirmative. The improvement in dynamic response which can be obtained is shown to be quite significant. Perhaps the most dramatic result is the tight regulation of pressure which is accomplished without significant increase in control action. This arises principally through coordination of fuel flow, air flow, and burner tilts. It is interesting to note that the optimal regulator takes advantage of the natural slowness of boiler temperature dynamics and manipulates air and burner tilts—normally associated with temperature control—to assist in regulating the faster pressure dynamics before the temperature transient becomes significant.

The simulation results were obtained using the nonlinear boiler model and the control system performed well even in the face of the quite irregular characteristics of the multiple governing valves. It should be noted that the plant characteristic would actually be smoother as the simulations were run with no valve overlap, which would not be the case in the field.

It is interesting to note that the pressure responses with the state estimator and state feedback systems are quite close, whereas there is significant difference between the corresponding temperature responses. This is to be expected as the process dynamics are directly reflected in the observer and the temperature dynamics are relatively slow. By including a degree of stability specification for the observer, however, the estimator tracking of the slow modes can be "sped up" to any desirable rate so that performance of the state estimator system approaches that of the state feedback system.

The approach to design afforded by modern control theory, exemplified in this paper, makes it particularly easy to investigate the effect of using different combinations of controls and outputs, the effect of emphasizing the importance of regulating specific process variables (such as first-stage temperature), and the effect of restricting the use of specific control variables.

Acknowledgment

The authors wish to acknowledge the advice and support of their colleague L. H. Fink throughout the course of this project.

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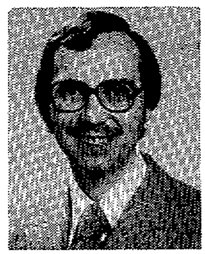

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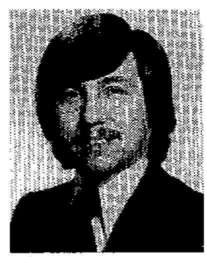

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