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Moving-Bed Gasification-Combined-Cycle Control Study.
Volume 2. Results and Conclusions, Case 2—Oxygen-Blown, Slagging-Ash Operation

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ABSTRACT

A computer simulation study has been conducted to investigate the process dynamics and control strategies required for operation of an oxygen-blown, slagging, moving-bed gasifier combined cycle (GCC) power plant in a utility power system. The gasifier modeled is of the modified Lurgi type as developed by the British Gas Corporation. This study is a continuation of a study on moving-bed GCC control analysis. Work reported on previously (EPRI report AP-1740) was for an air-blown, dry-ash Lurgi GCC power plant and results are compared to this study.

The simulated GCC plant configuration is similar to that developed in earlier EPRI economic studies (EPRI report AF-642). The computer model used in the air-blown, dry-ash GCC study was re-configured to represent the oxygen-blown slagging GCC cleanup process and a new gasifier model included. Gas turbine-lead and gasifier-lead control modes were evaluated with respect to power system dynamic requirements. The effect of gasifier output fluctuations, as observed in actual gasifier process development unit operation, was modeled and investigated.

In comparison to the air-blown GCC power plant, the oxygen-blown fuel process and power generation process are not as integrated, resulting in less system interaction and reduced difficulty of control. As concluded in the air-blown GCC system study, the turbine-lead control mode is the preferred control strategy because it can effectively meet power system requirements. The large storage volume of the cleanup system is used to advantage and control of the combined cycle is maintained close to that of a conventional-fueled combined cycle. The oxygen-blown system is more responsive than the air-blown system system and can successfully meet power system requirements.

EPRI PERSPECTIVE

PROJECT DESCRIPTION

This report, Moving-Bed Gasification-Combined-Cycle Control Study--Volume 2: Results and Conclusions, Case 2--Oxygen-Blown, Slagging Ash Operation, is the second and concluding part of a computer simulation analysis of process dynamics and control for advanced coal gasification-based power plants conducted under RP914-1. The first phase of the study (see EPRI Final Report AP-1740, Volume 1) was devoted to an analysis of gasification-combined-cycle (GCC) plants based on moving-bed gasifiers with dry-ash operation (Lurgi type). In the second phase, attention was shifted to the analysis of GCC plants based on moving-bed coal gasifiers with slagging operation (British Gas Corporation-Lurgi type).

EPRI Final Reports AF-642 and AP-753 document the economic advantages offered by the slagging, moving-bed gasifier for electric power generation. Experimental studies by British Gas Corporation (BGC) verify the steady-state and dynamic performance potential of the slagging gasifier operation on U.S. caking coal from the Pittsburgh seam (see EPRI Final Report AP-1922).

Results of a similar control study for a GCC plant based on an entrained flow gasifier have been reported by Fluor Engineers and Constructors, Inc., and Westinghouse Electric Corporation (see EPRI Final Report AP-1422).

PROJECT OBJECTIVES

The GCC plant control studies were intended to evaluate the operability of these advanced power plants in which moving-bed gasifiers and associated fuel process components are closely integrated with the gas and steam turbines in a combined-cycle process. Inherent characteristics were to be determined for dynamic operation of the GCC plants, and preferred control strategies were to be recommended.

PROJECT RESULTS

Simulation models were formulated for the GCC plant based on the slagging, moving-bed gasifier, using flowsheets originally conceived for economic evaluations for

EPRI by Fluor Engineers and Constructors (see AF-642). Since no power plants have yet been built with that type of gasifier, the computer analyses reported here are for screening purposes only, and they remain to be verified experimentally. Nevertheless, they have demonstrated the capability for analyzing process transients and designing process control systems.

As in the earlier study of dry-ash, moving-bed GCC plants, the process dynamics for the overall plant with the slagging gasifier appear to be uniquely determined by the inherent characteristics of the gasifier. The principal difference between the dry-ash and slagging reactors lies in the makeup of the blast stream, i.e., the steam and oxidant feed streams. The dry-ash unit for the earlier study is air-blown and consumes large quantities of steam to cool the ash below the fusion temperature; i.e., uses steam well in excess of that needed for the steam-carbon reactions. The slagger, on the other hand, is oxygen-blown and uses far less steam since it achieves higher temperatures to intentionally slag the coal ash. Without the introduction of nitrogen from the air and with a much lower steam flow, the slagger achieves a significantly higher throughput. In the earlier study the large steam consumption of the dry-ash gasifier was the source of anomalous transients at the steam turbine, which appeared initially to counter load-change demands. Such interactions between the fuel process and the turbine systems are greatly reduced in the analyses of the slagger GCC plant.

The turbine-lead control strategy appeared again to be the preferred control mode, as in the earlier phase, with the large inherent volume of the cleanup system appearing advantageous. However, the gasifier-lead strategy was not effective, again similar to Phase I, and the large cleanup system volume appeared to act as an impediment. Interestingly, the gasifier-lead mode of control has only appeared beneficial in the studies of the entrained flow GCC plant by Fluor and Westinghouse (see AP-1422) and later duplicated in power system integration studies by Philadelphia Electric Company (see EPRI Final Report AP-2053, Volumes 1 and 2). The results of this study imply that the cleanup system volume, if large, can be an important factor governing the effectiveness of this particular strategy. This may also hold for entrained flow GCC plants. Experimental tests of these alternative control strategies will be included as part of the test plan for the entrained GCC demonstration plant under the Cool Water Gasification Program (see EPRI Interim Report AP-2487).

Another concern addressed in this study was the effect of irregular heating value output variations as experienced in experimental trials of the BGC-Lurgi slagging gasifier at the BGC Westfield Development Center (see AP-1922). However, the simulated response was well modulated by the action of the plant controls in the turbine-lead mode and the attenuating effect of the large cleanup system volume. Therefore, variations of the fuel gas-heating value at the gas turbine inlet were minimal.

The computer simulation models from these GCC plant studies will have an impact on further development of this technology. For example, they formed the basis of later models used for the design of control systems in the Cool Water GCC demonstration plant. Similarly, they provided background for the study of an integrated gasification test facility at the General Electric Corporate Research Center under a study sponsored by the DOE.

For those engineers who may wish to obtain copies of the computer simulation codes for the various moving-bed or entrained flow GCC plants, simplified versions of such programs are available from the Electric Power Software Center. Descriptions of these programs, as developed by Philadelphia Electric in a related project, are covered in AP-2053.

This report should be of general value to those examining GCC plant applications but of more particular interest to those studying the load-following capabilities of moving-bed GCC plants.

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SUMMARY

EPRI studies (1, 2) have identified coal gasification-combined cycle (GCC) power plants as a potential economic and environmentally attractive means of coal-based electric power generation. An important consideration for any future source of electric power generation will be the ability to meet dynamic response requirements imposed by interconnected power system networks. An electrical network imposes load demands on individual generating plants to satisfy requirements for:

- Frequency regulation
- Tie-line flow regulation and thermal backup
- Daily load-following

These demands cover a wide spectrum of response magnitude and duration. Industry standards provide guidelines for acceptable performance of generating plants based on known power system requirements, as shown on Figure 6-1 and summarized in Table 6-1 of (3). For reliable power system operation, individual plants must perform the required dynamic maneuvers while using the equipment and premium fuels efficiently.

It is important to establish the capability of such new plants for dynamic operation and to develop successful plant control strategies. Prior to actual process design and construction, analysis of process dynamics and control can be accomplished using computer simulation. Such a control study has therefore been undertaken based on simulation of a gasification-combined cycle (GCC) power plant with the following objectives:

- Obtain responsive plant control while maintaining critical process variables within design limits for safe and reliable operation.
- Identify necessary control procedures for contingency operation under emergency conditions.

- Evaluate capability of the resulting power plant response characteristics to satisfy power system requirements.

This report comprises the second volume of the study of moving-bed coal gasification-combined cycle (GCC) plant controls. The first volume (3) presented results and conclusions for a dynamic simulation and control analysis of an air-blown dry-ash Lurgi GCC plant and utility system and identified power plant operational requirements and plant protective logic and operational procedures under emergency conditions. This volume presents results for a dynamic simulation and control study of an oxygen-blown slagging gasifier of the modified Lurgi type as developed by the British Gas Corporation and compares simulation results to the earlier air-blown dry-ash GCC study. Previous EPRI studies (4, 5) have investigated entrained GCC plant control and operation of both entrained and moving bed GCC plants within a utility network.

The oxygen-blown moving-bed GCC plant, as configured for this study, has a nominal rating of 240 megawatts (MW) when burning medium Btu fuel gas of ~350 Btu/SCF. Two state-of-the-art gas turbines are represented, each rated as 85 MW and operating at 2000 degrees F turbine inlet temperature with 11:1 pressure ratio. A single automatic-extraction, condensing steam turbine has a nominal 70 MW rating. The fuel process consists of one gasifier and cleanup train and an over-the-fence oxygen supply. A low temperature, Selexol-type physical absorption system is used for sulfur removal from the gas stream; i.e., selectively removing hydrogen sulfide (H_2S) from the clean fuel stream while leaving carbon dioxide (CO_2) as a working fluid for the gas turbine. The plant is configured with commercial-size equipment to represent the characteristics of large integrated plants, including the interaction and sequencing of operations under various loads and conditions.

THE SIMULATION APPROACH

To evaluate control strategies, the digital computer simulation model used for the air-blown dry-ash GCC plant was modified for oxygen-blown slagging operation. The plant simulation model was developed from basic physical laws of conservation of mass, energy and volume, with appropriate simplifying assumptions for heat and mass transfer dynamics of multi-phase, multi-component mixtures. A lumped parameter model was derived and, where accuracy demanded, a particular section was represented as several lumped volumes connected in series. Individual equipment models were general in nature and were simply used in a different configuration to model the oxygen-blown fuel gas system.

The degree of model detail was sufficient to achieve the control study objectives in an efficient and economic manner. (Cost of a fifteen minute, real-time run of the entire simulation model was less than fifty dollars.) The emphasis was on modeling important dynamic effects with reasonable accuracy. The model was intended primarily for control analysis and not as a process design tool; therefore, a detailed multi-dimensional representation has not been developed and there has been no attempt to optimize plant performance.

Gasifier

The gasifier model used was adapted from a moving-bed gasifier model developed by K.J. Daniel of the General Electric Company Corporate Research and Development Center and described in (6) and reprinted in Appendix A with the permission of the A.I.Ch.E. The gasifier model is based on a simultaneous solution of a set of equations derived from fundamental chemical, thermodynamic and heat and mass principles. Reaction kinetics and other time-dependent phenomena are included. Kinetic coefficients, gasifier geometry, and heat transfer coefficients used by the model were adjusted at steady-state conditions to represent the oxygen-blown slagging gasifier and provided a reasonable match with published data on gasifier performance.

Cleanup System

The cleanup system model developed for the air-blown dry-ash GCC simulation was configured for use with the oxygen-blown slagging GCC. The model includes components which directly impact fuel gas properties and have a significant effect on power plant performance.

Included were the following:

- Wash Cooler
- Product Gas Exchanger
- Gas Cooler
- Trim Cooler
- H_2S Absorber

The overall cleanup system model is structured to accept gas flow, temperature, and gas composition from the gasifier model. In the H_2S absorber, differing amounts of H_2S and CO_2 are absorbed; therefore, at the tower exit the dry gas composition is recomputed. Gas properties, such as specific heat, molecular weight, and heating value are calculated in the model as a function of the variable gas composition, temperature and pressure as the gas passes through each component of the cleanup system. The residence time of gas in the cleanup system is significant as a result of its large volume. A sizeable transport lag in composition and temperature therefore occurs across the cleanup system.

Combined Cycle

The combined cycle power plant includes gas turbines, heat recovery steam generators, and a steam turbine. The models are derived from fundamental thermodynamic relations with energy and mass balance equations. Representative hardware data is used for the simulation.

Gas Turbine

Models for compressor, combustor, and turbine are derived from basic governing equations. Air flow is calculated as a function of speed and pressure ratio with an appropriate correction factor for ambient conditions.

The model is capable of simultaneous operation on dual fuel. The combustor model provides for operation on either light fuel oil or a low Btu gas with variable composition.

Steam Turbine

Steam from two heat recovery steam generators flows into a common header to the steam turbine. The steam turbine is a single automatic extraction condensing unit with both high and low pressure sections. Steam is extracted from the turbine for the gasifier feed flow. For given input conditions the enthalpy drop across each section of the turbine is calculated using representative values of thermal efficiency, thereby determining the steam turbine power output and extraction steam conditions.

Heat Recovery Steam Generator

The superheater, evaporator, and economizer are the three heat transfer sections of the heat recovery steam generator (HRSG); the exhaust gas from the gas turbine flows across these three sections in a cross-flow arrangement. Due to the relatively low temperature of the turbine exhaust gas, convection is the only mode of heat transfer involved. A lumped parameter model is used to represent heat transfer effects.

Plant Control Systems

For a simple and efficient control study model, it is not deemed necessary to include all auxiliary subloops and secondary control loops in the overall plant control model. The only subloops included are those in which response and control directly affect fuel condition and plant output.

Computing absolute values of variables was not considered as important as predicting trends for initiating timely and appropriate control action. Judicious care was exercised to retain important dynamic effects to predict trends correctly during a transient. Also, since this was not a detailed design study, optimization of controller settings was not attempted.

A gas turbine control system including protective control logic was simulated. This includes speed-load control, temperature control, and inlet guide vane control. A pressure control system was also implemented when the systems were run in a gasifier-lead control mode.

The steam turbine control valves were on pressure control responding to steam flow changes and regulating throttle pressure at a 5% droop setting. The speed control was active only as a backup to provide overspeed protection for the steam turbine. Extraction steam pressure control was provided for the process steam to the gasifier.

SYSTEM COMPARISON

The significant differences between the oxygen-blown and air-blown moving-bed GCC systems are:

- Oxidant supply
- Process steam requirement
- Gasifier response characteristics
- Fuel gas moisture content

The oxygen-blown gasifier is supplied oxygen from an electrically-driven air separation plant. The air-blown gasifier is supplied air from gas turbine compressor extraction. The process steam requirement of the oxygen-blown system is much less than that required by the air-blown system (~0.3 lb steam per lb coal vs. ~1.4 lb steam per lb coal). The net effect of these differences is the reduction in the degree of integration of the fuel process and power generation process for the oxygen-blown system as compared to the air-blown system.

The heating value of the fuel gas from the air-blown gasifier varies significantly (plus or minus 10%) when the gasifier feed is changed due to behavior of volatile release. Also, moisture content of the fuel gas supplied to the gas turbine by the air-blown system varies causing the heating value to change. The heating value variations in fuel gas supplied to the gas turbine during a system transient are less than 1% for the oxygen-blown system.

CONTROL ANALYSIS

To provide baseline knowledge of GCC plant dynamic response, open-loop runs were made using the plant simulation model. Then automatic control was implemented to evaluate the overall closed loop response of a GCC plant. The gas turbine-lead control mode and gasifier-lead control mode were considered during the course of the study.

The control alternatives were evaluated to determine those which were likely to be successful. However, no attempt was made to seek an optimal control strategy. That is, a fully coordinated plant control mode was not developed.

Gas Turbine-Lead Control Strategy

As concluded in the air-blown dry-ash GCC study, the gas turbine lead control strategy gives the best dynamic performance in tracking load demand. Specifically, the turbine-lead control scheme was found to be successful for the operation and control of the air- or oxygen-blown moving bed GCC power plants. This judgment was based on a number of factors as follows:

- Average power system response requirements of 2% per minute can be satisfied.
- Growth potential exists for a faster response requirement.
- Overall output response is 30 to 50% faster than comparable gasifier-lead mode response.
- When supplemented by fuel system control modifications, it can effectively minimize fuel system and combined cycle transients.
- Basic gas turbine control philosophy is retained with only minor modifications.
- This control scheme also makes use of inherent storage of the fuel system during transient load changes.

An important advantage of the gas turbine-lead control strategy is that it can be achieved with a control system which is both familiar to and well-accepted by utility system operators.

Gasifier-Lead Control Strategy

This control strategy has a major drawback that prohibits its use as an effective control scheme. Plant power output in this mode is regulated by modulating the gasifier feed rate according to measured plant power output error by a proportional plus integral controller. The gas turbine fuel valves in this control mode open or close to maintain a set level of pressure. Pressure at the gas turbine fuel valve lags the gasifier considerably due to the capacitance effect of the large volume of the cleanup system, thus the plant power output considerably lags the demand.

CONCLUSIONS

The dynamic simulation of moving-bed GCC systems reported here is on the order of a screening analysis to investigate potential control strategies and to highlight any major obstacles to realization of moving bed GCC plant control capable of meeting power system requirements and provides a framework for future projects requiring evaluation of more detailed plant-specific configurations.

As concluded in the air-blown dry-ash GCC study (3), the process dynamics observed for the overall plant are uniquely determined by the gasifier operating characteristics. This is demonstrated through the comparison here between the air-blown and oxygen-blown moving-bed GCC systems. The oxygen-blown, slagging, moving bed GCC system is much less integrated than the air-blown system and control is less difficult. Fuel process and power system interactions are much reduced in the oxygen-blown system and control of the combined cycle is similar to that of a conventional fuel combined cycle.

The turbine-lead control mode is the preferred control strategy because it can effectively meet power system requirements. Control of the combined cycle in this mode is maintained close to that of a conventional combined cycle. The large storage volume of the cleanup system is used to advantage in this control mode.

The effect of gasifier output fluctuations as represented in this study do not appear to present a signifigant problem to the the plant control system in the turbine-lead control mode. The action of the control system and the attenuating effect of the large volume of the fuel cleanup system and associated piping hold variation of plant power output within an acceptable tolerance (±0.6%).

The gasifier-lead control mode has a serious disadvantage. Plant response lags demand by the time required for the pressure to change at the gas turbine fuel valve. Pressure changes at the gas turbine fuel valve lag changes in gasifier feed due to the large storage volume of the cleanup system. Thus the large storage volume of the cleanup system is not used to advantage in this control mode. Gasifier output fluctuations cause plant power output to vary by ±0.8%.

RECOMMENDATIONS

Dynamic simulation fits into the scheme of overall plant design as a tool to be used for evaluating:

- plant control strategy
- plant operations strategy
- plant transient performance

Dynamic simulation is effectively used in an iterative fashion throughout the plant synthesis and design cycle to evaluate the ability of hypothesized plant equipment configurations and control schemes to meet power system operational and transient performance requirements. It facilitates the prediction and evaluation of transient requirements of plant equipment and allows for the synthesis of coordinated controls used to minimize possible life shortening component duty and yet meet required response. With the advent of digital computer-based control systems, the dynamic simulation model may be used in the implementation of the actual plant control. Furthermore, the computer models developed and used in the plant dynamic simulation may also be used in a plant simulator for training of plant operating personnel.

It is with the perspective of the previous discussion on the application and benefits of process and control dynamic simulation that recommendations are made. Future EPRI efforts to facilitate the incorporation and effective use of process and control dynamic simulation in the entire plant design cycle as outlined above will be to the benefit of utility plant owners and operators. Ability of a plant to meet transient performance requirements and operational requirements would be fully considered in the plant design rather than incorporated as an afterthought. Effective design will require interaction between the plant designers and owner/operators to make trade-offs in costs, performance, reliability, control and operational considerations and can be realized through use of plant process and control dynamic simulation.

Verified Component Models

The end use of dynamic simulation results is limited by the accuracy of simulation models of the plant components. Component models also must represent component operational limits. It is recommended that models of plant components in appropriate detail, verified to be suitably accurate by actual component testing, be created and maintained to enable useful dynamic simulation.

Plant Specific Operation and Control

For a specific plant, the overall plant operations scheme and control need to be integrated. For example, it is possible that the plant could be operating at part load for a considerable time, and it may be advantageous from a heat rate point of view to turn down only one gas turbine (if the system power needs are met at this level of output). This, in turn, requires that a specific plant control scheme be formulated to operate the overall plant efficiently, i.e., while individual gas turbines are at different load points.

Also, for a specific plant, dynamic simulation can provide insight into off-normal operations following the loss of key process components. It is recommended that plant-specific simulation models of sufficient detail be developed for careful examination of such normal, near normal or contingency operations. Similarly, more comprehensive study of multi-unit and multi-train plant loading alternatives should be considered.

Coordinated Plant Control Strategy

There is potential benefit to coordinated control of the plant and its components. Using component models appropriate in level of detail and accuracy, a coordinated control scheme may be developed to yield the best possible results in terms of plant transient performance, operability and reliability. For example, severe pressure or temperature transients could jeopardize the reliabiltity of key plant equipment. By developing specific load-change schedules for key process components, a fully coordinated control strategy could evolve that would reduce the transient duty of these critical components and yet still meet overall plant transient performance requirements. It is recommended that coordinated control be studied on as near a realistic plant basis as possible and that benefits be identified.

Section 1

INTRODUCTION

Coal gasification combined cycle power plants offer a power generation alternative to the utility industry which is attractive in terms of both economics and environmental considerations. Heat rate and cost-of-electricity advantages result from a high degree of integration of fuel plant and combined cycle plant. This high degree of integration also gives rise to a concern regarding plant control and operation. The individual subsystems such as gasifiers, gas cleanup systems and combined cycle power plants have been operated commercially in different industrial applications as separate units. However, reliable operation and effective control of such integrated systems in power generation service (which remains to be demonstrated) are important for the commercial acceptance of new power generation schemes.

New fossil fuel power plants must be capable of responding dynamically to meet reqirements imposed by interconnected electrical networks. In the future, a large proportion of power system requirements may be met by advanced gasification-based GCC power plants. With a large number of such plants operating in a given power pool, transients associated with bringing a plant to new load points can lead to interaction among units connected to the system, interaction with automatic dispatching controls, and power system control difficulties. These new units must meet maneuvering requirements for the effective operation of the power system thereby assuring the most economic use of equipment and premium fuels. Therefore, at a very early point, it is important to establish the capability of such plants for dynamic operation and to develop successful plant control strategies. Prior to actual process design and construction, analysis of process dynamics and control can only be accomplished using computer simulation. Such a control study has therefore been undertaken based on simulation of a gasification-combined cycle (GCC) power plant with the following objectives:

- Obtain responsive plant control while maintaining critical plant process variables within design limits for safe and reliable operation.

- Evaluate the capability of the resulting power plant response characteristics to satisfy power system operational needs.

- Identify necessary override or protective logic and operational procedures under contingency conditions.

Computer simulation models of major plant components suitable for dynamic control studies were developed as an intermediate step in the study to meet the final objective of control strategy evaluation. The selected control systems investigated were configured for automatic control of major plant variables at both full and part load conditions. However, no attempt has been made to study control strategy or develop special logic for plant startup and shutdown.

A report previously published under this study (3) presented results from an investigation of the air-blown, dry-ash Lurgi gasification-combined cycle (GCC) power plant. Beyond GCC plant dynamic simulation, control analysis and utility system simulation, the study identified power plant operational requirements and plant protective logic and operational procedures under emergency conditions.

This report presents the results of a dynamic simulation study of a GCC plant based upon an oxygen-blown slagging gasifier of the modified Lurgi-type as developed by the British Gas Corporation. EPRI studies have shown that a GCC plant based on the oxygen-blown slagging gasifier appears economically competitive for electric power generation (1, 2). Tests have been conducted under another EPRI study (RP 1267) (7) by British Gas Corporation at the Westfield Development Center to obtain operating data under transient conditions typical of those required for electric power plants. This study also evaluates the impact on GCC system control of gasifier fluctuations identified from this testing and further described by Albrecht, (8). Simulation results are compared with those of the earlier air-blown dry-ash GCC study.

Section 2

SYSTEM DESCRIPTION

PLANT CONFIGURATION

Figure 2-1 is a simplified block diagram representing major subsystems of an oxygen-blown GCC plant. Figure 2-2 shows details of the oxygen-blown moving-bed arrangement used for this control study investigation. The fuel gas process consists of one moving-bed slagging gasifier and one fuel gas cleanup train. Steam required for the gasification process is partially supplied by steam generation in the gasifier water jacket with the balance being supplied by steam turbine extraction used for this control study investigation.

The moving-bed slagging gasifier, as in Figure 2-3, consists of a number of vessels stacked vertically. These are, from top to bottom: coal lock hopper, water jacket and gasifier chamber, slag quench vessel and slag lock hopper. A rotating distributor is located near the top of the gasifier chamber mantaining uniform coal flow and providing a level coal bed surface. The fuel bed is situated above a refractory hearth with facilities for running off liquid slag via a tap hole into a chamber below, where it is quenched in water. The quenched slag is discharged from the pressurized quench chamber by means of a slag lock hopper.

The coal flowing slowly down through the gasifier is a semi-solid bed of continuouusly changing composition. Preheated steam and oxygen are introduced near the bottom of the bed through a number of tuyeres and are further heated by contact with the high temperature slag leaving the reaction zone. The gasifier bed may be viewed conceptually as having five zones with varying physical and chemical conditions. These are, from the hearth upward: slag/ash, oxidation, reaction, devolatilization and drying zones. The relative thickness of each zone varies due to differences in the properties of coals and gasifier operation conditions. The vessel water jacket generates saturated steam for use in the process.

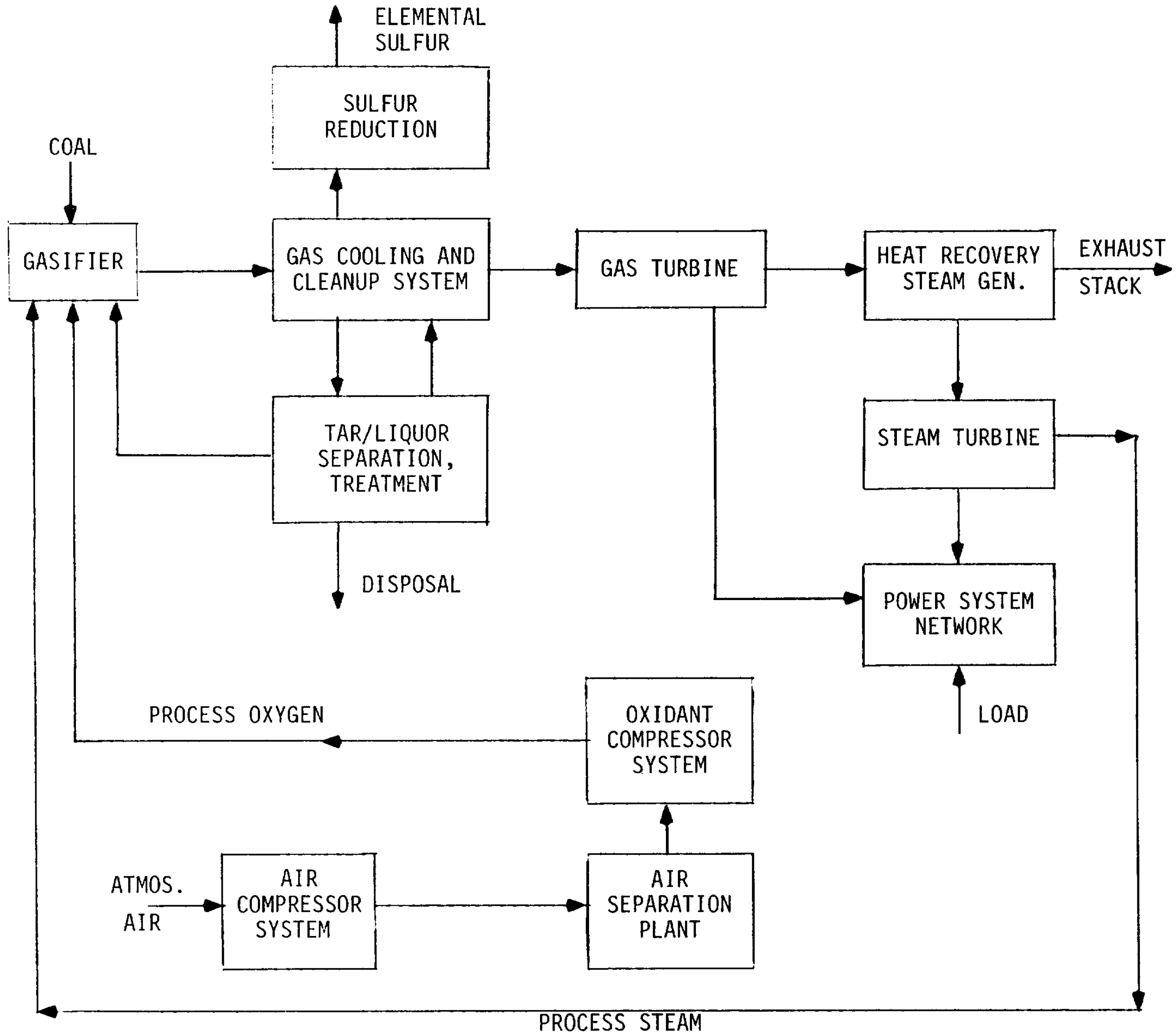


Figure 2-1. Simplified Block Diagram of an Oxygen-Blown GCC System

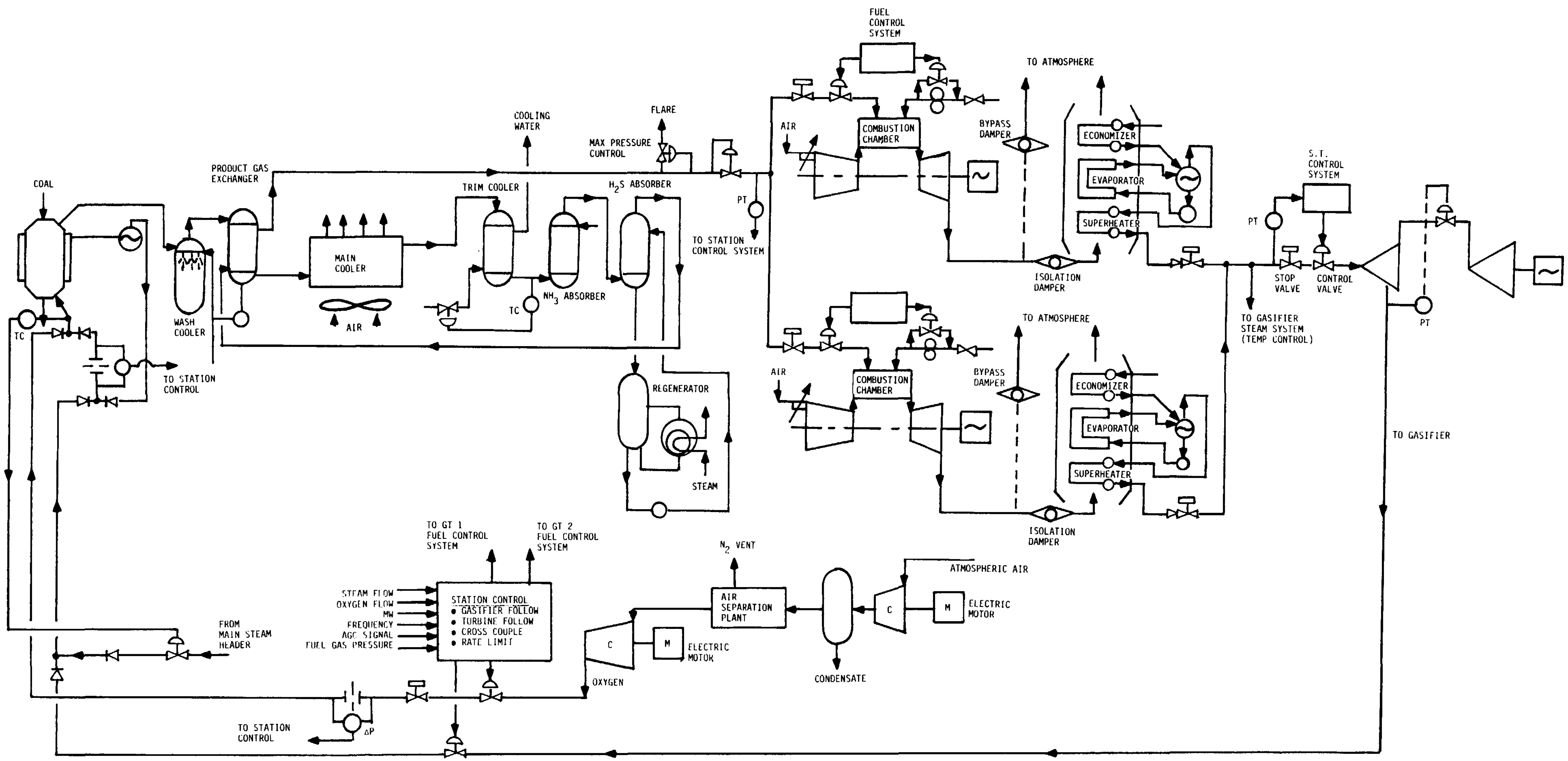


Figure 2-2. Oxygen-Blown, Moving-Bed Gasification Combined Cycle Plant Schematic

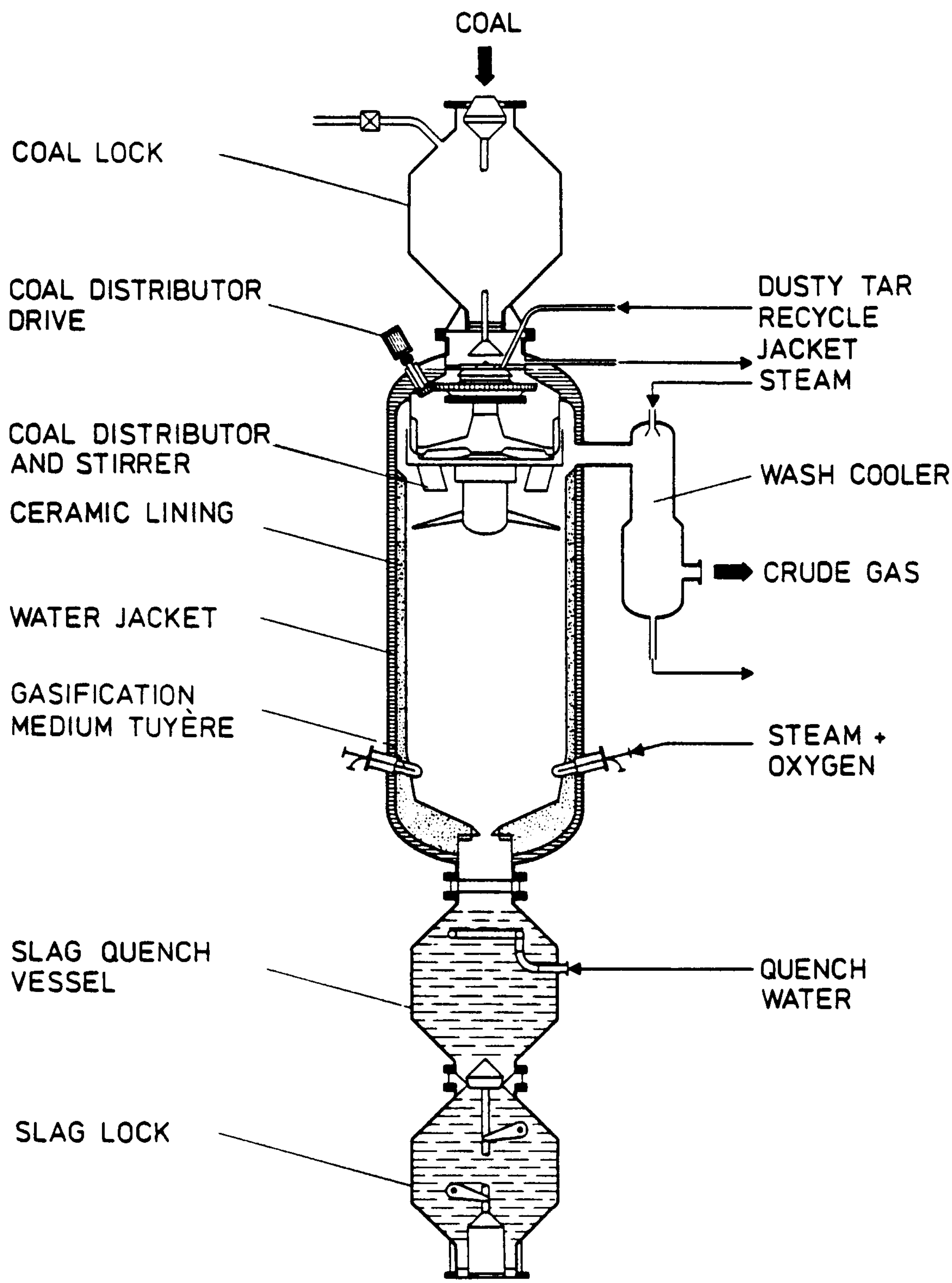


Figure 2-3. Moving-Bed, Slagging Gasifier Schematic*

*Figure taken from J. J. Albrecht and R. Reinert, (8) The fuel gas exits the gasifier at a high temperature. Entrained in the fuel gas stream are acid gas, ammonia, oils, tars and particulate matter. However, the gas turbine system and the environmental considerations dictate that relatively "clean" gas be furnished to the gas turbine combustion system. This is accomplished in the gas cleanup system by removing ammonia (NH_3), a major portion of the acid gas which is primarily hydrogen sulfide (H_2S), and particulate matter from the product gas stream. A low-temperature gas absorption process based on physical principles is used for selective hydrogen sulfide (H_2S) removal; therefore, it is necessary to lower the fuel gas temperature to 100-150 degrees F.

For this study, a single cleanup train was considered. Adjacent to the gasification reactor vessel is a wash cooler, into which flows raw gas from the top of the gasifier bed. A liquid stream spray directly quenches the fuel gas, lowering the gas temperature, and separates particulate matter from the gas stream to be removed by blowdown. Also a fraction of the heavy tar is condensed out and is removed by blowdown.

The saturated gas stream from the wash cooler is successively cooled in a number of indirect shell-and-tube type heat exchangers where condensate is removed. Ammonia compounds in the fuel gas are absorbed by water sprays in the NH_3 absorber, prior to the H_2S absorption tower where most of hydrogen sulfide (H_2S) and a fraction of carbon dioxide (CO_2) are removed. The solvent used for this acid gas removal flows into the regenerator where the absorbed H_2S and CO_2 are driven off, forming an acid gas stream to the sulfur reduction plant.

<u>The combined cycle plant</u> consists of two gas turbines, two heat recovery steam generators and one steam turbine. Each gas turbine nominally rated at 85 MW at approximately 2000 degrees F turbine inlet temperature and a pressure ratio of 11. Gas turbine exhaust gas is directed into heat recovery steam generators. Superheated steam is delivered to a common steam header. A low pressure economizer is used to heat water to generate low pressure steam for the deaerator. A <u>condensing steam turbine with a nominal rating of 70MW</u> is used. Steam control valves are used for steam pressure regulation and automatic extraction is used to supply process steam needs.

An electric motor-driven compressor provides compressed air to the air separation plant, which delivers low pressure oxygen. This oxygen is then pressurized by another motor-driven compressor for delivery to the gasification system. Discharge pressure and temperature are controlled. The electric drives associated with oxygen supply are assumed to be supplied with grid power for the purposes of this study.

Section 3

SIMULATION APPROACH

A control analysis has been conducted using computer simulation to study interactions associated with integrated GCC power plant operation and to evaluate alternative control strategies. The plant was configured with multiple components and commercial-size equipment to represent operating characteristics of large integrated plants including interaction of plant components under various loads and operation conditions.

The objective of this study is to model the oxygen-blown, slagging, moving-bed gasification-combined cycle and to use the model to predict plant response and evaluate different plant control strategies. This study is a continuation of a similar study of an air-blown, dry-ash moving bed gasification-combined cycle plant, whose approach and results are reported in EPRI Report AP1740(3). In effect, this provides a comparison of the operation and control of simulated GCC plants based on alternative moving-bed gasifiers of the Lurgi-type (dry-ash) or the BGC/Lurgi-type (slagging). To provide a meaningful, comparative analysis, the model developed under the earlier study has been used with appropriate substitution of modules in the gasification system area.

The following steps were taken in the execution of this study, and reflect of the modular substitution noted above:

- Select oxygen-blown slagging gasifier cycle configuration and sizing.
- Formulate analytical models for slagging gasifier and related components.
- Develop operational digital computer simulation of slagging gasifier and related components.
- Substitute oxygen-blown slagging gasifier and related component models into existing system simulation.
- Evaluate alternate plant control strategies.
- Compare results with previous air-blown dry-ash case.

A modular program structure was selected for flexibility in the GCC Plant Control study. Each component model was individually checked out with specific input variables before connecting it to upstream and downstream components to form a subsystem model. The fuel Process system, gas turbine and heat recovery steam generator are examples of such subsystems. The subsystem models are integrated to form an overall GCC plant model and a station control system is implemented to operate the plant automatically; i.e., under closed loop control. The ease with which the oxygen-blown slagging gasifier was substituted into the previously developed air-blown gasification system simulation, forcefully confirmed the validity of the modular program structure concept adopted for these system evaluations.

Dynamic simulation results in unique requirements for solution techniques which can avoid time-step numerical instability and yet realize economy in long duration computer runs. Time constants in various parts of such a large system are significantly different. Some process equations necessitate a smaller time interval for stable integration. The modular program structure allows the flexibiility to solve sub-routines with different integration time steps. For this study a one second time interval was selected as the primary integration step size while the gasifier subsystem used a variable time step and the pressure-mass flow model required a .01 second time step.

Section 4

MODEL DESCRIPTION

To permit evaluation of component interactions in the GCC plant system, and also to evaluate control concepts, a dynamic digital simulation was used. This simulation necessarily had to represent all the major components and their dynamic characteristics.

In an undertaking of this size, care must be exercised regarding the degree of detail included in individual component models. The scope of the study requires a balance between the size of digital simulation program and complexity of individual models for reasons of efficiency and economy. This can not be achieved by sacrificing representation of major dynamic effects; with too much simplification there lies a danger of "assuming the problem away" thus making the control study trivial. Therefore, it is important to identify those components in the cycle which have significant direct effects on fuel gas conditions and combined cycle power output. From the basic physical laws and chemical reaction relationships, a model can then be developed to simulate plant operation including both steady-state and dynamic behavior. The result is a set of non-linear differential and algebraic equations which must be solved simultaneously. The models must be in sufficient detail to predict equipment interactions in a multi-unit plant and should be usable to study off-design operating conditions.

Brief descriptions of the various component models developed for this study are provided below.

GASIFIER

The gasifier model was adapted from a model developed by K.J. Daniel of the General Electric Company Research and Development Center. This model was described in Reference (6), which is reprinted by permission of A.I.Ch.E. in Appendix A. A summary of the important features of this model and its adaptation for use in this study follows. The gasifier has modeled by a set of simultantous equations based on fundamental chemical, thermodynamic, and heat and mass principles. Reaction kinetics and other time-dependent phenomena are included. The equation set was adjusted at steady state conditions to achieve reasonable performance prediction as determined from published test data for the oxygen-blown, slagging gasifier.

The processes which significantly influenced the simulated transient response of a fixed bed gasifier were: (1) drying and devolatilization of the coal, and (2) thermal energy storage within the bed. For gasifiers operating with a low steam/oxidant ratio, these two processes are equally important in determining the transient change in the raw gas heating value. The thermal mass effects decayed over a longer time period than the drying and devolatlization effects. Furthermore, there is an interaction between changes in the bed temperature profile and the rate of formation of methane by chemical reaction. However, this interaction has a relatively small effect on raw gas composition since the model predicts that the majority of the methane is obtained by devolatilization. Devolatilization rate is, of course, directly affected by the temperature at the top of the fuel bed.

Changes in the raw gas heating value following a transient were found to be moderate. For a rather severe transient, involving a 20 percent step decrease of blast flow rate, the heating value of the raw gas increased by approximately 4.3 percent.

Major approximations in this model include the following:

- Extremely short duration transient effects are approximated as instantaneous
- Long duration effects are approximated as being constant
- Transients of an intermediate time scale are modeled in detail.

For this reason the model provides information only for transients on the time scale ranging from one-half to several minutes. Examples of transients which are assumed to occur immediately are: (1) changes in the combustion zone output conditions, and (2) propogation of gases through the reactor. Examples of transients which occur slowly and are assumed to be constant are: (1) the location of the combustion zone, and (2) the height of the coal bed. As a result of these assumptions, the thermal mass of the bed and drying time of the raw coal are characteristics that dominate the transient response of the gasifier.

Following the analysis of Yoon (1978) (9) and (10) both solid and gaseous phases are assumed to be at the same temperature throughout the gasifier. While this approximation is adequate for the prediction of trends in the output gases, it should be used with caution when attempting to predict conditions within the bed. Experimental data reported for Lurgi gasifiers show that the temperature difference between phases can be as much as 700 degrees K near the combustion zone (Rudolph, 1972) (11).

The gasifier is conceptually divided into three zones for the purposes of modeling the combustion zone, the reduction zone, and the devolatilization zone as depicted in Figure 4-1. These three zones are then divided into nineteen sections for modeling purposes.

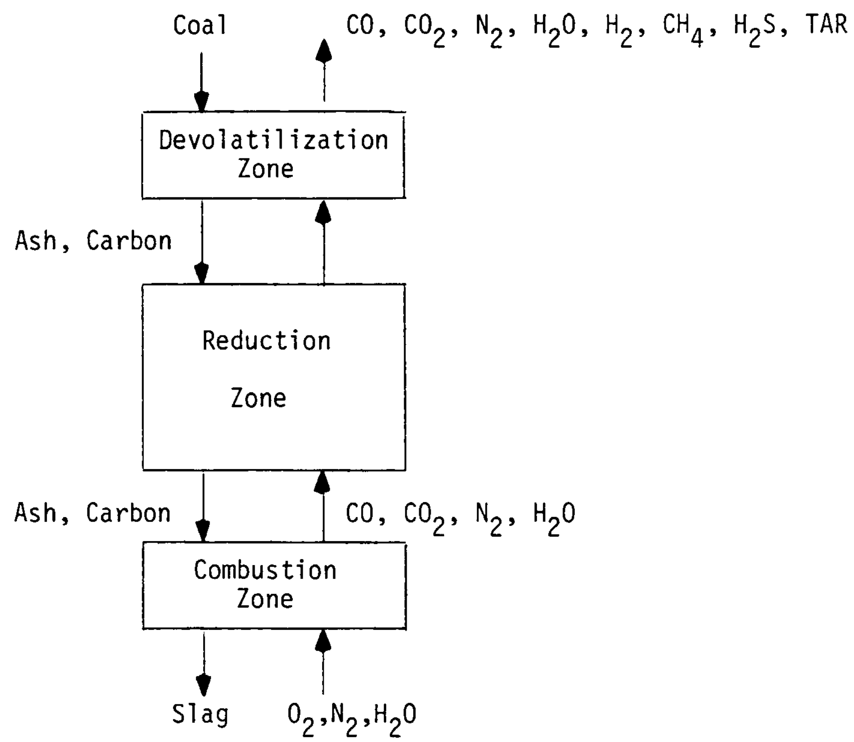
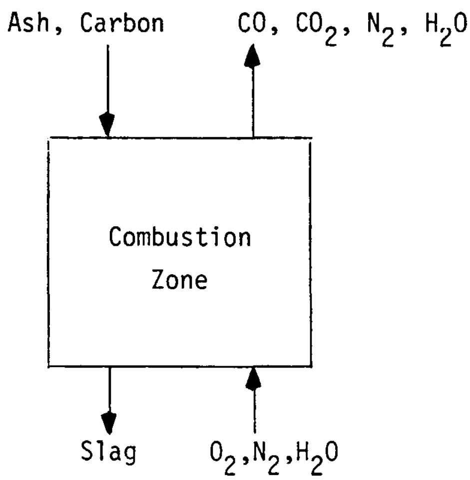


Figure 4-1. Reduction Zone Model

The reactions that occur in the combustion zone are very fast compared to other transient phenomena in the gasifier and dynamic effects are assumed to be instantaneous for this zone. The solution conditions for the combustion zone reactions are given in Figure 4-2.



STEADY STATE REACTIONS

$C + 1/2\ O_2 = CO$

$C + O_2 = CO_2$

$2H_2 + O_2 = 2H_2O$

$2CO + O_2 = 2CO_2$

$H_2O + C = CO + H_2$

$H_2O + CO = CO_2 + H_2$

SOLUTION CONDITIONS

$H_2O_{in} = H_2O_{out}$

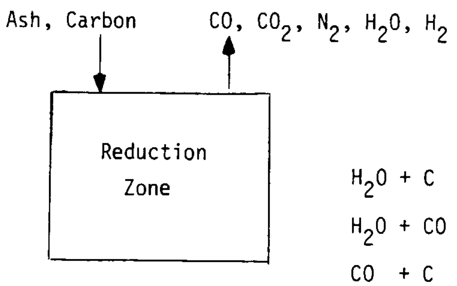
$\dot{\emptyset}\ O_{2\ in} = \dot{\emptyset}\ CO_2 + 1/2\ \dot{\emptyset}\ CO$

$\dot{H}_{blast} + \dot{Q}_{React} = \dot{H}_{gas}$

$$\frac{(P_{CO})^2}{P_{CO_2}} = \beta(\dot{m}_B) K_{eg} T_c$$

Figure 4-2. Combustion Zone Model

The reduction zone solution conditions are given in Figure 4-3. The combustion zone represents the 1st of the 19 sections and the reduction zone represents the next 17 sections. The equation calculating net generation or consumption of chemical species is solved by integrating spatially for the 17 bed segments. The chemical kinetic equations determine the rate of production or disappearance of the species present, and water-gas shift equilibrium conditions are imposed on each segment to determine a proper balance among major gas species (CO, CO_2, H_2, H_2O). The energy equation then is solved to give the time rate of change of temperature throughout the bed.



$$H_2O + C = CO + H_2$$
$$H_2O + CO = CO_2 + H_2$$
$$CO + C = 2\ CO$$
$$C + 2H_2 = CH_4$$

$$\frac{\partial C_i}{\partial t} = \frac{\partial C_i U}{\partial x} + \sum_j n_j a_{ij} R_j(T)$$

$$\rho_c A C_{p,c} \frac{\partial T}{\partial t} = -\sum \Delta Hf_i^o \frac{\partial \dot{m}_i}{\partial x} - \sum \dot{m}_i C_{pi} \frac{\partial T}{\partial x}$$

$$- h\pi d(T = T_{wall})$$

Figure 4-3. Reduction Zone Model

The devolatilization zone is the last section and is modeled by assuming that volatile products are released from the coal at a rate proportional to the coal feed, but with a suitable time lag.

The general moving-bed gasifier model described in the appendix, can represent any moving-bed gasifier from dry-ash or slagging, and air- or oxygen-blown. It is required only to change the gasifier geometry, the jacket heat loss, and the products of devolatilization based on the specific coal used. Proper selection of the reaction kinetic coefficients brought the gasifier model results to within an acceptable tolerance of the desired values. Table 4-1 shows the model final results with tuned kinetic coefficients; Figure 4-4 shows a typical bed temperature distribution predicted by the model.

The effects of gasifier bed phenomena of bridging, hang-slip and channeling were modeled by varying the gasifier steam and oxygen feed in a fluctuating pattern. This results in fluctuations in both coal consumption and exit gas flow of varying composition, for seemingly constant gasifier input. The net effect is similar to the type of fluctuations in gasifier output observed in actual process development unit testing performed under EPRI Project RP1267-1 at the Westfield Development Center (Scotland) of British Gas Corporation (see EPRI Report AP-1922, Ref. (7) and the Lurgi paper by Albrecht and Reimer, Ref. (8)).

Table 4-1

COMPARISON OF MODEL RESULTS WITH BRITISH GAS SLAGGER DATA

	BGC DATA (Ref. 1)	REVISED MODEL
CO, Mole %	54.2	53.2
CO_2 Mole %	1.8	3.1
H_2, Mole %	28.8	31.5
CH_4 Mole %	8.3	5.8
N_2, Mole %	0.84	0.9
H_2O Mole %	4.64	4.05
H_2S Mole %	1.3	1.4
Coal Flow, lb/hr	138,888	137,445
Exit Gas Temp., ^oF	820	853
Combustion Zone Temp., ^oF	3500	4000
Effluent Gas HHV, Btu/SCF	360	341

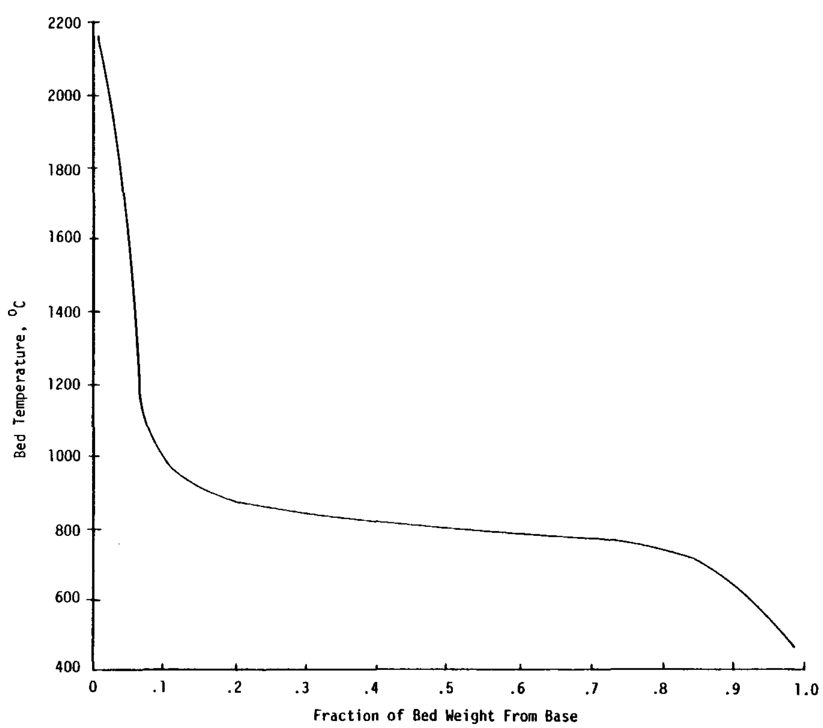


Figure 4-4. Predicted Bed Temperature Distribution

Cleanup System

A mathematical representation has been developed for those cleanup system components which directly impact fuel gas properties, flow, pressure and temperature, and ultimately have a significant effect on the power plant performance. These components include:

- Wash Cooler
- Waste Heat/Product Gas Exchanger
- Gas Cooler
- Boiler Feedwater Heater
- Trim Cooler
- H_2S Absorber

The overall cleanup system model is structured to accept gas flow, temperature, and composition from the gasifier. The total flow is made up of dry gas components and water vapor. In several stages of cooling, moisture is added (in the wash cooler) or condensed (in the gas cooler). Therefore, the moisture content of the product gas varies significantly through the cleanup system. The dry components (CO, CO_2, H_2S, H_2, N_2, CH_4) remain unchanged through several stages of cooling and therefore the dry gas flow and volume fractions may be treated separately from the water vapor flow and volume fraction.

In the H_2S absorber, differing fractions of H_2S and CO_2 are absorbed and thus at the tower exit the dry gas composition is changed and recomputed in the model.

Gas properties, such as specific heat and molecular weight and the universal gas constant, are calculated in the model as a function of the varying gas composition, temperature, and pressure in each component of the cleanup system.

The large volume of the cleanup system (approximately 7,000 cu ft) leads to a substantial gas residence time in the cleanup system. Therefore, gas composition and temperature experience a significant transport lag as the gas traverses the cleanup system. Components of the dry gas undergo different dynamic changes in the gasifier, and to represent the accompanying variations in gas properties and the effect on the H_2S tower absorption performance, it is necessary to apply the transport lag to each component of the dry gas. A variable transport lag effect has been modeled to account for the flow change effects on the residence time.

The cleanup system model generates output to the gas turbine in terms of gas flow, temperature, pressure, and composition (as mass fraction).

Both the sour gas stream composition out of the H_2S regenerator and the H_2S tower pressure drop are calculated on a dynamic basis.

For modeling purposes, the fuel gas cleanup system components may be divided into two categories: (1) saturated vapor-gas mixture components where the moisture content is high, and (2) dry gas components where the moisture is superheated or the moisture content is very low. The pressure and temperature dynamics for the two types of components are different because flashing and condensation of the moisture occur with a corresponding latent heat release.

Appropriate assumptions have been made to develop a simple, lumped parameter model of process dynamics. For those components, such as the gas cooler where the temperature and/or the moisture content vary significantly from the inlet to the outlet, the equipment is modeled as several sections connected in series; each section is representated by the lumped parameter model. In this way, a distributed-capacity model has been aproximated which more closely represents the actual process flow and temperature dynamics.

The individual component models were derived from the basic mass balance, energy balance and volume balance equations. These equilibrium equations were written, in turn, for the dry gas, vapor, liquid and the vapor-liquid interface. The equations were combined and solved simultaneously to yield the temperature-rate and pressure-rate equations.

The cleanup system components and inter-connected piping comprise a significant volume or capacitance effect. The total volume was divided into several control volumes to preserve the transient flow-pressure characteristics. Equations were written for the flow rates between the control volumes and the pressure in each node was determined based on the flow rates, the changes in the gas-liquid storage mass and the energy transfer rates. All components within a given control volume undergo the same pressure-rate change.

The cleanup system models can operate at low loads. Under very low load conditions, however, some reduction of the calculation time inteval is necessary since the pressure-flow system time constants are very small at the low flow rates.

Combined Cycle

The combined cycle power plant includes gas turbines, heat-recovery-steam generators (HRSG's) and steam turbine. The models were derived from fundamental thermo-dynamic relations with energy and mass balance equations. Representative hardware data was used for simulation models.

Figure 4-5 shows the flow of data between components. Gas composition, flow, temperature, and pressure are input from the fuel gas cleanup system to the gas turbine, while gasifier jacket steam flow and temperature are input to the steam-oxygen mixer. Outputs from the combined cycle plant include power to the utility system, and extraction steam flow and temperature to the steam-oxygen mixer. Flows, temperatures, and pressure are exchanged between component models. For example, the temperature of the steam-oxygen mixture (or blast) to the gasifier is calculated as shown in Figure 4-6, and regulated by extraction steam flow.

The models for compressor, combustor and turbine are derived from basic governing equations. Air flow is calculated as a function of speed and pressure ratio with appropriate correction factors for ambient temperature and pressure. Representative values of compressor and turbine efficiencies are used to calculate actual temperatures and hence the power output. A turbine control system, including protective control, logic is also simulated. This includes speed-load control, temperature control, and inlet guide vane control (see Figures 4-7 and 4-8). A pressure control system is also implemented when the system is run in a gas turbine follow control mode. The model is capable of simultaneous operation on dual fuel. The combustion system model is expanded for operation both on light fuel oil and on a low Btu, variable composition gas.

Heat Recovery Steam Generator

The superheater, evaporator and economizer are the three heat transfer sections of heat recovery steam generator. The exhaust gas from gas turbine flows across these three sections in a cross-flow arrangement. Due to the relatively low temperature of the exhaust gas, convection is the primary mode of heat transfer involved. A lumped parameter heat transfer model is used. The superheater is divided into two sections connected in series for a better representation of temperature transients. Mass, energy and volume equations for the evaporator and steam drum are written for the liquid and vapor phases. These equations when combined with heat transfer and pressure flow equations yield a steam pressure

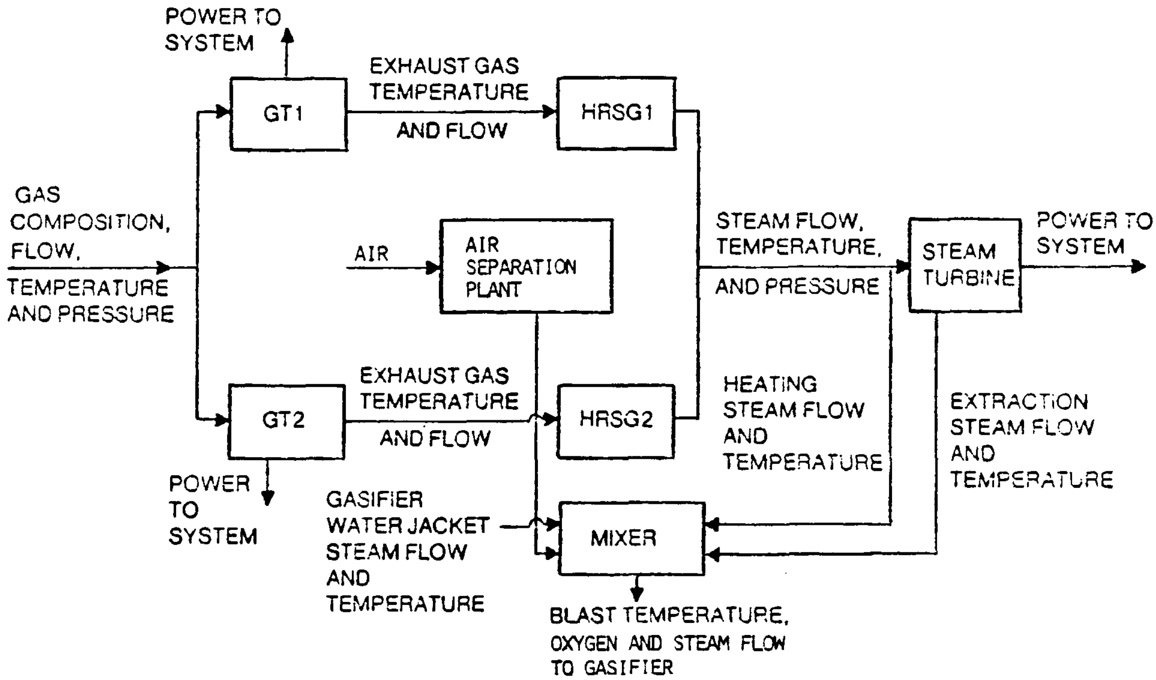


Figure 4-5. Combined Cycle Model Schematic

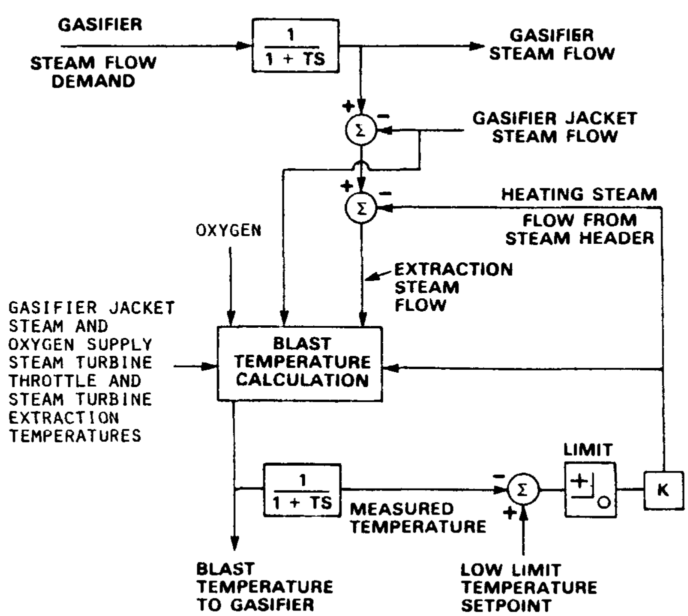


Figure 4-6. Gasifier Blast Temperature Control Schematic

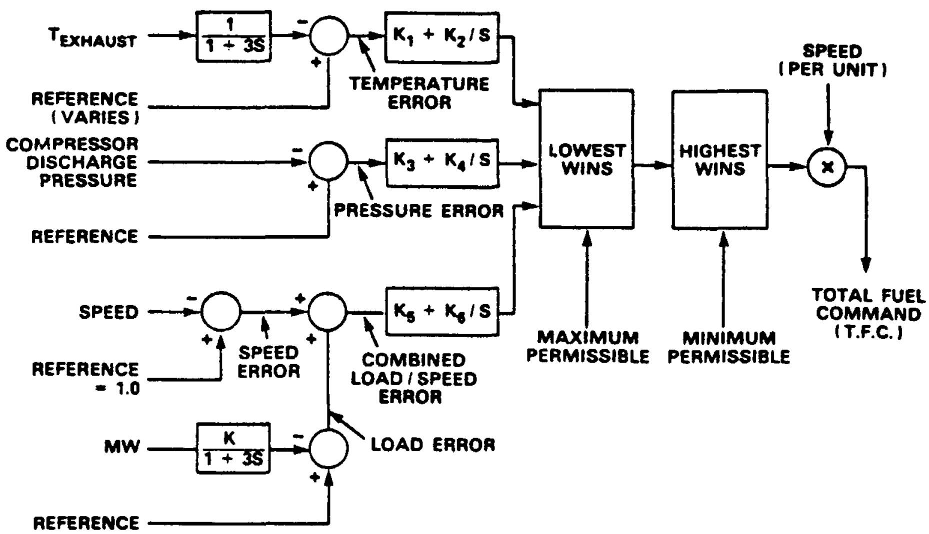


Figure 4-7. Gas Turbine Fuel Control Schematic

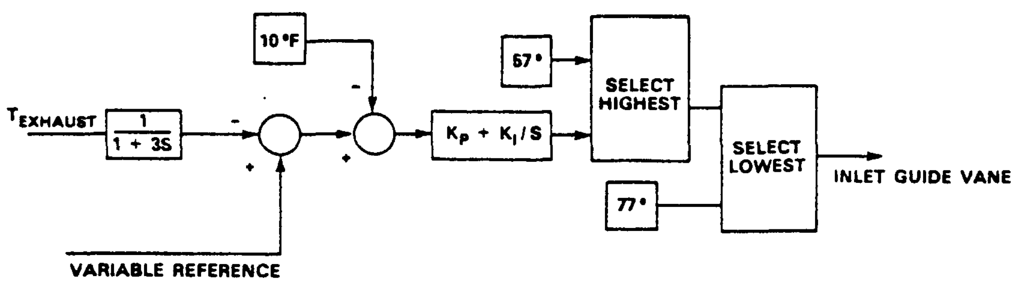


Figure 4-8. Gas Turbine Inlet Guide Vane Control Schematic

rate equation. A three-element drum level control can tightly regulate the drum water level by matching feedwater and steam flow. In the model, it is assumed that the feedwater flow lags the steam flow by five seconds. Feedwater from the economizer enters the steam drum and mixes with saturated water from the evaporator. Feedwater entering the evaporator is nearly saturated. For model simplicity, the subcooling effects (which are small) are neglected in the evaporator dynamics.

Steam Turbine

Steam from the two heat recovery steam generators flows into a common header and is delivered to the steam turbine. The steam turbine is a single, automatic extraction condensing unit with high pressure and low pressure sections. Steam pressure control is provided for process steam extraction. The steam turbine control valves are on pressure control and respond to steam flow changes to regulate throttle pressure at a five percent droop control. The speed control is active as a back-up only for overspeed protection. For reasons of efficiency, it is desirable to operate with a large control valve opening. Three levels of pressure setpoint are established as a function of valve position between 70 percent and 95 percent. The pressure setpoint is ramped from one level to another at 40 psi/minute. Enthalpy drop across each section of the steam turbine is calculated using a representative efficiency value for known steam conditions. The steam turbine output and extraction condition are then known.

Oxygen Supply System

For the purposes of this particular evaluation study, an oxygen plant was not included in the model. It was assumed that adequate oxygen would be available at all times. Furthermore, it was assumed that electric motor drives were used for the air and oxygen compressors with electric power supplied the grid, thus resulting in no direct impact on the combined cycle steam supply or power output. In reality, such a system would of course reduce available plant electrical output. This effect was not considered important at this level of simulation.

Section 5

RESULTS

As was done in the air-blown dry-ash GCC control study, open loop runs were made (i.e., without automatic controls) to investigate individual component integrity and response. The component models were then integrated and automatic (closed-loop) plant control was implemented to evaluate plant response to load demand.

Plant response was evaluated for both the gas turbine-lead control mode and the gasifier-lead control mode. One investigation made here that was not done for the air-blown dry-ash GCC control study was the simulation analysis of the effects of irregular fluctuations of gasifier output (due to bridging, channelling or hang-slip phenomenon in the gasifier bed) on the overall system.

Results are presented for comparison to the air-blown dry-ash GCC system. They also demonstrate the degree of interactions among components of the integrated plant in alternate control modes (i.e., gasifier-lead and gas turbine-lead).

Block diagrams depicting the overall control schemes used for the gas turbine-lead and gasifier-lead control schemes are shown in Figure 5.1 and 5.2 respectively.

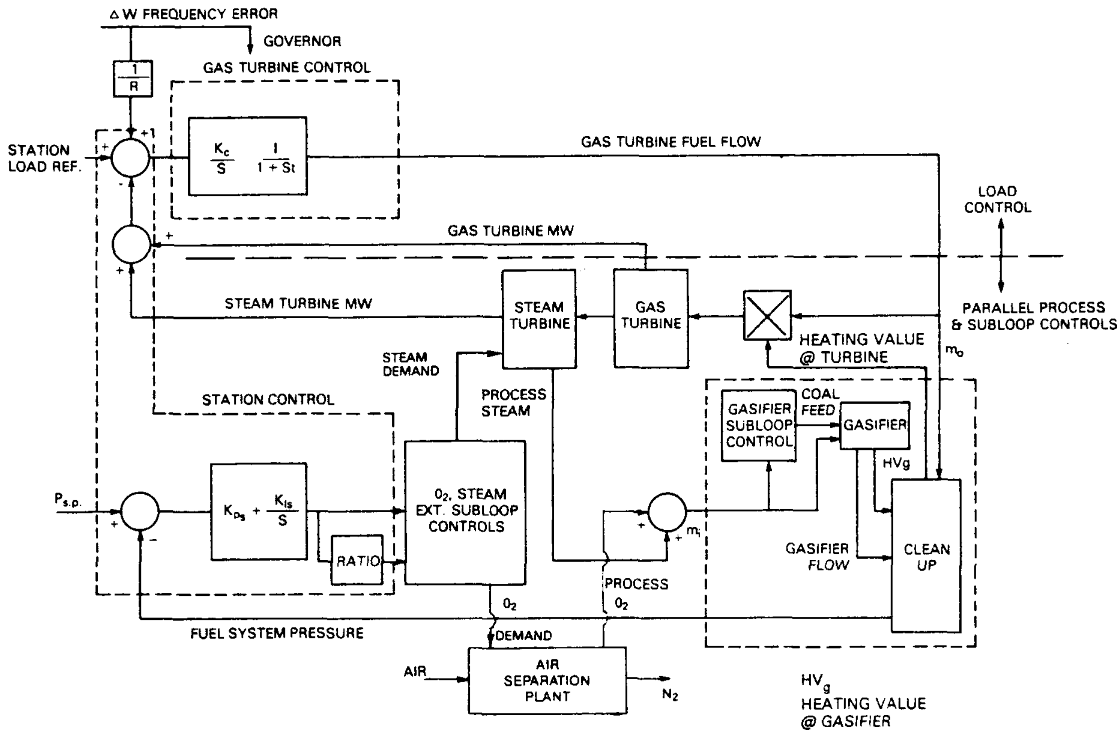


Figure 5-1. Plant Control Schematic for the Gas Turbine-Lead Mode

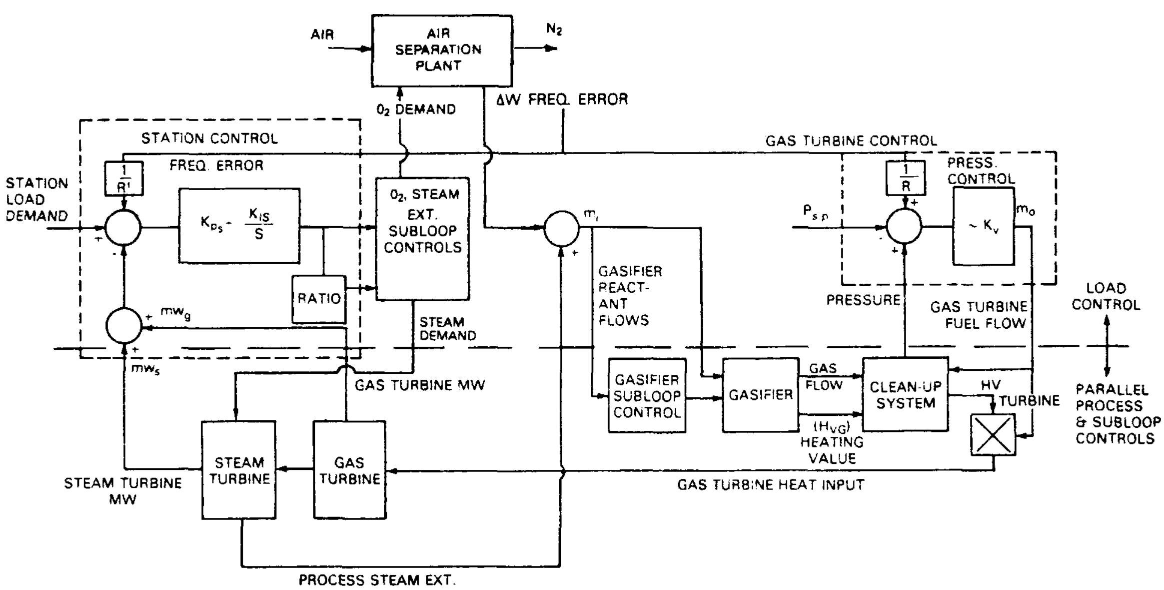


Figure 5-2. Plant Control Schematic for the Gasifier-Lead Mode

Case 1: GCC Plant Response, Turbine-Lead Control Mode

Commercial combined cycle plants based on conventional fuels are capable of changing load (up or down) at 8% per minute. However, cleanup system considerations such as potential for tar and fines carryover from the gasifier and solvent carryover from the H_2S absorption column, require that GCC plant load changes be accomplished at a reduced rate. A 4% per minute rate was found to be satisfactory for the air-blown, dry-ash moving bed GCC study and was used here.

In the turbine-lead control mode, the plant controls manipulate the gas turbine fuel valve in response to station load demand by adjusting the gas turbine load set point, while gasifier steam and oxygen feed are controlled by a proportional-plus-integral (PI) controller to maintain a set value of fuel pressure at the gas turbine fuel valve (see Figure 5.1).

The height of coal in the gasifier bed is maintained at a fixed level by action of the distributor plate. The distributor plate is fed by the lock hopper which is periodically recharged. Ash is removed from the bottom of the gasifier as slag.

The plant load set point was ramped down 20% at a 4% per minute rate. Figure 5.3 shows that the plant power output follows the load set point very tightly throughout the transient with a slight undershoot at the end of the set point ramp. The plant unloading rate closely follows the demand, i.e., at a rate of 4% per minute.

As shown in Figure 5.3, the gas turbine initially unloads at a rate greater than 4% per minute while the steam turbine unloads at a lesser rate. This lag is due to a lower extraction steam demand and the energy storage capacity of the HRSG-steam system. The rates are reversed near the end of the ramp change in load (at about 320 seconds), where the steam turbine unloading rate exceeds 4% per minute while the gas turbine unloading rate is less than 4% per minute. Consequently the gas turbines appear more than fast enough to compensate for either over- or under-action of the slower responding steam turbine. The gas turbine does undergo a small undershoot at the end of the load ramp.

Figure 5.3 also shows that the fuel lower heating value (LHV) at gasifier exit rises only abut 0.8% above the nominal value and even less at the gas turbine fuel valve due to the effective gas residence time in the cleanup system. This presents no problem to the gas turbine. Gas turbine fuel flow changes about 17% for the 20% decrease in load. This reduction in fuel flow is similar to that experienced by a conventional combined cycle plant.

The gasifier and cleanup system both experience moderate, and well controlled changes uniformly over the transient period as shown in Figure 5.4. The use of the tight control (PI) on pressure, limits the pressure overshoot to 2.9 psia (~1%) while resulting gasifier feed changes are on the order of 3-1/2% per minute.

The effects of the transient on the gas turbine are shown in Figure 5.5, and the effects on the steam turbine are shown in Figure 5.6.

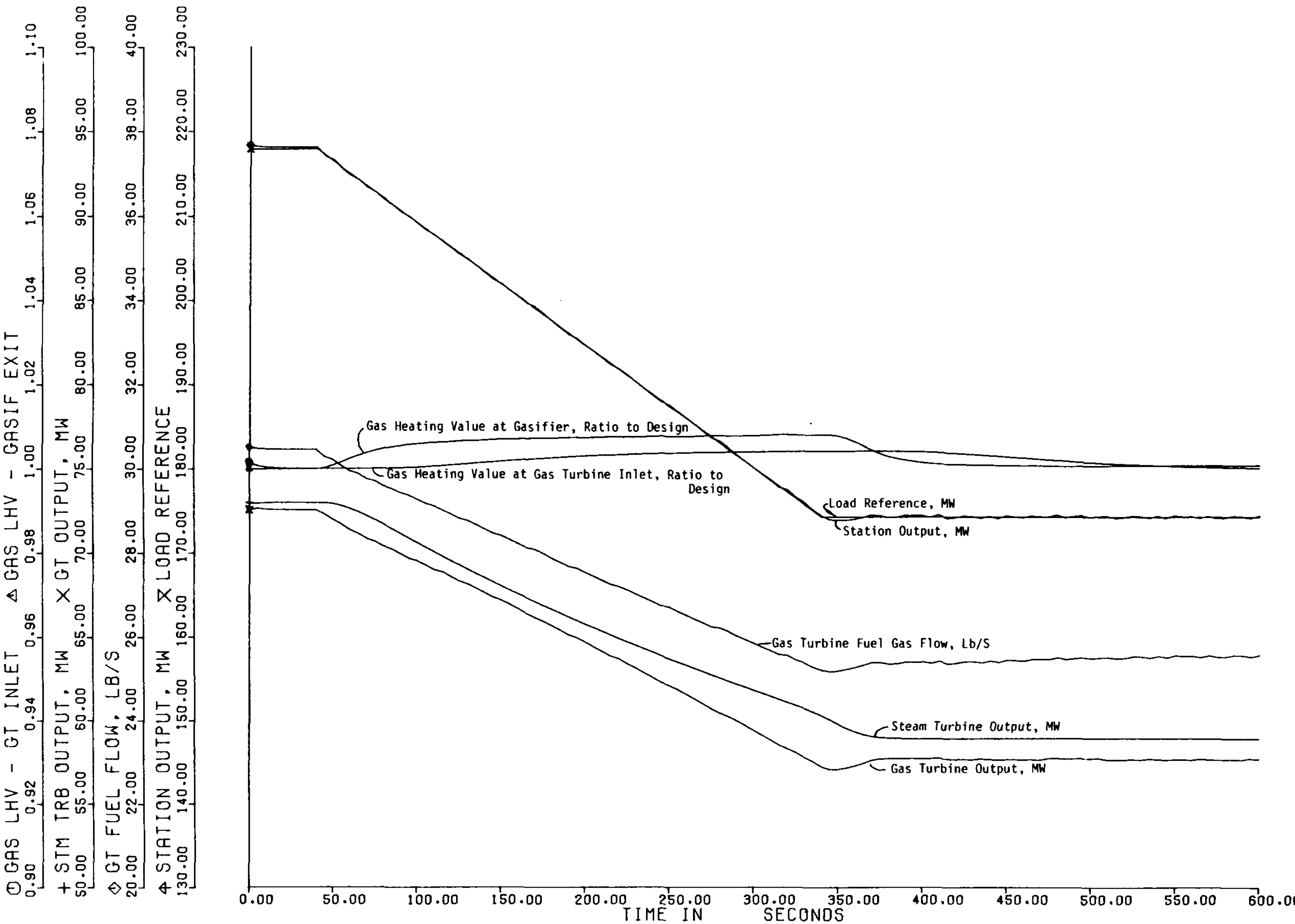


Figure 5-3. Case 1-GCC Plant Megawatt Output and Heating Value Transients in Turbine-Lead Control Mode for a 20% Decrease in Load Demand at 4%/Minute

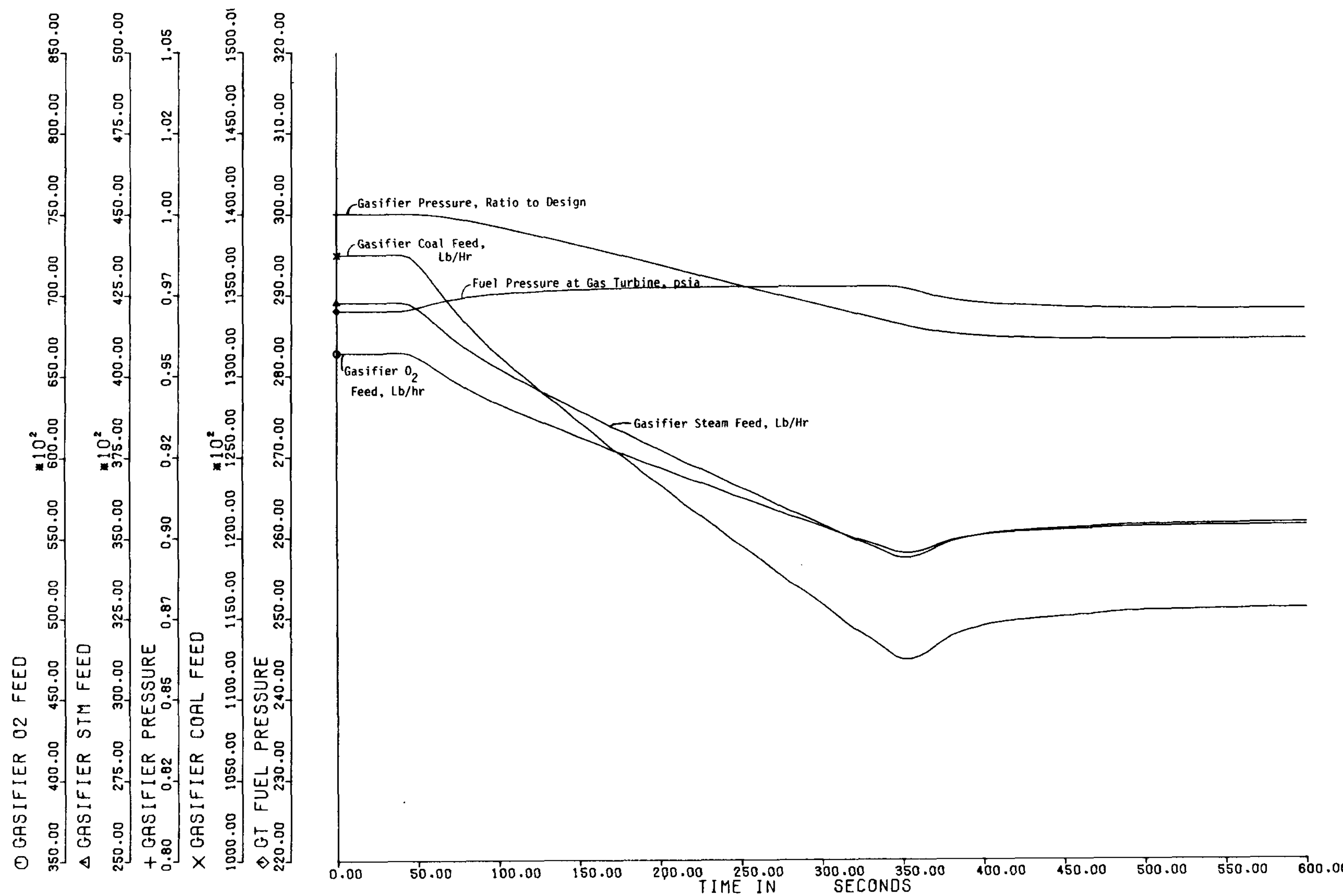


Figure 5-4. Case 1-GCC Plant Gasifier Transient in Turbine-Lead Control Mode for a 20% Decrease in Load Demand at 4%/Minute

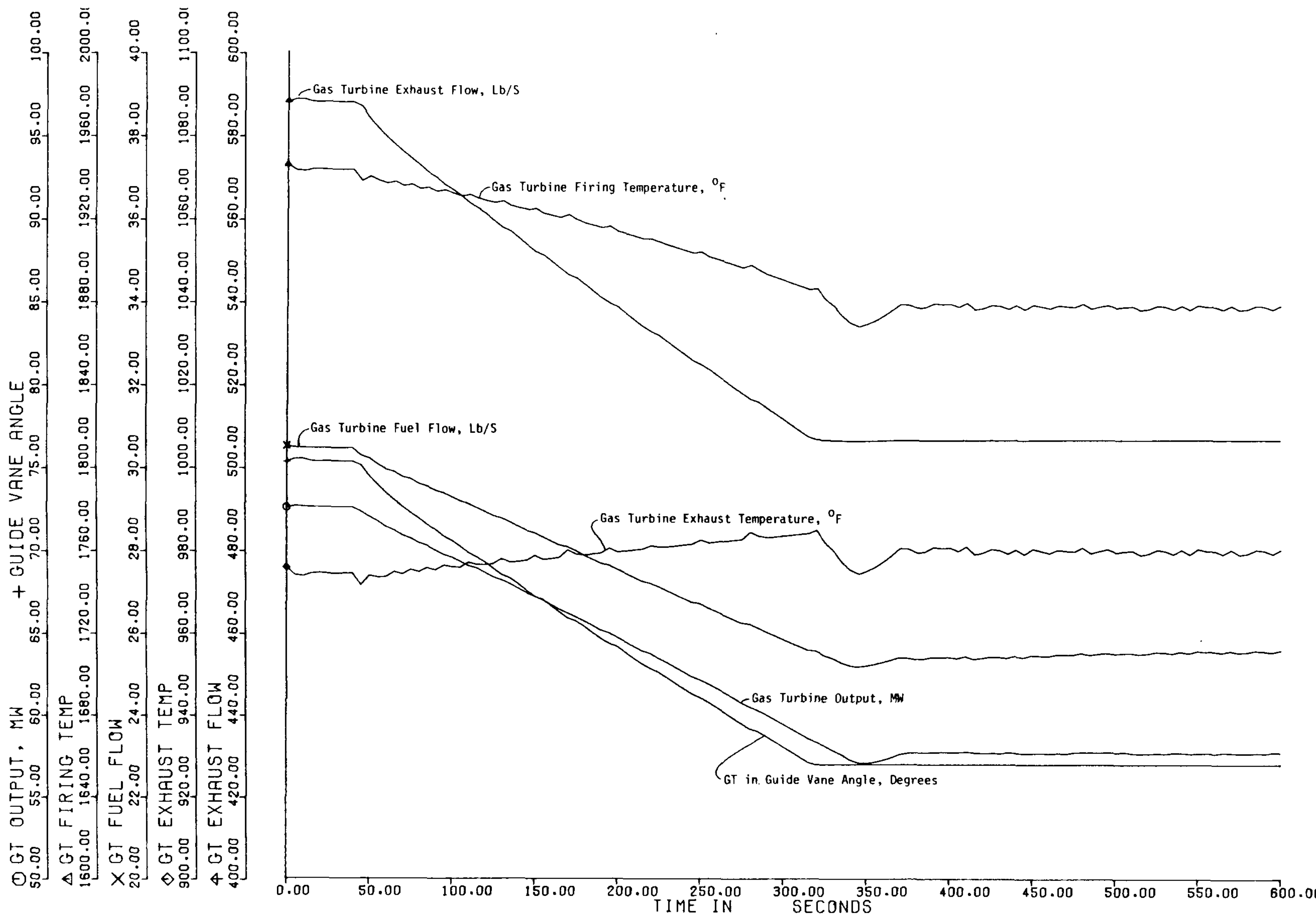


Figure 5-5. Case 1-GCC Plant Gas Turbine Transient in Turbine-Lead Control Mode for a 20% Decrease in Load Demand at 4%/Minute

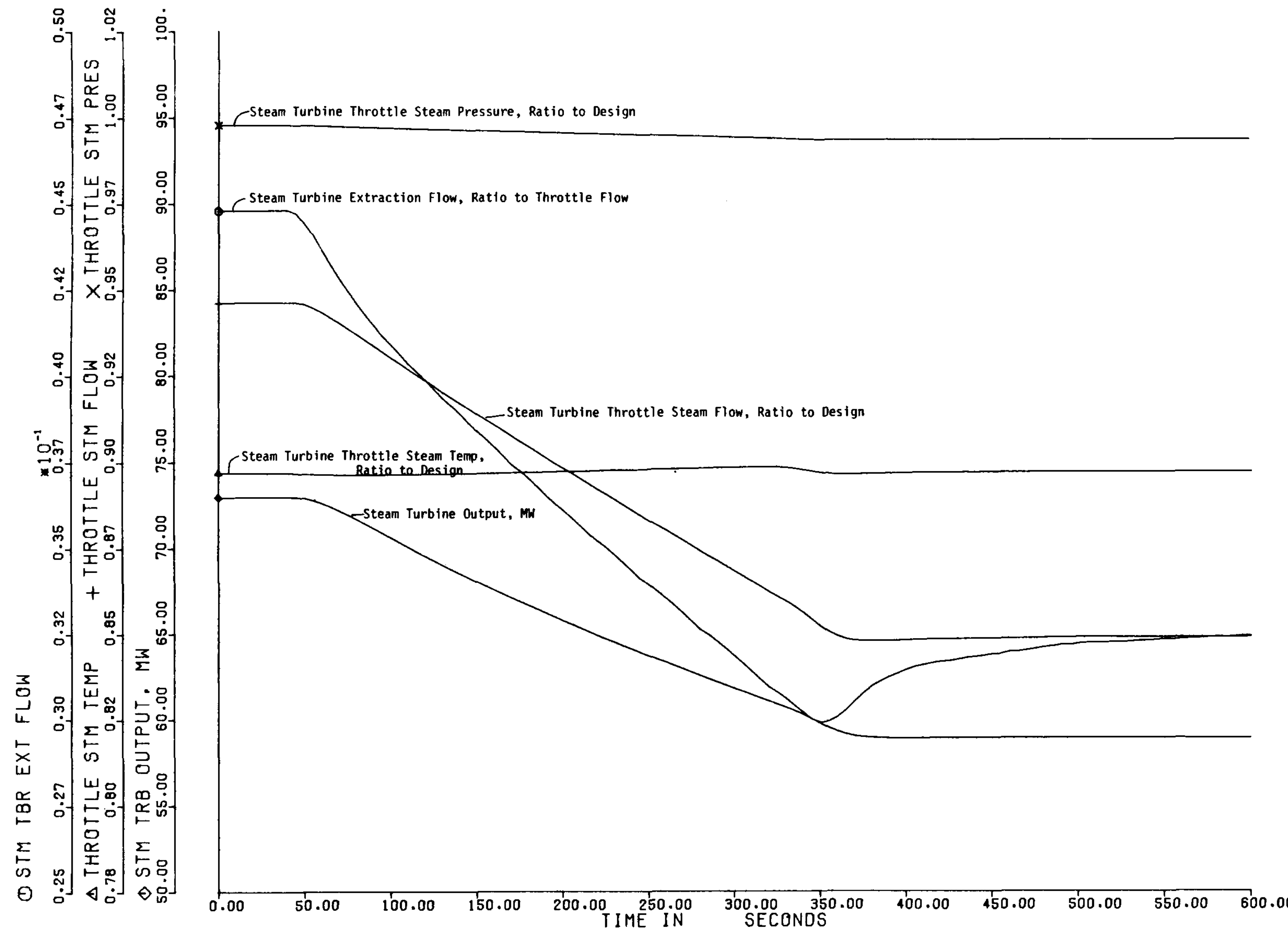


Figure 5-6. Case 1-GCC Plant Steam Turbine Transient in Turbine-Lead Control Mode for a 20% Decrease in Load Demand at 4%/Minute

Case 2: GCC Plant Response, Gasifier-Lead Conrol Mode

In this control mode, the plant load set point directly modulates the gasifier steam and oxygen feed with a proportional plus integral controller. Increases or decreases of gasifier feed streams are reflected in increased or decreased pressure at the turbine control valve. The increase or decrease in pressure, of course, is not reflected immediately at the turbine fuel valve due to the long gas residence time created by the capacitance effect of the large volume of the cleanup system. The fuel system pressure is regulated by the gas turbine fuel control valves with a proportional control having 4% droop.

The plant load set point was moved at a 4% per minute rate for a 20 percent decrease in load.

As shown by Figure 5.7, the response rate of 4% per minute is achieved after an initial delay of about 10-15 seconds due to the inherent lag of the fuel system. Also, the plant is slow in reaching the final steady state value. There is no undershoot in this case. Since the gas turbine is not the leading control element, it only responds following fuel system changes which are then reflected in the GCC plant output.

The transients in the gasifier and cleanup system, as shown in Figure 5.8 are similar to those of Case 1. There is no pressure overshoot, of course, due to the regulating action of the gas turbine fuel valve.

The gas turbine response shown in Figure 5.9 and the steam turbine response shown in Figure 5.10 are also similar to those of Case 1.

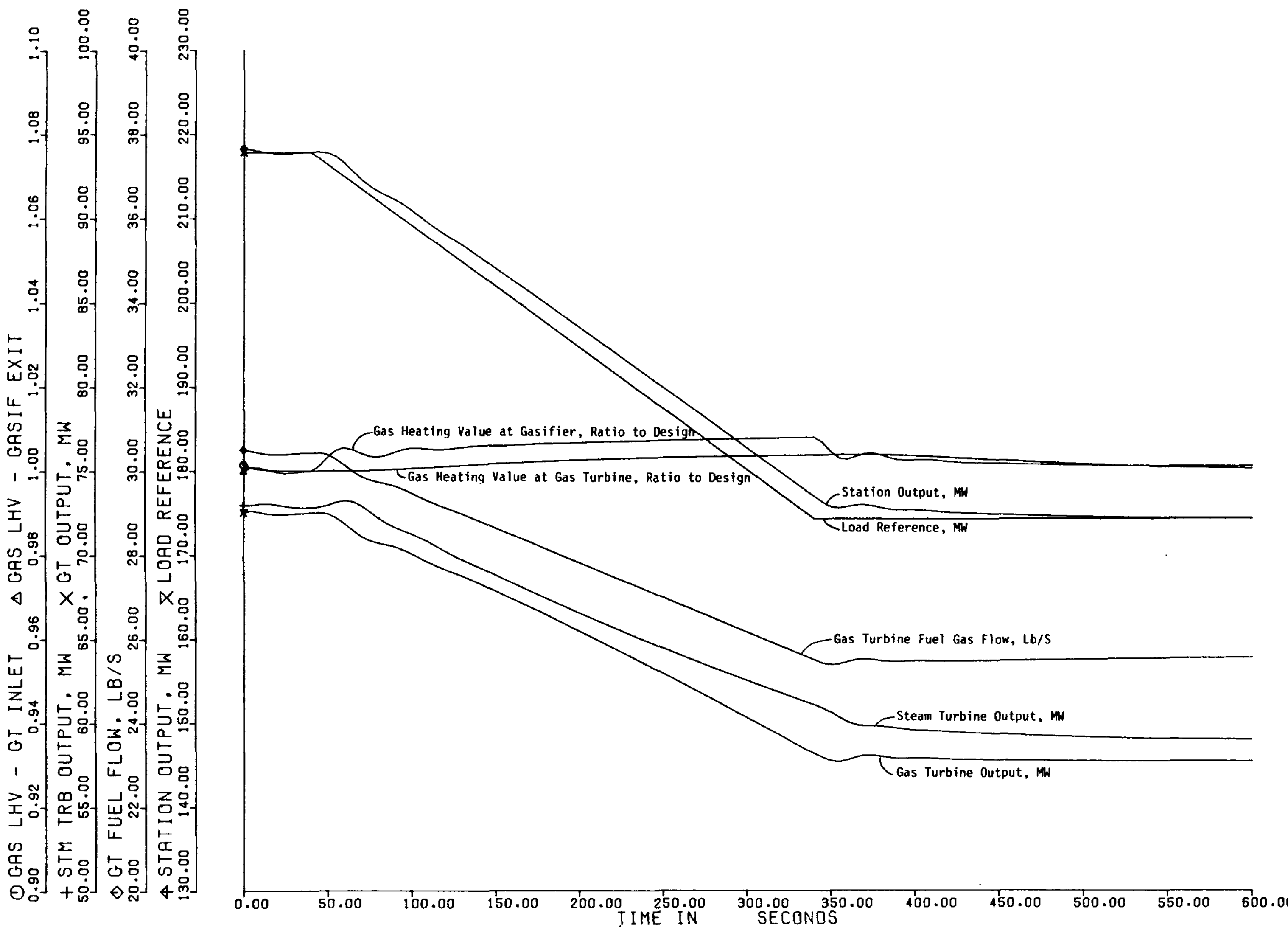


Figure 5-7. Case 2-GCC Plant Megawatt Output and Heating Value Transients in Gasifier-Lead Control Mode for a 20% Decrease in Load Demand at 4%/Minute

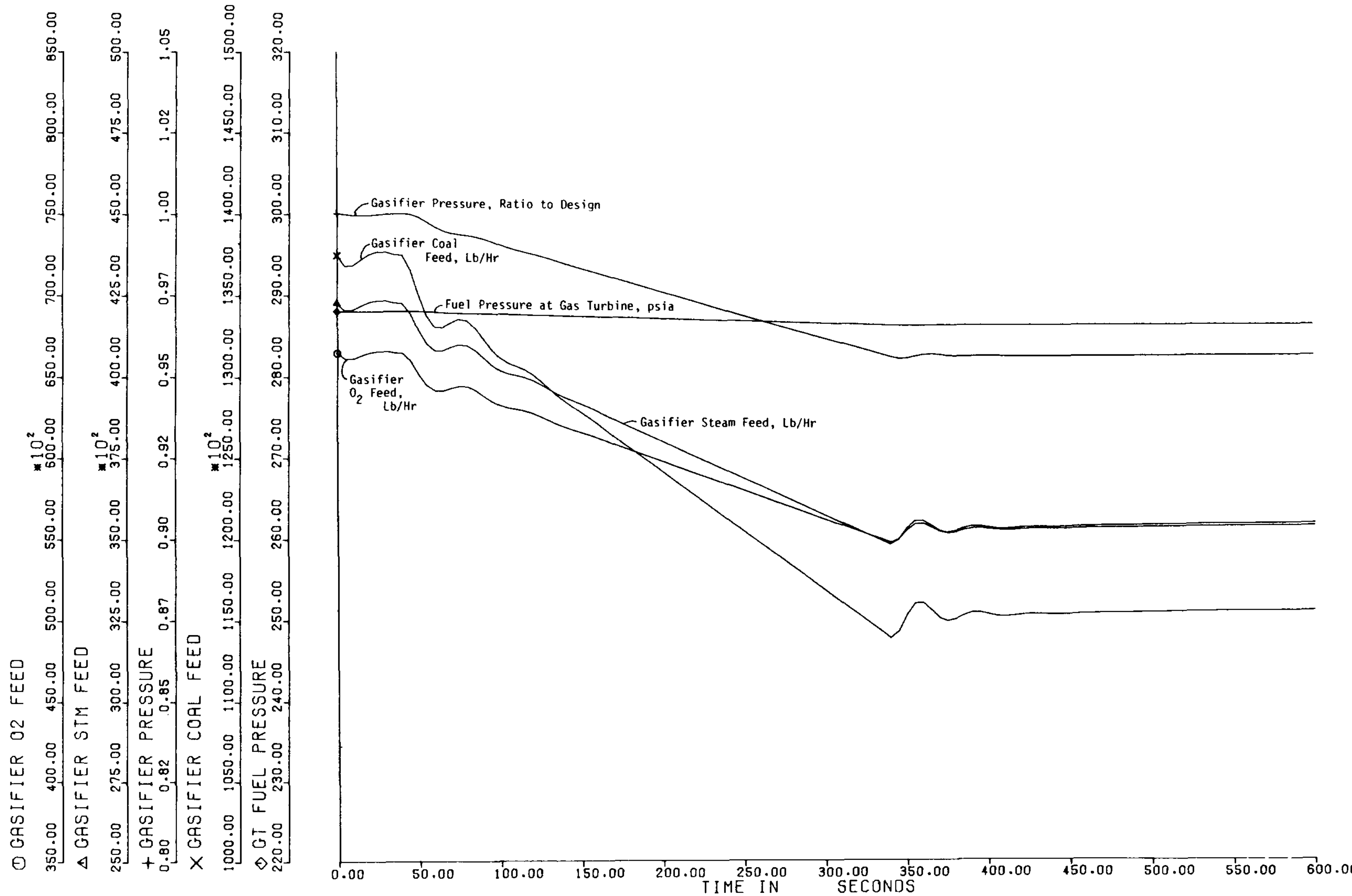


Figure 5-8. Case 2-GCC Plant Gasifier Transient in Gasifier-Lead Control Mode for a 20% Decrease in Load Demand at 4%/Minute

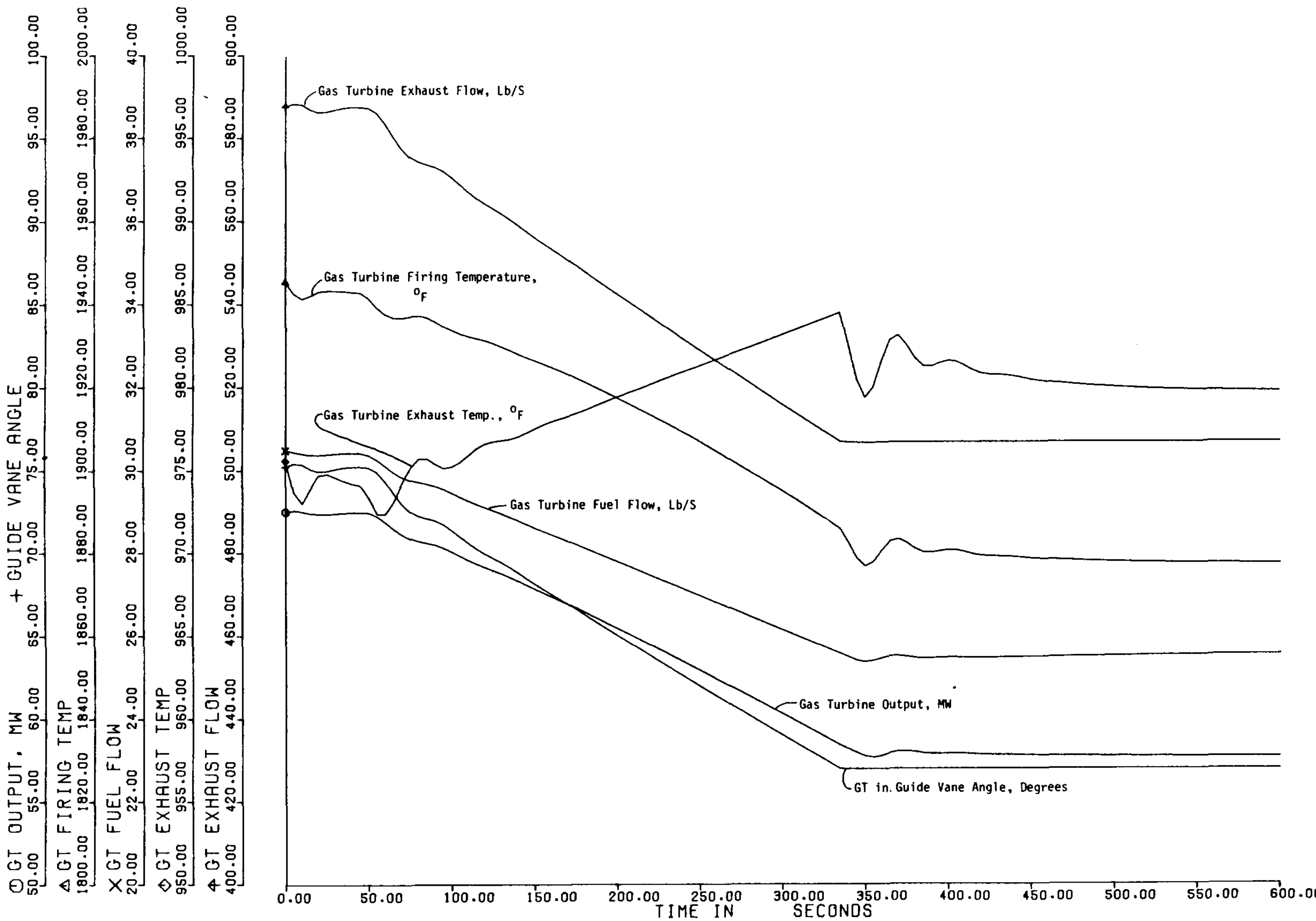


Figure 5-9. Case 2-GCC Plant Gas Turbine Transient in Gasifier-Lead Control Mode for a 20% Decrease in Load Demand at 4%/Minute

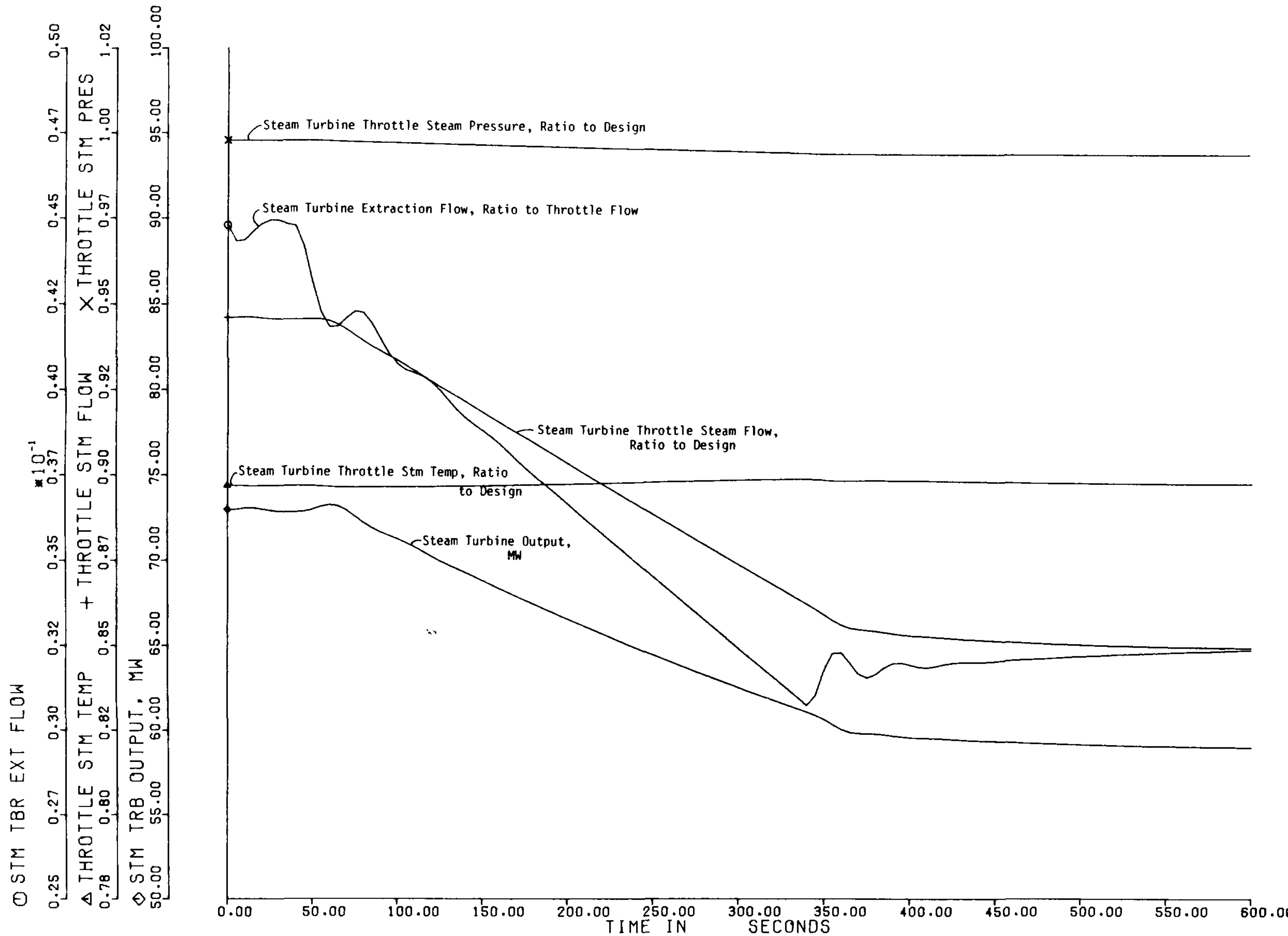


Figure 5-10. Case 2-GCC Plant Steam Turbine Transient in Gasifier-Lead Control Mode for a 20% Decrease in Load Demand at 4%/Minute

Case 3: Plant Response to Gasifier Fluctuations, Turbine-Lead Control Mode

Figure 5.11 shows the response of the GCC system in the turbine lead control mode to fluctuations in gasifier output. The plant megawatt demand is constant for this transient. The gasifier fluctuations are initiated at 40 seconds and the plant control system attempts to maintain a constant megawatt output. In the turbine lead control mode (see Figure 5.1) the gas turbine fuel control responds to regulate plant output to the desired set point and the proportional-plus-integral control on gasifier steam and oxidant feed attempts to maintain the pressure at the turbine fuel valve at the set level.

The fuel gas lower heating value at the gasifier exit varies by 4.3% peak to peak but remains constant at the gas turbine fuel valve due to the attenuating effect of the large volume of the cleanup system. The gas turbine fuel flow varies due to the flow pressure disturbances and the result is a slight increase or decrease in gas turbine power output. Variations in steam turbine output lag the changes in gas turbine output. The resulting plant power output varies by 0.6% for the gasifier fluctuations modeled here. This level of variation in output is acceptable in normal power system operation, but it should be minimized if possible.

Figure 5.12 shows the gasifier and cleanup system response to gasifier fluctuations. Due to the capacitance effect of the large volume of the cleanup system, fluctuations at the gasifier outlet are attenuated by the time they reach the gas turbine fuel valve. The pressure sensed at the gas turbine control valve lags the disturbance at the gasifier considerably and the attempt to regulate the pressure by adjusting gasifier steam and oxidant feed may worsen the situation by moving the feed in the same direction as the fluctuation.

Figure 5.13 gives the gas turbine response to the fuel system fluctuations and Figure 5.14 presents the steam turbine response. The variations in operating condition are slight and should present no problem to these systems.

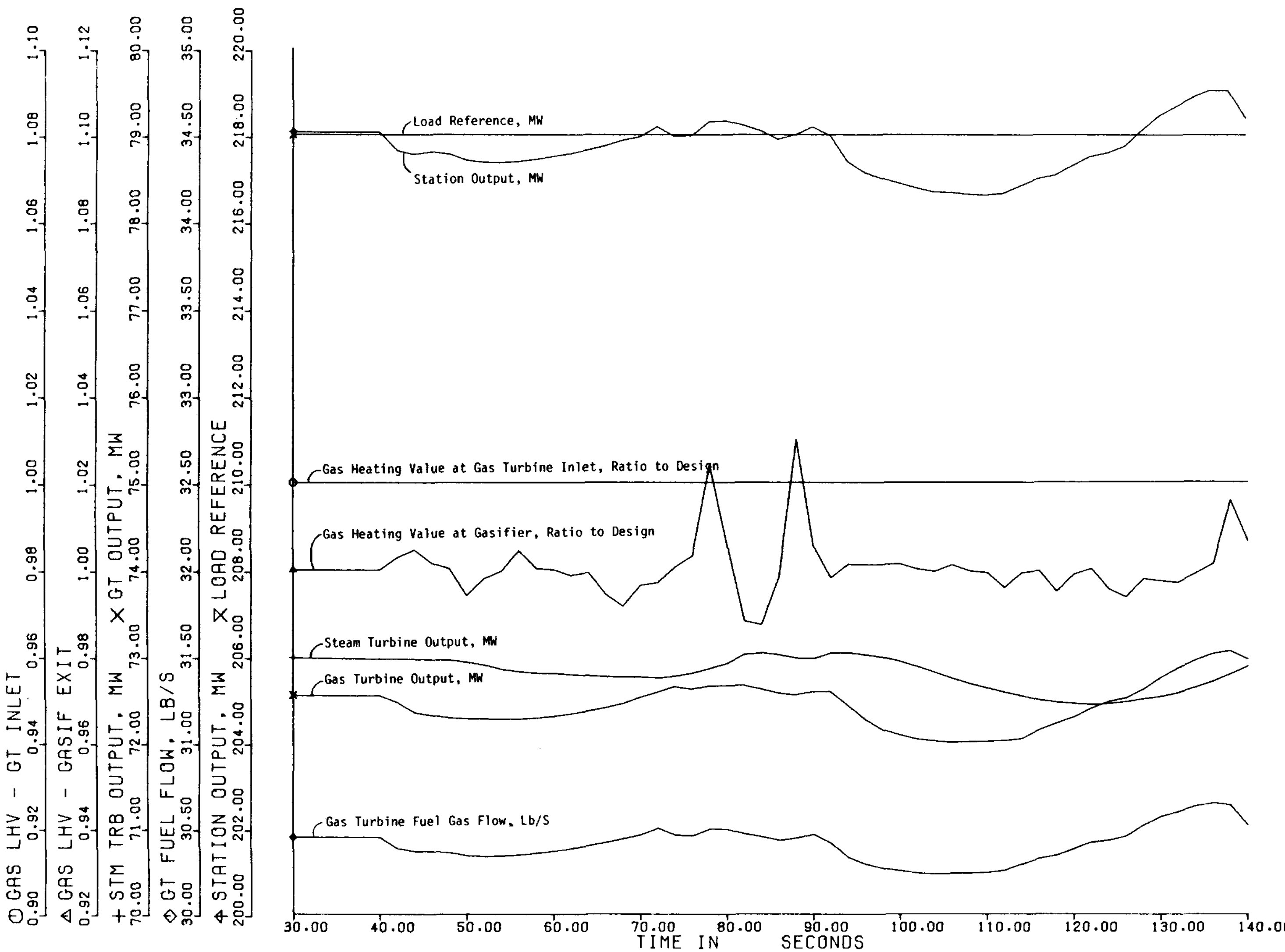


Figure 5-11. Case 3-GCC Plant Megawatt Output and Heating Value Transients in Turbine-Lead Control Mode in Response to Gasifier Fluctuations

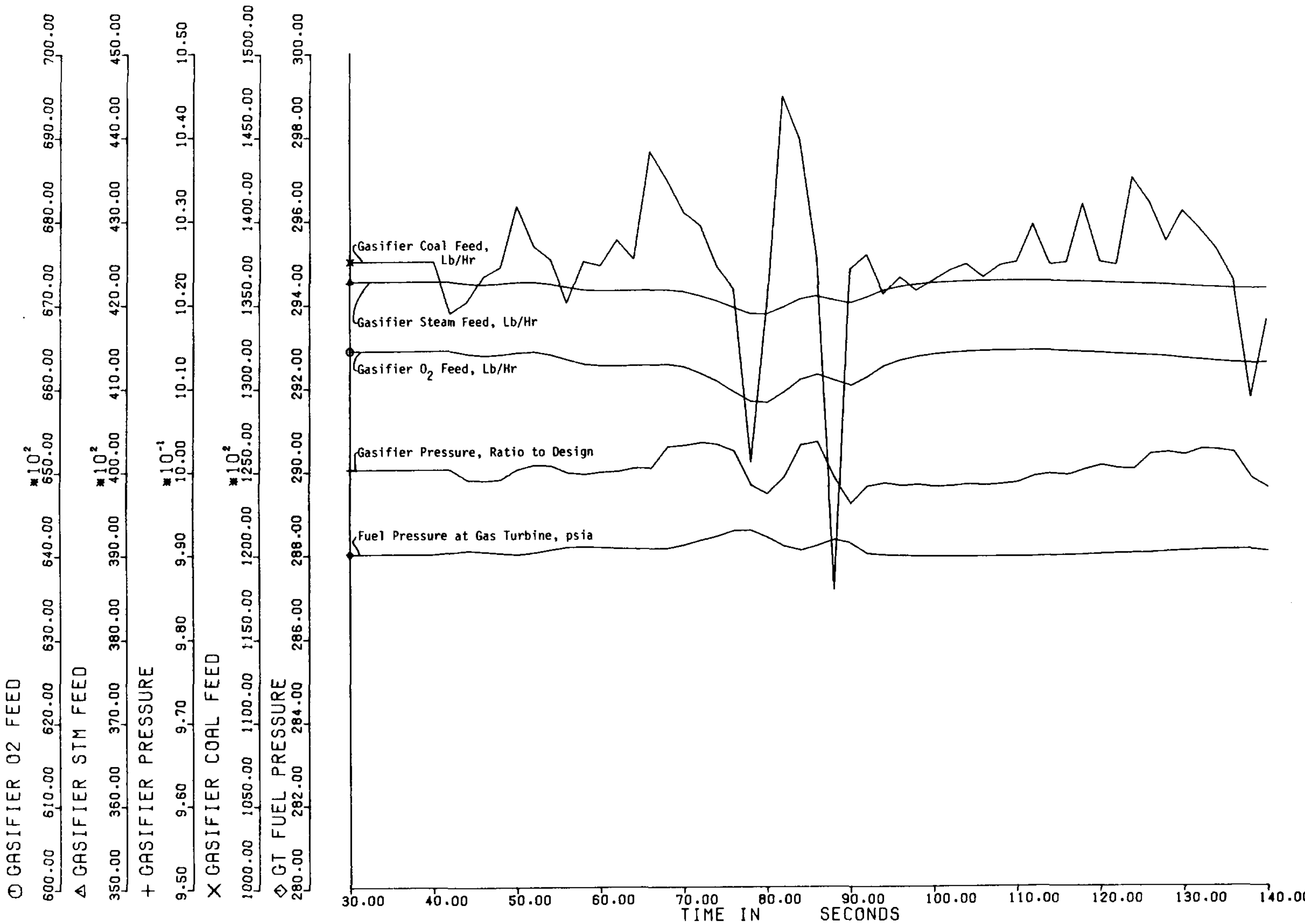


Figure 5-12. Case 3-GCC Plant Gasifier Response During Gasifier Fluctuations, Turbine-Lead Control Mode

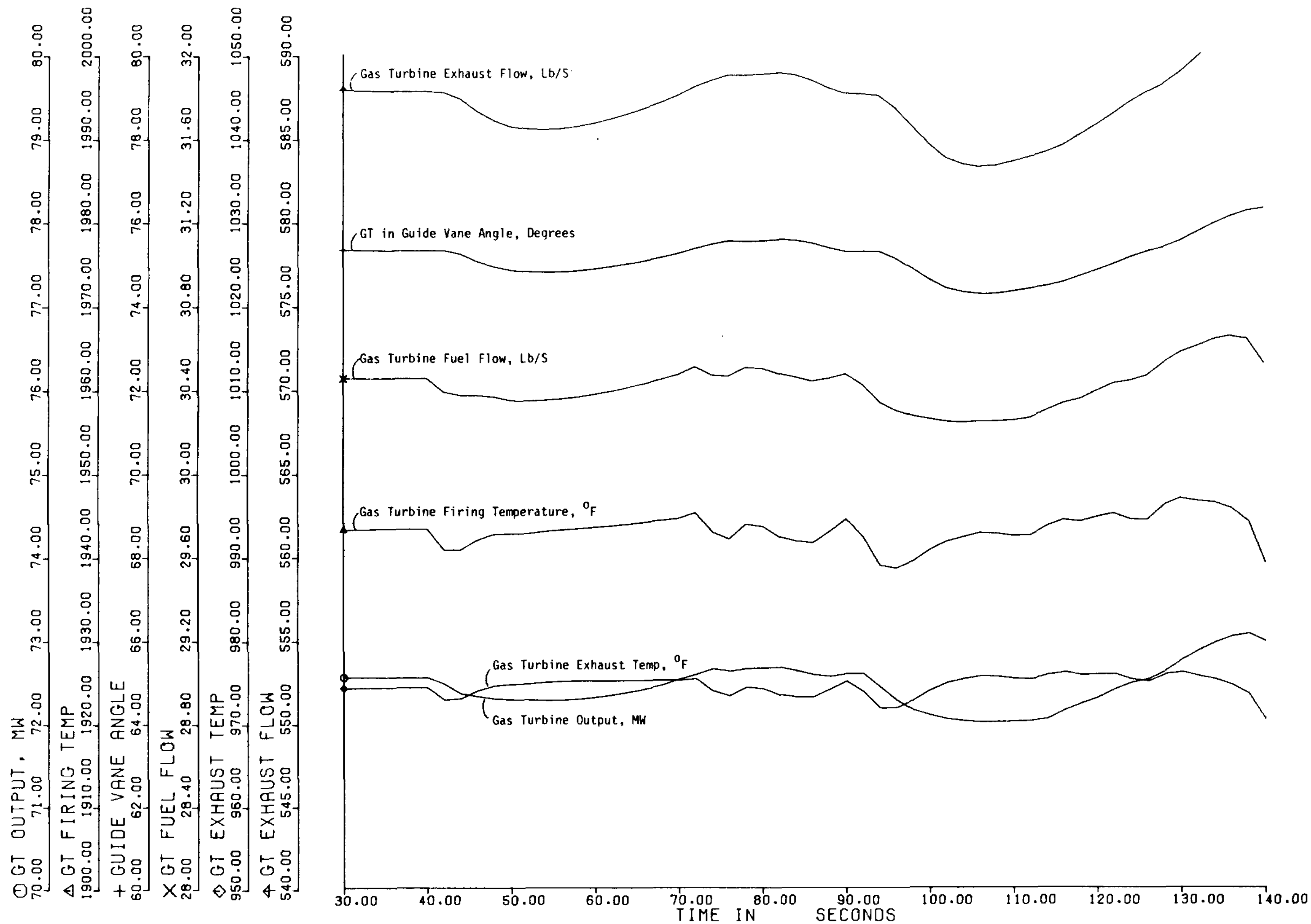


Figure 5-13. Case 3-GCC Plant Gas Turbine Transient in Turbine-Lead Control Mode in Response to Gasifier Fluctuations

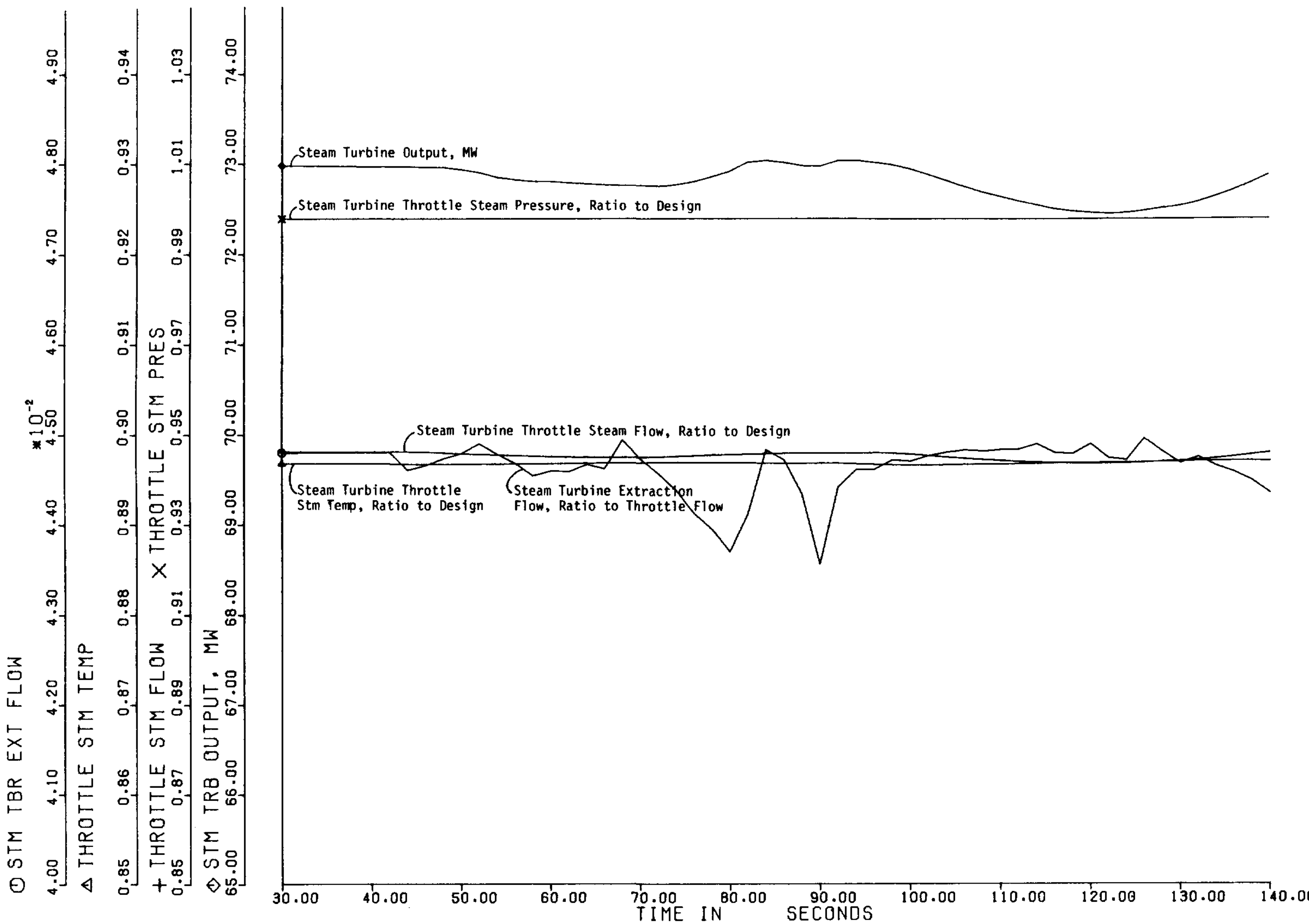


Figure 5-14. Case 3-GCC Plant Steam Turbine Transient in Turbine-Lead Control Mode in Response to Gasifier Fluctuations

Case 4: Plant Response to Gasifier Fluctuations, Gasifier-Lead Control Mode

In this transient, as in the previous case, the plant megawatt set point was held constant and gasifier fluctuation initiated at 40 seconds to observe the effects of gasifier fluctuations on individual components and plant output in the gasifier lead control mode. In this control mode, gasifier feed is controlled through a proportional plus integral controller to regulate plant power output. The gas turbine fuel valve is controlled by a proportional controller with 4% droop to maintain pressure at the fuel valve.

The gasifier flow fluctuations are the same as that imposed in Case 3, but the overall system response is different due to the different closed loop control action. Figure 5.15 shows the plant response to the gasifier flow fluctuations. Plant power output is maintained within plus or minus 0.8% of the set point. Gas turbine power output follows gas turbine fuel flow and steam turbine power output lags the gas turbine output. In this instance, changes in steam turbine power output are contrary to the changes in output of the gas turbine. That is, when the gas turbine power output goes up, the steam turbine power output goes down and vice versa. As in Case 3, the lower heating value of fuel gas varies more than 4% at gasifier exit but hardly at all at the gas turbine fuel valve.

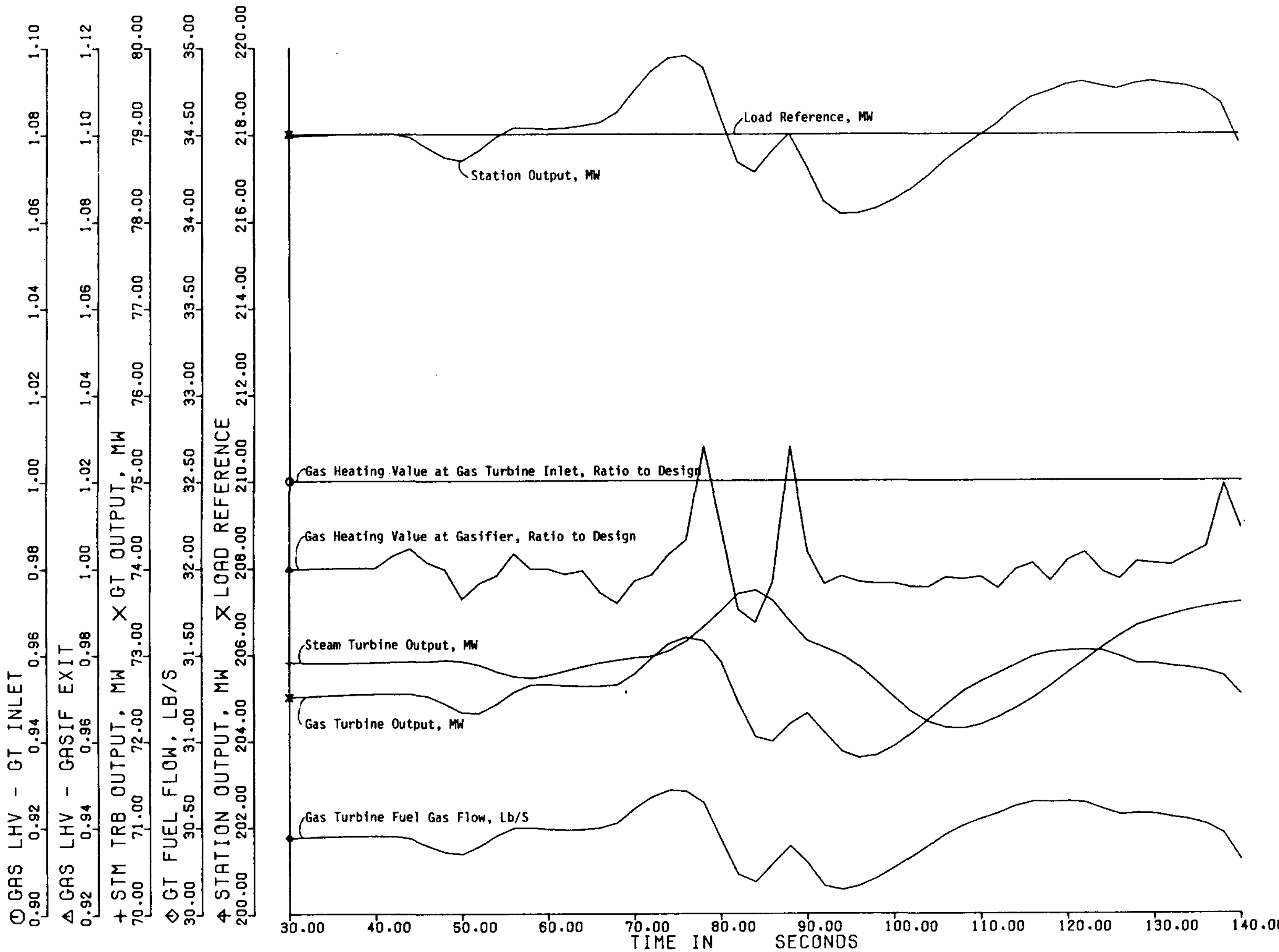


Figure 5-15. Case 4-GCC Plant Megawatt Output and Heating Value Transients in Gasifier-Lead Control Mode in Response to Gasifier Fluctuations

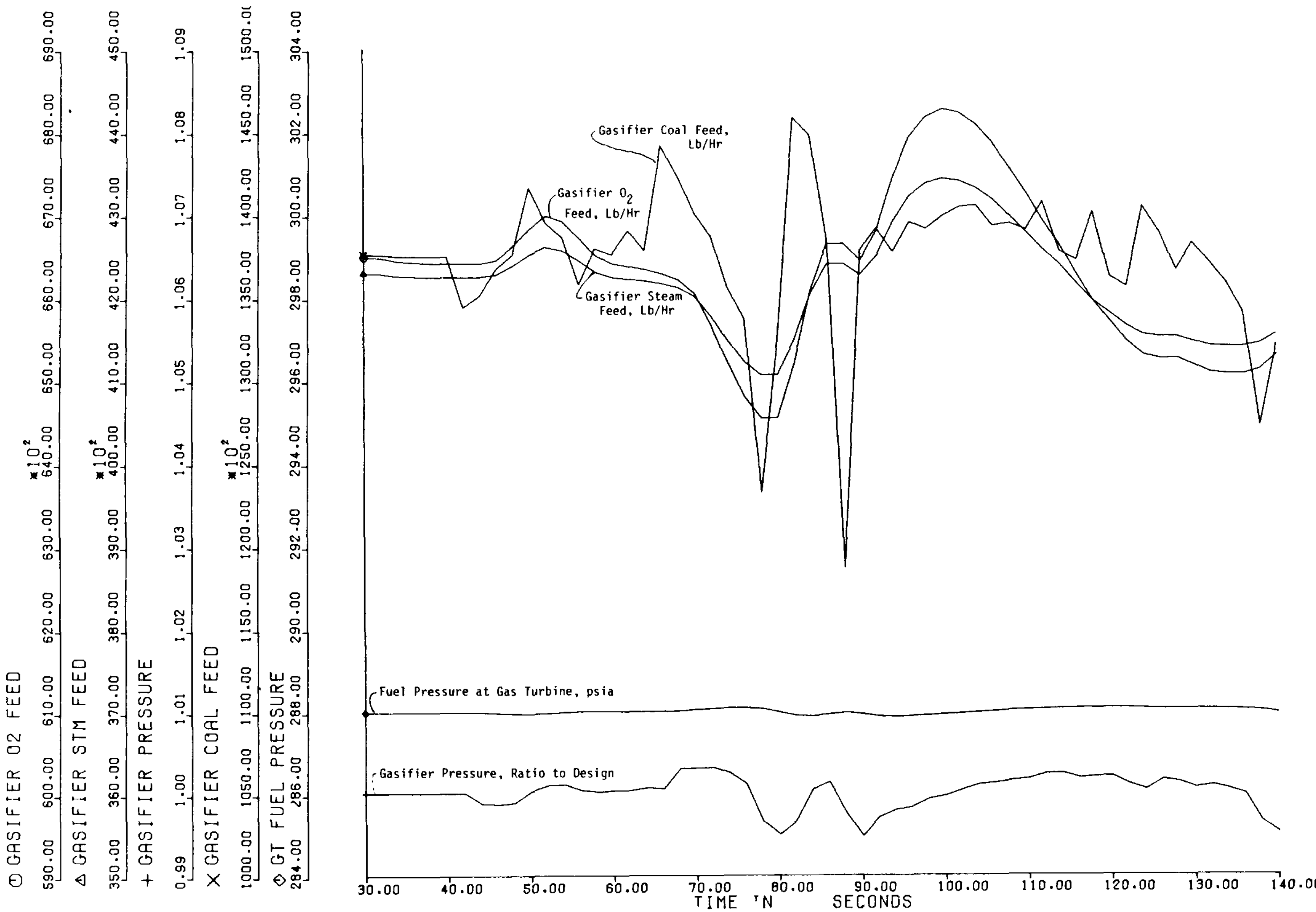


Figure 5-16. Case 4-GCC Plant Gasifier Response During Fluctuations, Gasifier-Lead Control Mode

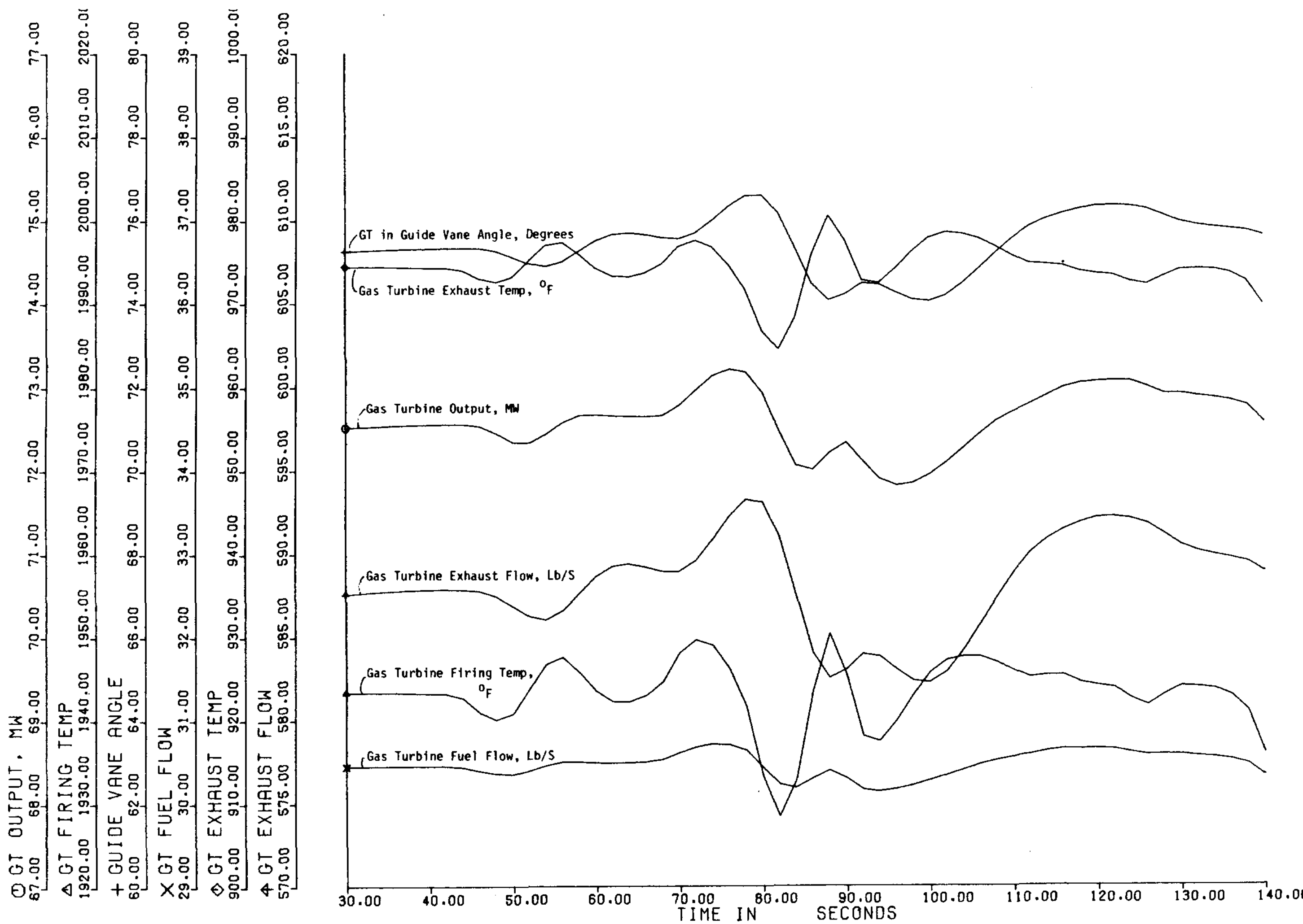


Figure 5-17. Case 4-GCC Plant Turbine Transient in Gasifier-Lead Control Mode in Response to Gasifier Fluctuations

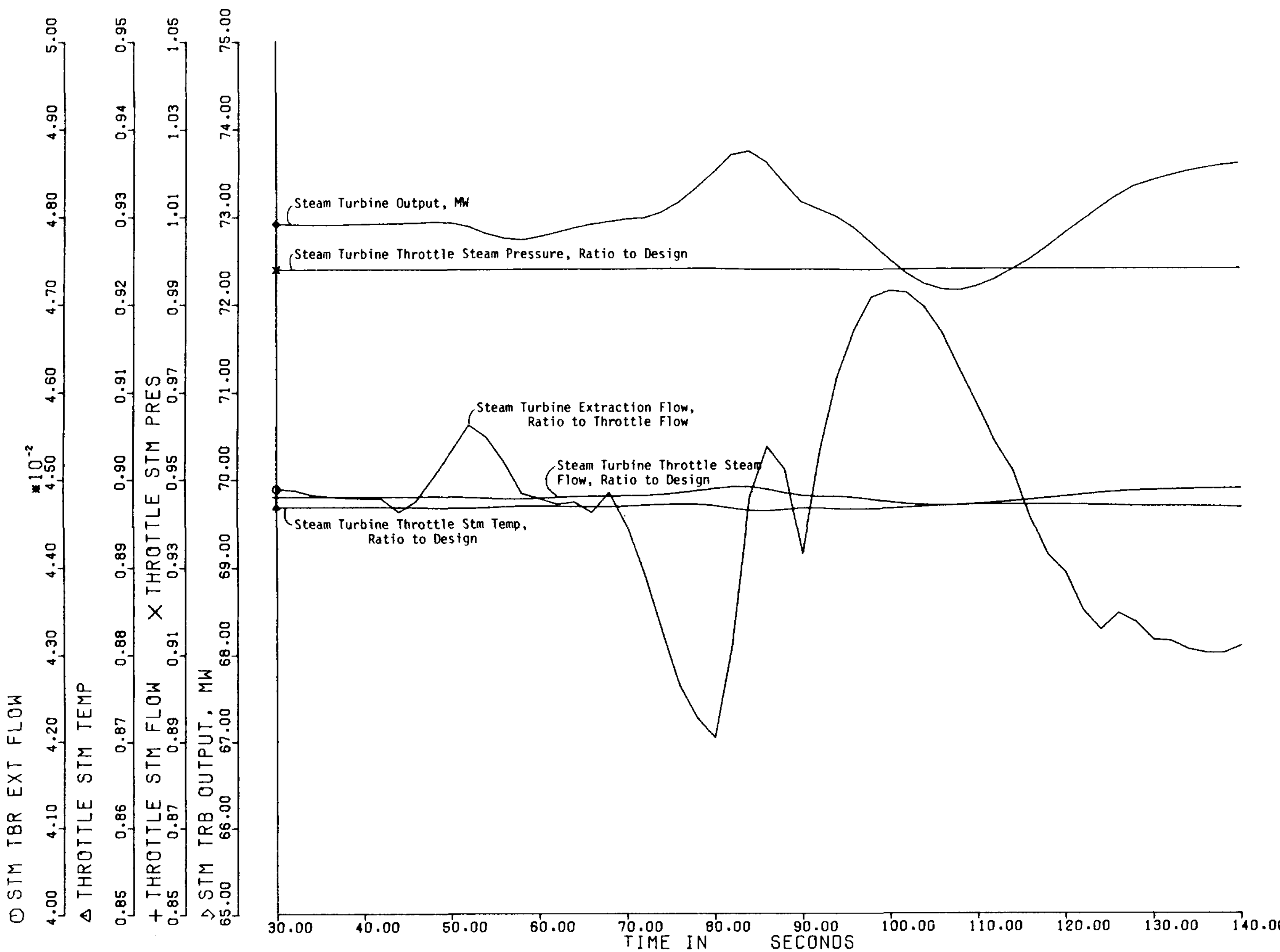


Figure 5-18. Case 4--GCC Plant Steam Turbine Transient in Gasifier-Lead Control Mode in Response to Gasifier Fluctuations

Section 6

CONTROL ANALYSIS

A comparative system and control analysis is presented in this section. The oxygen-blown slagging GCC is compared to the air-blown, dry-ash GCC.

POWER SYSTEM REQUIREMENTS

The power system requirements for either plant are the same. See Section 6 of volume 1 of this report (3). The average power system response requirement is on the order of 2% per minute or less for daily load following and 2-5% per minute for tie line regulation.

SYSTEM COMPARISON

A GCC system using an oxygen-blown slagging gasifier is similar in many respects to an air-blown system and yet the differences can significantly affect plant dynamic response. The basic configurational difference is that oxygen is supplied by an air separation plant for the oxygen-blown gasifier while air is extracted from the gas turbine compressor for the air-blown unit. In this model, it was assumed that the compressors in the air separation plant were electrically driven with power being supplied by the grid. This assumption should not significantly affect conclusions of this study as air separation plant and oxidant compressor power requirements are about 6% of gross output. Plant control requirements for the air separation unit and the oxygen compressor must be addressed, however, if the air separation plant is to be closely integrated with the gasification plant. The situation analyzed here is that of over-the-fence oxygen supply. In any event, the dynamic requirements of oxidant supply must be addressed and limitations assessed when actual plant components are selected for any specific oxygen-blown slagging GCC power plant design.

Another difference is that the process steam requirement for the oxygen-blown gasifier is significantly reduced from what the air-blown dry-ash gasifier requires. As a direct consequence, the amount of steam extracted for the gasifier is less and a larger percentage of plant power is produced by the steam turbine. The end result is that changes in gasification process steam demand affect the plant power output to a much lesser degree than in the air-blown dry-ash gasification system.

Additionally, there is no gas saturator in the oxygen-blown system used for the slagger study, as in the earlier air-blown study. Therefore, the moisture content of the fuel gas is much less than in the air-blown case. Because of this, variation of fuel heating value due to varying amounts of moisture in the fuel gas are not a problem in the oxygen-blown slagging GCC system.

Another difference between the two systems is that in the oxygen-blown system the volumetric flow of fuel gas per Btu delivered to the gas turbine is reduced for two reasons. One is because nitrogen is separated out in the air separation plant and does not dilute the fuel. Another reason is that the fuel delivered to the gas turbine in the oxygen-blown slagging GCC has a low moisture content. The end result is that variations of fuel gas heating value at the gas turbine are considerably reduced from what they are in the air-blown dry ash GCC. Gasifier output (in terms of fuel gas heating value) is more closely controlled by regulating gasifier feed for the oxygen-blown gasifier. Coal devolatilization, while a large contributor, does not predominate in determining the heating value of the fuel gas from the oxygen-blown moving bed slagging gasifier. This is because lower the heating value (on a volumetric basis) of the gaseous products of devolatilization (neglecting H_2S) is on the order of 30% greater than the lower heating value of gas emanating from the gasification zone of the slagging gasifier. In the air-blown system the heating value of the products of devolatilization are about 300% greater than the gas coming from the gasification zone.

A significant result of the configurational difference between the air-blown dry ash GCC system and the oxygen-blown slagging GCC system is that the oxygen-blown gasification process is not as closely coupled with the power generation turbines. Also, the selective removal of H_2S for the oxygen blown system is less difficult because the concentration of H_2S in the fuel gas is higher (since it is not diluted with nitrogen).

The oxygen-blown slagging GCC is able to perform load changes of at least 4% per minute and probably faster. The limiting factors are the same as for the air-blown dry-ash GCC; and they include carryover of fines into the cleanup system and carryover of solvent from the H_2S absorber. The real limitations can only be determined from actual component testing. The oxygen-blown system is able to follow load demand much more tightly than the air blown system because the gasification process and power generation process are essentially de-coupled. The large amount of gas turbine compressor extraction air required for the air-blown plant always acts in a manner opposite to the desired change. For example, if an increase in power were desired, the extraction air fed to the gasifier must be increased, decreasing the amount of air available for the turbine. If a decrease in power were desired, the extraction air is decreased making more air available for the turbine. Since no extraction air is required in the oxygen blown-system, the gas turbine is able to respond immediately as in a conventional fuel system.

CONTROL STATEGIES

Gas Turbine-Lead Control Mode

In this control mode the gas turbine fuel valve receives a command signal from the station controller to increase, decrease or maintain plant power output depending on the measured difference between megawatt demand (i.e., set point) and plant output. The combined cycle then operates in a conventional mode, subject to normal limiting and protective functions of the gas turbine control. The gasifier oxidant and steam feed are controlled by a proportional plus integral controller to mantain the fuel gas pressure at the turbine fuel valve at a set value. This control mode and sensor setup was used in the air-blown dry ash GCC study and was simply carried over for use in this oxygen-blown slagging GCC study. Again, as in the air-blown study, no attempt was made to tune or optimize the system to achieve a given set of performance specifications.

For both the oxygen-blown, slagging and air-blown, dry-ash GCC, the turbine-lead control mode gave the best dynamic performance in tracking load demand. It offers the advantage that the combined cycle unit is controlled in a manner similar to the conventional combined cycle plant. The disadvantage is that the gasifier is controlled by sensing a pressure far removed from the gasifier, and therefore control action may not be in phase with that required at the gasifier. For instance, in Case 3 (described earlier in Section 5 of this report), gasifier pressure fluctuations are simulated and the controls acting in a direction which worsens the effect of the gasifier fluctuation. The gasifier fluctuations were amplified by a positive feedback effect created by inherent time lags in the control system. The cleanup system is also subject to these fluctuations; operation of the cleanup train and the combined cycle would be more stable if the fluctuations could be minimized.

Gasifier-Lead Control Mode

In the gasifier-lead control mode the plant load controller regulates gasifier feed through a proportional-plus-integral controller to maintain plant power output at the desired level. Fuel system pressure is regulated by the gas turbine fuel control valves with a proportional control having 4% droop or proportional band (i.e., a controller gain of 25-to-1).

When operating in this mode, the gas turbines consume fuel gas at whatever rate it is provided from the cleanup system at the set pressure, as long as the gas turbine control is not operating on a protective limit signal (e.g., in a temperature-protection mode). The large inherent volume of the cleanup system causes the output of the cleanup system to lag considerably behind its input. (Cleanup system residence time is on the order of one minute.) That is, the gas turbine cannot respond to a changing power demand until fuel gas flow through the cleanup system has responded first. Consequently, when the plant responds to a change in power demand, the power output lags the demand considerably. The results in section 5, Case 2, volume 2 of this report show that the plant responds as an overdamped system due to the large capacitance effect of the cleanup system.

The gasifier-lead control mode does not perform as well as the gas turbine-lead control mode in the situation where the gasifier output is fluctuating due to hang-slip, channelling or bridging phenomena. Again this is due to the effect of the large volume of the cleanup system. A change in power output of the plant requires a change in volume flow of fuel at the set pressure at the gas turbine fuel valve. The change in volume flow rate must be impressed across the cleanup system and the long residence time causes the change in flow at the exit to lag the flow at the inlet of the cleanup train.

Control Modifications for the Oxygen-Blown Slagging GCC

A possible way to minimize the effect of gasifier fluctuations in either control mode would be to sense gasifier exit pressure and control gasifier feed over a limited range about a reference value to hold gasifier exit pressure at a desired level. The reference value of gasifier feed would come from the plant pressure control in the gas turbine-lead control mode, or from the plant load controller in the gasifier-lead control mode. In the above scheme the nominal amount of gasifier feed would be set by either the plant load control (in the gasifier-lead control mode) or by the plant pressure control (in the gas turbine-lead control mode) and the gasifier local controls would bias the gasifier feed (set by either of the above controls) to minimize local fluctuations.

The fluctuations of gasifier output would be minimized, reducing the effect on the cleanup system and combined cycle. This would, however, require rapid changes in steam and oxidant feed to the gasifier. In effect, this approach shifts the fluctuating duty from the cleanup train-combined cycle to the steam and oxidant supply system. This fluctuating duty shift effect has been referred to in discussions of EPRI tests of the BGC/Lurgi gasifier pilot plant at Westfield appearing in EPRI report AP-1922 (7).

Figure 6.1 shows how this gasifier local fluctuation control scheme would fit into the gas turbine-lead plant control scheme; Figure 6.2 shows how it would fit into the alternate gasifier-lead plant control scheme.

As found in the air blown-dry ash GCC study, control logic must be added to keep the gas turbine load controller from controller saturation (or windup) when the gas turbine is operating on a protective or physical limit while the plant megawatt demand is greater than plant output. If this logic is not added, the load controller may become saturated and the ability of the plant to respond rapidly to a decrease in demand will be delayed while the plant load controller integrates back to the proper controlling regime.

Also, as was found in the air-blown dry-ash GCC study, lock hopper operation causes a temporary reduction in plant output while the lock hopper is being pressurized with fuel gas. One way the reduction power could be minimized would be to bleed a small amount of fuel gas continually and store it for use in pressurizing the lock hopper. The plant response when using this or other schemes can be evaluated readily with a dynamic simulation.

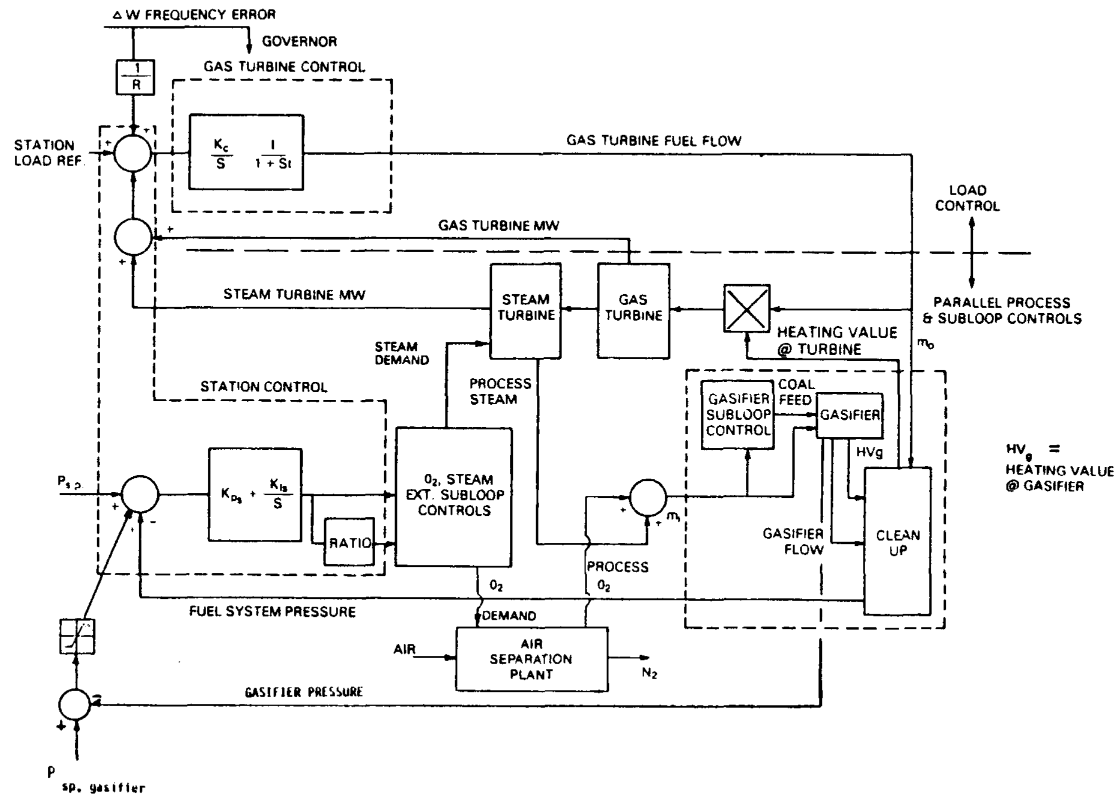


Figure 6-1. Plant Control Schematic for Gasifier Fluctuation Control in the Gas Turbine-Lead Mode

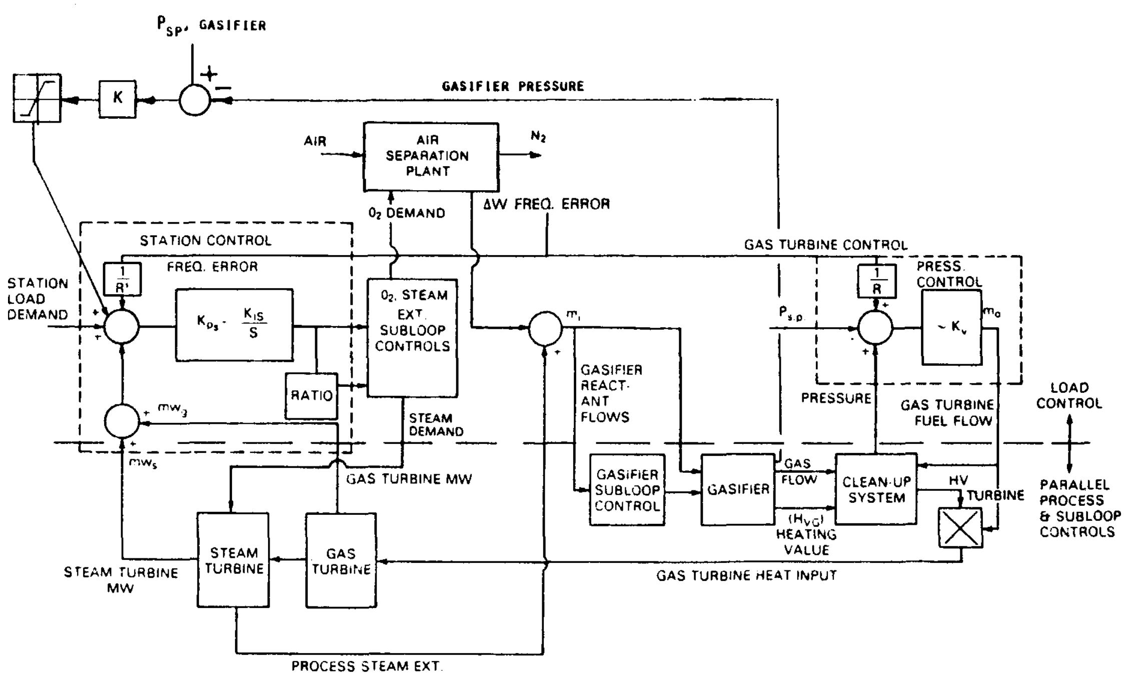


Figure 6-2. Plant Control Schematic for Gasifier Fluctuation Control in the Gasifier-Lead Mode

RECOMMENDATIONS

Dynamic simulation fits into the scheme of overall plant design as a tool to be used for evaluating:

- plant control strategy
- plant operations strategy
- plant transient performance

Dynamic simulation is effectively used in an iterative fashion throughout the plant synthesis and design cycle to evaluate the ability of hypothesized plant equipment configurations and control schemes to meet power system operational and transient performance requirements. It facilitates the prediction and evaluation of transient requirements of plant equipment and allows for the synthesis of coordinated controls used to minimize possible life shortening component duty and yet meet required response. With the advent of digital computer based control systems, the dynamic simulation model may be used in the implementation of the actual plant control. Furthermore, the computer models developed and used in the plant dynamic simulation may also be used in a plant simulator for training of plant operating personnel.

It is with the perspective of the previous discussion on the application and benefits of process and control dynamic simulation that recommendations are made. Future EPRI efforts to facilitate the incorporation and effective use of process and control dynamic simulation in the entire plant design cycle as outlined above will be to the benefit of utility plant owners and operators. Ability of a plant to meet transient performance requirements and operational requirements would be fully considered in the plant design rather than incorporated as an afterthought. Effective design will require interaction between the plant designers and owner/operators to make trade-offs in costs, performance, reliability, control and operational considerations and can be realized through use of plant process and control dynamic simulation.

Verified Component Models

The end use of dynamic simulation results is limited by the accuracy of simulation models of the plant components. Component models also must represent component operational limits. It is recommended that models of plant components in appropriate detail, verified to be suitably accurate by actual component testing, are required to perform a useful dynamic simulation.

Plant Specific Operation and Control

For a specific plant, the overall plant operations scheme and control need to be integrated. For example, it is possible that the plant could be operating at part load for a considerable time, and it may be advantageous from a heat rate point of view to turn down only one gas turbine (if the system power needs are met at this level of output). This, in turn, requires that a specific plant control scheme be formulated to operate the overall plant efficiently, i.e., while individual gas turbines are at difference load point.

Also, for a specific plant, dynamic simulation can provide insight into off-normal operations following the loss of key process components. It is recommended that plant specific simulation models of sufficient detail be developed for careful examination of such normal, near normal or contingency operations. Similarly, more comprehensive study of multi-uinit and multi-train plant loading alternatives should be considered.

Coordinated Plant Control Strategy

There is potential benefit to coordinated control of the plant and its component. Using component models appropriate in level of detail and accuracy, a coordinated control scheme may be developed to yield the best possible results in terms of plant transient performance, operability and reliability. For example, severe pressure or temperature transients could jeopardize the reliability of key plant equipment. By developing specific load-change schedules for key process components, a fully coordinated control strategy could evolve that would reduce the transient duty of these critical components and yet still meet overall plant transient performance requirements. It is recommended that coordinated control be studied on as near a realistic plant basis as possible and that benefits be identified.

Section 7

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APPENDIX A

K.J. Daniel, "Transient Model of a Moving Bed Coal Gasifier," Paper 52f, A.I.Ch.E. 88th National Meeting, Philadelphia, Pa., June 1980.

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TRANSIENT MODEL OF A MOVING BED COAL GASIFIER

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ABSTRACT

A transient model for air and oxygen blown moving bed coal gasifiers has been developed, and the use of the model shows good agreement with available steady-state data. It was found that the major factors affecting transient performance are the amount of thermal energy stored in the coal bed, and the rate of drying and devolatilization of the raw coal near the top of the gasifier.

SCOPE

Combined cycle power plants utilizing coal gasifiers as a source of gaseous fuel appear to be a viable power generation alternative for the future. One attractive configuration for such a plant employs a moving bed gasifier such as the Lurgi-British Gas Slagger or the General Electric moving bed gasifier (GEGAS). Steady-state performance estimates of such power plants have been made and are satisfactory for evaluation purposes, but better understanding of the transient operating conditions is required. One of the important parameters that affect the dynamic response of the power plant is the transient performance of the moving bed gasifier. This report describes a transient mathematical model for a moving bed gasifier which is used to predict responses to step changes in the input blast conditons.

Four detailed moving bed coal gasifier models have been described in the literature (Yoon et al., 1978a; Schlich, 1977; Biba et al., 1978; Amundson and Arri, 1978). Kosky and Floess (1979) have recently developed a simplified model that correlates well with other detailed models. All of these models were developed to give information about the steady-state performance of the gasifier. The model of Yoon, however, has been used to estimate very long term transients (hours) of the gasifier. The model described here is not concerned with these long term transients that are more easily controllable. It is concerned with shorter transients, that occur on the time scale from 0.5 to 10 minutes.

The scope of this study is limited to obtaining transient response of the output gas when changes are made in input blast conditions. Since the model is not intended to produce detailed information of processes occurring with the gasifier, many of the complicated processes are approximated.

Conclusions and Significance

The results of this model were compared to steady-state data from various sources (Hebden, 1975; Kimmel, et al., 1976; Chandra, et al., 1978). The model successfully predicted raw gas composition for both air and oxygen blown gasifiers.

The processes which were found to significantly influence the transient response of a fixed bed gasifier were: (1) drying and devolatilization of the coal and (2) thermal energy storage within the bed. It was found that for low steam/air ratio gasifiers, these two processes are approximately of equal importance in determining the transient change in the raw gas heating value. The thermal mass effects decayed on a longer time scale than the drying and devolatilization effects. In addition to these major processes, there is an interaction between changes in the temperature profile in the bed and the rate of the methane production reaction, which will also cause changes in the raw gas composition. However, this interaction has only a small effect since the model predicts that the majority of the methane is obtained by devolatilization.

Changes in the raw gas heating value following a transient were found to be moderate. For a rather severe transient of a 40% step decrease of blast flow rate, the heating value of the raw gas increased by approximately 15%.

The predicted response of the gasifier has been fitted with a simple exponential decay curve to facilitate the simple representation of a moving bed gasifier in integrated power plants without sacrificing accuracy.

Development of the Model

Two major approximations of this model are that extremely short duration transient effects are approximated as instantaneous and long duration effects are approximated as being constant. Transients of an intermediate time scale are modeled in detail. For this reason the model provides information only for transients on a time scale from one-half to several minutes. Examples of transients which are assumed to occur immediately are: (1) changes in the combustion zone output conditions; and (2) propagation of gases through the reactor of gases through the reactor. Examples of transients which occur slowly and are assumed to be constant are: (1) the location of the combustion zone; and (2) the height of the coal bed. As a result of these assumptions, the characteristics limiting the transient response of the gasifier are the thermal mass of the bed and drying time of the raw coal.

Following the analysis of Yoon (1978) both solid and gaseous phases are assumed to be at the same temperature throughout the gasifier. While this approximation is adequate for the prediction of trends in the output gases, it should be used with caution when attempting to accurately predict conditions within the bed. Experimental data reported for Lurgi gasifiers show that the temperature difference between phases can be as much as 700^{0}K near the combustion zone (Rudolph, 1972).

Combustion Zone

Conceptually, a moving bed coal gasifier can be divided into three zones as shown in Figure 1. The primary process occurring near the bottom of the gasifier is combustion. In this region, oxygen is consumed and heat is liberated. In most gasifiers this zone is small and will respond rapidly to changes in blast conditions. In order to eliminate short time constants associated with rapid transients and therefore make the problem tractable, it is assumed that the combustion is at steady state and responds immediately to changes in the blast. In reality the response of the combustion zone will require approximately 30 sec.

The steady state combustion zone model is based on the following chemical reactions:

$$C + 1/2\, O_2 = CO \qquad (a)$$
$$C + O_2 = CO_2 \qquad (b)$$
$$2H_2 + O_2 = 2H_2O \qquad (c)$$
$$2CO + O_2 = 2CO_2 \qquad (d)$$
$$H_2O + C = CO + H_2 \qquad (e)$$
$$H_2O + CO = CO_2 + H_2 \qquad (f)$$

Reactions (a) and (b) are simply the combustion of carbon which are very fast reactions with equilibrium strongly favoring the right-hand side. Reaction (c) is the combustion of hydrogen which is an extremely rapid reaction with equilibrium strongly favoring the right-hand side. Reaction (d) is the combustion of CO which occurs rapidly, but does not go to completion. Reactions (e) and (f) are the carbon steam reaction and the water-gas shift reaction respectively. Because reaction (c) is extremely rapid it will consume most of the hydrogen which is produced by reactions (e) and (f). Consequently, only a neglibible amount of hydrogen will be produced in the combustion zone. This has two major effects. First, the water gas shift reaction will not begin to attain equilibrium until O_2 is depleted so that reaction (c) becomes

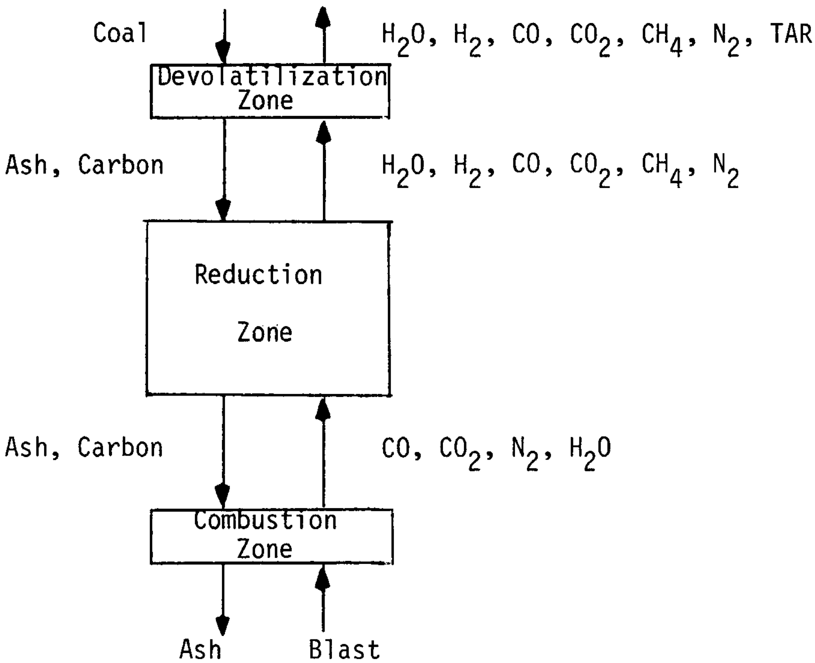

Figure 1. Conceptual Zones in a Fixed Bed Gasifier

unimportant. Second, only CO and CO_2 will be produced in the combustion zone and the concentration of water in the combustion zone will remain constant.

In an alternate approach, Yoon (1978) has assumed that the water gas shift reaction is in equilibrium throughout the combustion zone. Neither Yoon's assumption for the combustion zone, nor the one proposed here, has been fully justified. In any event, the difference of reaction mechanism occurring in the combustion zone does not affect the energy balance of the lower part of the gasifier or significantly affect the gas composition throughout the gasification zone.

Based on the previous discussion of reaction mechanisms, a simplified yet plausible model for the combustion zone can be written which consists of four equations. The first is hydrogen balance

$$\phi_{H_2O,\ in} = \phi_{H_2O,\ out} \tag{1}$$

The second is an energy balance

$$\dot{H}_{gas} = \dot{Q}_{React} + \dot{H}_B \tag{2}$$

The sensible energy of the solid species entering and leaving the combustion zone is assumed to be negligible. The third is an oxygen balance (Eq. 3).

$$\phi_{O_2,\ in} = \frac{1}{2}\,\phi_{CO,\ out} + \phi_{CO_2,\ out} \tag{3}$$

The fourth is an equation that determines the ratio of CO to CO_2 emerging from the combustion zone that takes the form

$$\frac{(P_{CO})^2}{P_{CO_2}} = \beta\ (\dot{m}_B)\ K_{eq}\ (T_{comb}) \tag{4}$$

where $\beta\ (\dot{m}_B)$ is an empirical function of the mass flow rate of the blast, $\dot{m}_B$.

Equation (4), with the exception of the function $\beta\ (\dot{m}_B)$ is obtained by combining the equilibrium expressions for reactions (a) through (f). It is unreasonable to assume that all of these reactions are in equilibrium at the top of the combustion zone. For this reason, the equilibrium constant, $K_{eg}\ (T_{comb})$, is multiplied by an empirical function $\beta\ (\dot{m}_B)$ which was chosen to fit experimental data of the peak combustion temperature in a fixed bed gasifier (Hebden, 1975). The dependence of β on the mass flow rate of the blast was assumed to be linear and was chosen to correspond to the single data point reported by Hebden (1975). Although the basis for Equation (4) may be tenuous, it reproduces the small amount of data available and it includes a reasonable temperature dependence for the overall reaction mechanism.

Equations (1) through (4) are solved simultaneously for the mass flux of CO and CO_2 leaving the combustion zone and the temperature of the combustion zone, T_{comb}. These quantities are then used as initial conditions for the gasification zone.

Gasification Zone

It is generally acknowledged that the overall chemistry in the gasification zone can be represented by reactions (e) through (h). Each of these reactions has been included in the model.

$H_2O + C$	=	$CO + H_2$	(e)
$H_2O + CO$	=	$CO_2 + H_2$	(f)
$CO_2 + C$	=	$2CO$	(g)
$C + 2H_2$	=	CH_4	(h)

It was assumed that the reactions (e), (g) and (h) are reversible and rate limited. It was assumed that reaction (f) is in equilibrium.

The species equation for the gasification zone then takes the form

$$\frac{\partial}{\partial t} C_i = \frac{\partial}{\partial x} C_i u + \sum_j \eta_j a_{ij} R_j (T) \tag{5}$$

The energy equation for the gasification zone is

$$\rho_c \, AC_{p,c} \frac{dT}{dt} = -\sum_i \Delta Hf^o{}_i \frac{\partial \dot{m}_i}{\partial x} - \sum_i \dot{m}_i \, C_{p,i} \frac{\partial T}{\partial x} - h\pi d \, (T-T_{wall}) \qquad (6)$$

where $\Delta Hf^o{}_i$ is the heat of formation of species i. This quantity, a weak function of temperature, is assumed to be constant (JANAF, 1971). The term on the left-hand side of Eq. (6) represents the rate of change of the stored thermal energy in the bed. The first term on the right-hand side of Eq. (6) accounts for changes between energy stored in chemical bonds and thermal energy. The second term on the right-hand side of Eq. (6) represents the change in the flux of sensible heat. The last term on the right-hand side of Eq. (6) represents the loss of heat to the wall.

After integrating Eq. (5) over the x coordinate, Eq. (6) was used to evaluate the time derivative of temperature for each axial step. A Gear implicit integration method was used to integrate over time.

Devolatilization Zone

The top portion of the coal bed is the devolatilization and drying zone. This section is treated differently than the gasification zone. The amount of steady-state devolatilization product is calculated using the product distribution in Table 1, the ultimate analysis of the coal, and the amount of fixed carbon consumed in the gasifier below the devolatilization zone. All hydrogen and oxygen in the coal is assumed to be given off in the proportion shown in Table 1. The volatile product distribution was chosen to fit GEGAS pilot plant gasifier data and has been adjusted to close the hydrogen, carbon, and oxygen balance.

Table 1

VOLATILE PRODUCT DISTRIBUTION

SPECIES	kg/kg DAF
H_2O (chemical)	.0730
Tars and Oils	.0563
CO	.0481
CO_2	.0232
H_2	.0173
CH_4	.1114

The first term on the right-hand side represents the change in mole flux to and from a location in the gasifier. The second term on the right-hand side represents the rate of chemical production or disappearance of a species. The term on the left-hand side represents the local rate of storage of species and is only significant immediately following a transient. It then rapidly decays to zero as chemical generation and species flux reach steady state. For the purpose of this study, the specie storage term is assumed to be zero. As a consequence, the effect of a blast transient is seen immediately in the raw gas composition rather than being delayed by the time required for gas propsagation through the coal bed. This delay is of the order of 10 sec. In addition, this assumption causes small errors in the predictions of the transient raw gas compositions.

Equation (5) is solved using only the heterogeneous reactions (f), (g) and (h). The solution is obtained by using a second order Runge-Kutta method, integrating from the bottom to the top of the reactor. The initial conditions are determined by the combustion zone model and an assumed initial temperature profile. After each finite step in the solution, water gas shift reaction equilibrium is imposed.

The rate constants that determine the rate of reaction are not well defined. In general, they are a function of heat treatment, percent burn-off, molecular structure of the coal and to some extent the inorganic species in the coal. Nevertheless, reactions (f), (g) and (h) were assumed to be reversible with rate constants of Arrhenius form. The rate constants used for reaction (e) and (g) were those used for Illinois No. 6 coal by Yoon (1978). The rate constants for reaction (h) were taken from Lowery (1963a). Only small amounts of methane are produced and consequently this reaction does not affect the results significantly.

The rate expression for reactions (e) and (g) are multiplied by a reaction efficiency that accounts for diffusion through the gas film, ash coating, and the reacting carbon (Ishida and Wen, 1968). The diffusion coefficient for all gaseous species was assumed to have the same temperature and pressure dependence as that presented by Reid (1977) and Field (1967) in addition to being proportional to the square of the porosity of the solid phase.

The expression for reaction efficiency assumes that the ash does not separate from the reacting coal particle. This assumption can be justified by recognizing that at the location where diffusion effects are important (the high temperature area in the lower gasification zone), ash occupies nearly half the bed volume and will probably still have a relatively high viscosity (Lowery, 1963b).

The rate of volatile product release during transients is difficult to predict because of uncertainties in properties, in the coal particle temperature distribution, and in the interaction between particle temperature distribution and devolatilization rate. For this reason, the volatile products are assumed to decay exponentially from the initial steady state value to the instantaneous steady state value as shown below

$$\dot{m}_{v,i} = \alpha_i \left(\dot{m}_c(t) + (\dot{m}_c(0-) - \dot{m}_c(t)) \exp(-t/\tau)\right) \tag{8}$$

where $\dot{m}_{v,i}$ is the mass rate of production of volatile species i, α_i is the proportionality constant determined from the volatile product distribution, $\dot{m}_c$ (0-) is the steady state coal consumption before a transient is imposed, and $\dot{m}_c$ (t) is the instantaneous coal consumption determined as if the volatile product release were in steady state with the instantaneous fixed carbon consumption. The magnitude of the time constant, τ, in Eq. (7) is a parameter that was varied in this study.

The energy equation for the devolatilization zone is

$$\rho_c A C_{p,c} \, \Delta x \, \frac{\partial T}{\partial t} = -\dot{m}_{cH_2O}(\Delta h_{vap} + C_{p,H_2O} \, (T_{vap} - T))$$

$$+ \left(\sum_i (\dot{m}_i \, C_{p,i} \, \Delta T\right) - h\pi \, d\Delta x \, (T - T_{wall})$$

where $\dot{m}_{cH_2O}$ is the rate of moisture vaporized from the coal input, h_{vap} is the latent heat of vaporization of the moisture and $C_{p_{H_2O}}$ is the specific heat of steam. Because of the low temperatures in this zone the chemical reaction rates are small. For this reason the chemical heat generation terms are not included in this equation. Moreover, devolatilization was assumed to be athermal as assumed by Yoon (1978).

Steady State Results

Table 2 lists the input parameters used for both the pilot plant and the full-scale gasifier.

Table 3 lists experimental data from the GEGAS-D pilot plant using a high steam/air ratio similar to that employed in Lurgi gasifiers. Also listed in this Table are the predictions of the model. Similarly, Table 4 contains pilot plant data and model predictions for low steam/air ratio conditions. The model reproduces the general trend between high and low steam/air ratios. However, the model predicts lower coal capacity and higher exit temperatures. One reason for these differences is the simplification of the combustion zone energy balance. Other simplifications such as the assumption of constant specific heat or constant heat of formation also probably contribute to the differences. Despite these differences, predicted gas composition is in reasonable agreement with the data.

Table 2

INPUT CONDITIONS FOR GEGAS PILOT PLANT AND FULL SCALE GASIFIER

	PILOT PLANT	FULL SIZE
Steam/Air (mass)	0.208	0.208
Pressure	2.07MPa	1.79MPa
Blast T	581 ^oK	608 ^oK
Heat Loss (% HHV Coal Throughput)	2.0	2.75
Height from combustion zone to top of coal bed	1.97 m	1.97 m
Diameter (ID)	88.9 cm	335. cm
Ultimate Analysis C(DAF)	0.7813	0.7726
(Illinois No. 6) H	0.0563	0.0592
O	0.1149	0.1114
S	0.0133	0.0139
N	0.0342	0.0429
Ash (AR)	0.0846	0.096
H_2O (AR)	0.110	0.02

Note: The full-scale gasifier volatile distribution is slightly different in order to insure mass balance for the slight variation in coal.

Table 3

COMPARISON OF EXPERIMENTAL AND ANALYTICAL DATA FOR HIGH STEAM:AIR RATIO

Input

Steam/Air (mass)	.686
Blast Temperature	561 ^oK
Gasifier Pressure	.689MPa
Blast Rate	.447 kg/S
Coal	Illinois No. 6

Output	Experimental	Analytical
H_2 % Vol.	13.7	16.5
CO	7.2	6.6
CO_2	11.3	12.4
N_2+Ar	30.5	30.0
CH_4	3.2*	2.8
H_2O	33.8	31.7
T_{exit} ^oF	792 ^oK	953 ^oK
Capacity kg/s	.118	.107

*$CH_{3.8}$

Confidence in the model is increased by the results shown in Figure 2. This figure compares the dry gas composition for several oxygen blown moving bed gasifiers to the gas composition predicted by the model for a commercial scale gasifier. The British Gas Slagger and the Lurgi data represents two extremes in the ratio of blast steam to oxygen employed in moving bed gasifiers. Overall, the model agrees well with the data.

All of the data shown in Figure 2 are for gasifiers using Illinois No. 6 coal except the experimental data of Hebden which was obtained using a bench scale moving bed gasifier, gasifying petroleum coke. This type of coke would be expected to have a much smaller methane yield because of its small amount of volatile matter. For this reason, the data of Hebden would be expected to show smaller amounts of methane. The results of this model confirm those of Yoon which show that practically all of the methane in the raw gas of a fixed bed reactor results from devolatilization.

Unfortunately, the numberical solution became unstable at low steam/oxygen ratios because of the increased gasification zone bed temperatures. This problem could be eliminated by neglecting methane production in the gasification zone when using the model for steam/oxygen molar ratios less than 1.4. Since the model predicts only a small amount of methane generation in this zone under these conditions, the total methane production is only decreased slightly and only slightly affects the other species concentrations. These changes are shown by a break in the continuous curves at the steam/oxygen ratio of 1.4.

The model predicted that the gasifier throughput of the British Gas Slagger and the Lurgi gasifier would differ by a factor of 2.99. This result is in reasonable agreement with the data presented at Kimmel et al., (1976) and Chanadra et al., (1978) which show a factor of 2.67.

<u>Transient Results</u>

The transient results for a 40% step decrease in blast flow rate while maintaining a constant bed height are shown in Figure 3 for the conditions corresponding to a full-scale gasifier (Table 2). This figure shows the raw gas heating value and the heating value of the gas leaving the gasification zone. Note that the transient imposed is rather severe. Normal power plant operation require a maximum change of 5% per minute. Nevertheless this figure illustrates the processes occurring following a transient.

Because the response of the combustion zone was approximated by an immediate response, an error is incurred for the first 30 seconds after the transient. Following this time period, all errors incurred due to this approximation are negligible.

The time constant for drying and devolatilization was estimated to be 90 s. The figure shows the raw gas heating value predictions for time constants of both 90 s and 0 s. From Eq. (8) it can be seen that the value of τ affects only the rate of decay of the effect of the drying and devolatilization zone, not the size of the effect. The size is determined by the amount of volatile matter in the coal and the magnitude of the change in gasification zone carbon consumption. The size of the transient effect caused by the devolatilization zone is the difference between the results for $\tau = 0$ s and $\tau = 90$ s. This effect is approximately the same size as that caused by the gasification zone.

The immediate increase in heating value of the gas leaving the gasification zone is caused by the relatively high initial temperature of the bed compared to its eventual steady state value. Figure 4 shows the decay of the temperature profile in the bed following this transient. The initial high bed temperature results in a large rate of reaction for the carbon steam reaction. As a consequence, a larger percentage of the steam reacts forming H_2 and CO thus increasing the percentage of these constituents in the raw gas. As the bed cools to its final temperature profile, less of the steam reacts and consequently the heating value of the gas drops. Surprisingly, the heating value of the gas does not return to its original value. There are two reasons for this. The first is that the methane production in the gasification zone does not decrease as much as the other constituents of the gas. The second is that slightly more water is reacted at the lower blast flux because of the longer residence time within the reactor.

Also shown in Figure 4 is the steady state temperature profile for a commercial scale gasifier operating at a high steam/air ratio. This profile is considerably lower than those shown for low steam/air ratio gasifiers and is in reasonable agreement with the predictions of Yoon (1978).

Figure 5 shows the change in the raw gas and gasification zone flow rates of the individual species. The molar flow rate of CO, CO_2, and H_2 exiting the gasification zone are almost identical to the flow rate of these species in the raw gas because devolatilization does not produce large quantities of these specifies. Consequently, the gasification zone molar flow rates of these species are not shown on this figure.

Table 4

COMPARISON OF EXPERIMENTAL AND ANALYTICAL DATA
FOR LOW STEAM:AIR RATIO

Input		
Steam/Air (mass)	.208	
Blast Temperature	466 °K	
Gasifier Pressure	2.06 MPa	
Coal	Illinois No. 6	
Output	Experimental	Analytical
H_2% Vol	16.8	14.7
CO	21.6	19.3
CO_2	5.7	8.1
N_2+Ar	40.1	41.9
CH_4	3.9*	4.3
H_2O	11.9	11.7
T_{exit}	866 °K	956 °K
Capacity kg/s	.209	.186

*$CH_{3.8}$

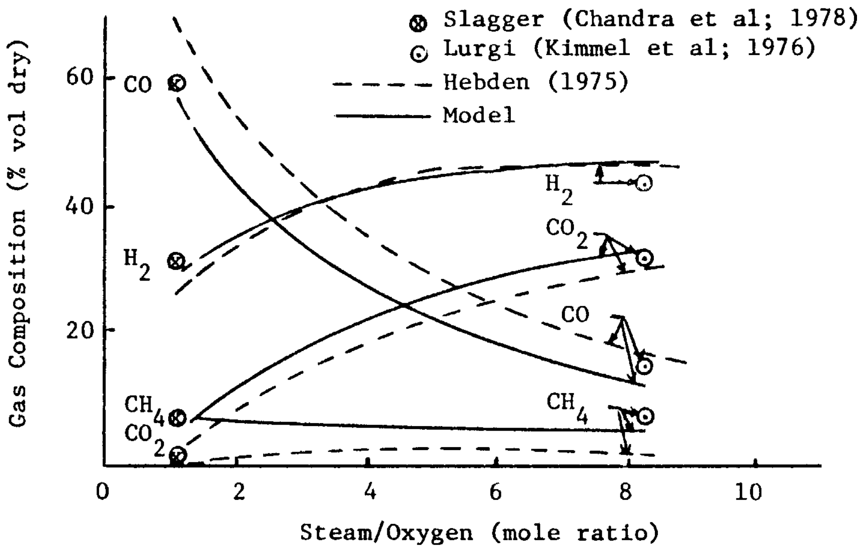

Figure 2. Dry Gas Composition Model Predictions for an Oxygen Blown Gasifier

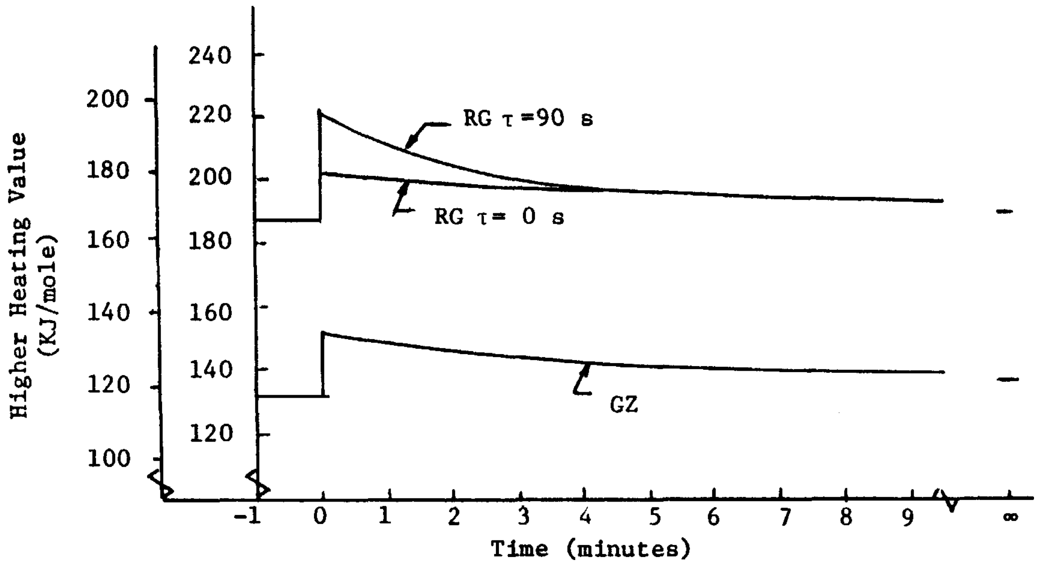

Figure 3. Heating Value of Raw Gas (RG) and Gas Exiting the Gasification Zone (GZ)

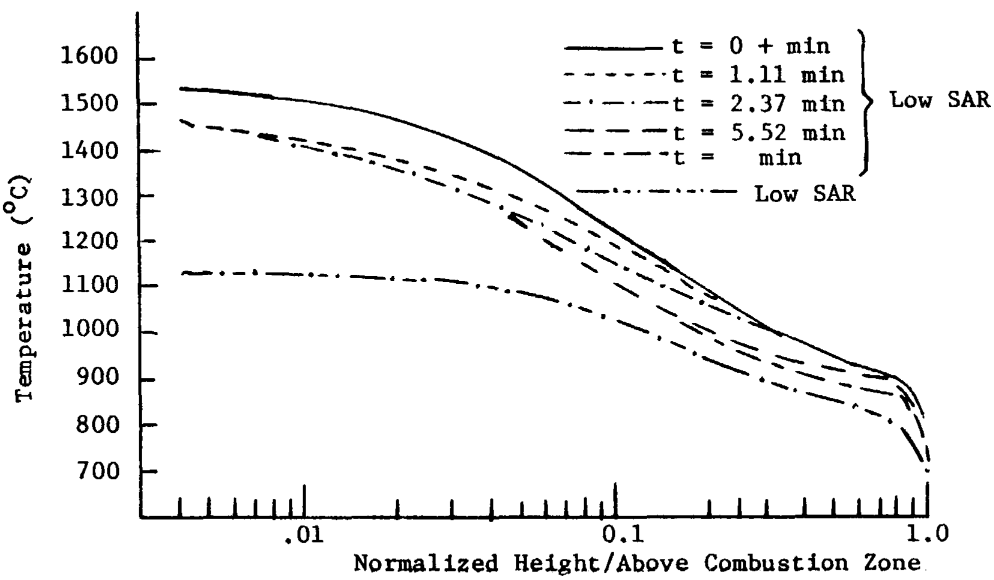

Figure 4. Predicted Temperature Profile Within the Gasifier

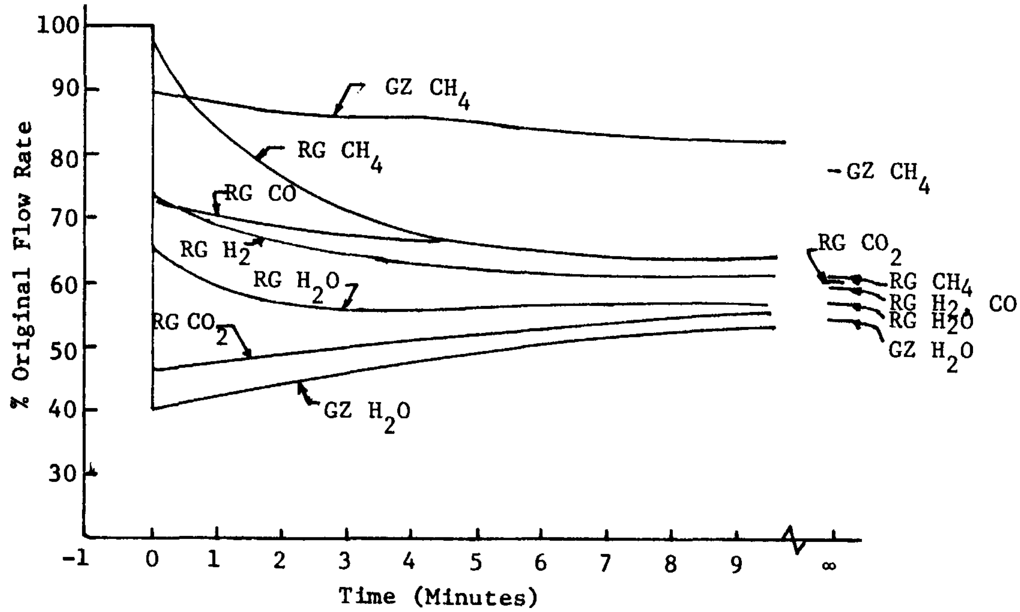

Figure 5. Change in the Flow Rate of Each Species Following a 40% Step Decrease in Blast (RG, raw gas; GZ, gasification zone exit)

Some interesting effects can be seen. The gas exiting the gasification zone shows a proportionately larger drop for water than carbon monoxide or hydrogen. The larger drop is due to the previously explained interaction between the bed temperature and the rate of the carbon-steam reaction. The amount of methane leaving the gasification zone is also affected by the transient. Initially, the rate of formation decreases slightly due to a decrease in the hydrogen concentration. Following this there is a long term decrease in the amount of methane produced caused by the propagation of the thermal wave up the bed. Since methane is produced only in the relatively cold upper regions of the gasification zone, methane transients are not seen as rapidly as transients in the other species which are formed in the hot regions deep in the bed. Methane formation in the gasification zone is minor. Consequently, transients in the raw gas methane content are dominated by changes occurring in the devolatilization zone.

The drop in heating value of the gas leaving the gasification zone following a decrease in blast can be approximated accurately by a least squares exponential curve fit of the form

$$\theta = 1 - e^{-t/\tau^*}$$

where θ is defined as

$$\theta = \frac{HHV(t) - HHV(t_{0+})}{HHV(t_\infty) - HHV(t_{0+})}$$

The value of * for this equation is a function of the magnitude of the step input. Table 5 lists the value of * obtained for various size blast step inputs. The cause for this variation can be traced to the basic nonlinearities of the problem.

Table 5

VARIATION OF PARAMETERS THAT DESCRIBE GASIFIER RESPONSE

$\Delta\% \, \dot{m}_B$	τ^*_{GZ}(min)	$\% \frac{HHV(t_{0+})-HHV(t_{0-})}{HHV(t_{0-})}$		$\% \frac{HHV(t_{0+})-HHV(t_{\infty})}{HHV(t_{0+})}$	
		RG	GZ	RG	GZ
-10	4.6	4.1	3.6	3.7	2.9
-20	5.1	8.6	7.0	7.6	5.1
-40	5.4	19.5	14.0	12.7	8.8

Also shown in Table 5 is the change of the heating value immediately following the step input and the magnitude of the decrease in heating value following the step input. Both of these quantities are listed for gas exiting the gasification zone and also for the raw gas. These quantities are roughly proportional to the magnitude of the step input.

Other methods of gasifier control give a similar result. Figure 6 shows how the heating value of the raw gas varies after a 20% decrease in the blast air flow rate. The same trend is evident. The heating value increases initially because the increased partial pressure of steam causes the carbon steam reaction rate to increase. However, this higher rate cannot be sustained because the peak combustion zone temperature decreases due to the higher steam content of the blast. As the bed cools, reactions slow and the heating value decreases.

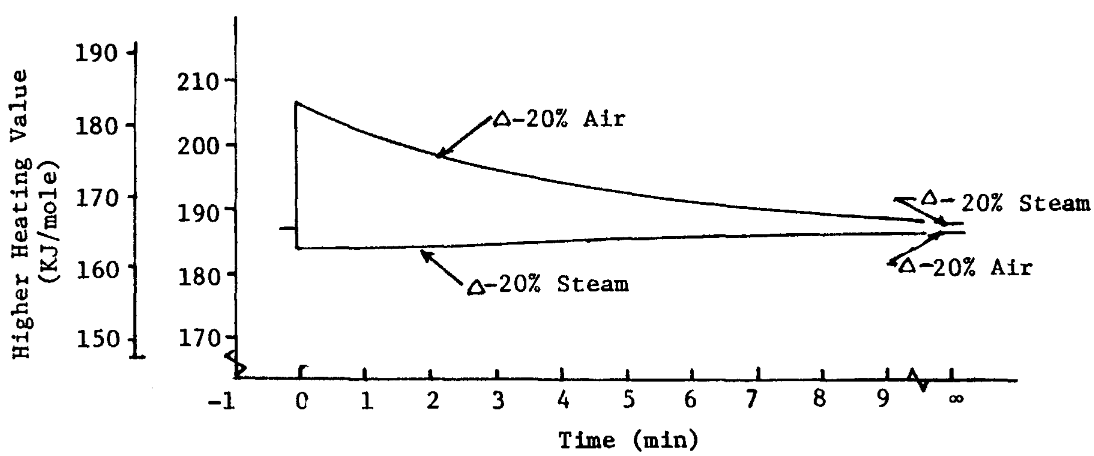

Figure 6. Change in Heating Value Following a 20% Decrease in Blast Steam and a 20% Decrease in Blast Air (RG, raw gas; GZ, gasification zone exit)

Decreasing the amount of steam in the blast has the opposite effect on gas heating value. Figure 6 shows the change in heating value of the raw gas after a 20% decrease in the steam contained in the blast. The heating value initially decreases because the partial pressure of steam decreases. This slows the carbon steam reaction. However, the decrease of steam in the blast causes the peak temperature in the bed to rise which feeds more sensible energy into the gasification zone. This eventually increases the temperature of the bed in the gasification zone causing the rate of the carbon steam reaction to increase. Consequently, as the temperature of the bed rises the heating value increases.

Summary

The major finding of this study is that despite rather severe changes in blast input conditions, a low steam/air ratio moving bed gasifier is well behaved during transients. There is nothing to suggest that it could not be integrated with a combined cycle power plant. The gasifier responds much faster than would be required by normal load changes of a utility power plant.

LIST OF SYMBOLS

A	cross-sectional area of gasifier cm^2
a_{ij}	stoichiometric coefficient of species i and reactant j
C_i	concentration of species i gm moles/cm^3
$C_{p,i}$	specific heat of species i J/kg
d	diameter of gasifier m
$\dot{H}_B$	rate of enthalpy entering the combustion zone due to blast gases J/S
$\dot{H}_{gas}$	rate of enthalpy leaving the combustion zone due to blast gases J/S
h	effect heat transfer coefficient to wall of gasifier
K_{eq}	equilibrium constant for water gas shift reaction
$\dot{m}_i$	rate of mass flow of species i kg/s
$\dot{Q}_{React}$	rate of energy release from combustion reactions J/mole S
P_i	partial pressure of species i Pa
T	temperature ^oK
T_{comb}	temperature of gases leaving the combustion zone ^oK
T_{wall}	wall temperature
t	time s
u	vertical gas velocity m/s
X	vertical coordinate m
α	proportionally constant that determines the volatile product distribution
β $(\dot{m}_B)$	empirical function chosen to yield correct peak combustion temperature
ΔHf^o	heat of formation J/mole
Δh_{vap}	heat of vaporization of water J/kg
η_i	reaction efficiency of reaction i
ρ	density gm/cm^3
τ	time constant min^{-1}
ϕ	mole flux mole/min

SUBSCRIPTS

B	blast
c	carbon
in	input
i	species
out	output
vap	vaporization
V	volatile material

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