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Analysis and Control of the METC Fluid-Bed Gasifier

Quarterly Report
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I. Overview of Present Work

This document summarizes work performed for the period 10/1/94 to 2/1/95. The initial phase of the work focuses on developing a simple transfer function model of the Fluidized Bed Gasifier (FBG). This transfer function model will be developed based purely on the gasifier responses to step changes in gasifier inputs (including reactor air, convey air, cone nitrogen, FBG pressure, and coal feedrate). This transfer function model will represent a linear, dynamic model that is valid near the operating point at which the data was taken. In addition, a similar transfer function model will be developed using MGAS in order to assess MGAS for use as a model of the FBG for control systems analysis.

II. Discussion of FBG Data

The data for which the transfer function model is developed is taken from gasifier run #10 (October 1994) only. During the previous gasifier run (run #9), the gasifier was operated over a fairly wide range of operating conditions in an attempt to seek an optimal set of operating conditions. A 'good' condition was identified during run #9. That condition was used as the baseline operating point for run #10 (see Table 1 below).

Coal Type	Montana #7
Coal Feed rate	70 lb/hr
Reactor Air flow	1000 scfh
Convey Air	1600 scfh
Steam flow rate	55 lb/hr
Cone Nitrogen flow	100 scfh
Nitrogen Underflow	300 scfh
Operating Pressure	425 psi

Table 1: FBG Run #10 Baseline Operating Condition

The objective of run #10 was to make step changes in the cone nitrogen flow, reactor air flow, reactor pressure, steam flow, coal feed rate, and underflow nitrogen flow around this optimal condition.

Gasifier run #10 went smoothly for step changes made in reactor air and cone nitrogen flow. For each, a positive step change followed by a 2X negative step change, and finally a positive step change (back to the original value) were made. The data is reasonably good for these changes in reactor air and cone nitrogen. However, the next scheduled change was reactor pressure which is maintained by a pressure controller (which manipulates the outlet gas flowrate). When a pressure setpoint change was made, it appears that the pressure controller overreacted by closing the valve on the exit stream. This likely had serious consequences on the bed. As a result, the gasifier run was terminated at that point. We therefore report only the part of the transfer function matrix for which data is available from run #10.

Additional data is available from gasifier runs #8 and #9, however, it is unreasonable to develop a linear model over such a wide range of operating conditions. This additional data will be used in later modeling efforts (see Section VI). The additional data for the transfer function model will be gathered during a run in May 1995.

III. Discussion of Methods Used

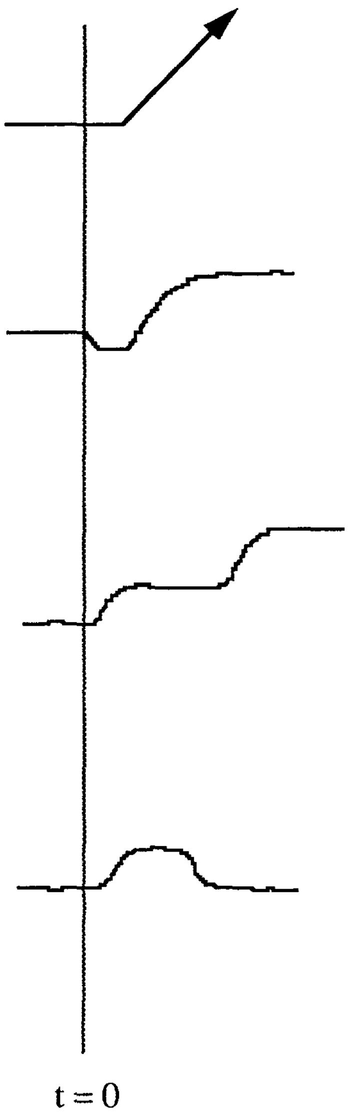
This section will discuss the methodology applied in developing transfer function models from the FBG data. This method is typically used in industry for developing simple control relevant models from process data. It will also be used on simulation data from MGAS to evaluate the applicability of using MGAS for control studies on the FBG.

The method for deriving transfer function models involves two steps: first, pose a reasonable form of the model, and second, evaluate model parameters. Defining a reasonable model form is the more important step. In Figures 1 and 2 below, a number of

Type	Model
1st order Lag	$G(s) = \dfrac{K_p}{\tau_p s + 1}$
1st order Lag + deadtime	$G(s) = \dfrac{K_p e^{-\theta s}}{\tau_p s + 1}$
2nd order underdamped	$G(s) = \dfrac{K_p e^{-\theta s}}{\tau_p^2 s^2 + 2\zeta\tau s + 1}$
2nd order overdamped	$G(s) = \dfrac{K_p e^{-\theta s}}{(\tau_1 s + 1)(\tau_2 s + 1)}$

Figure 1: Some common open loop step responses and their appropriate transfer function models

Response



Type	Model
Pure integrator	$G(s) = \dfrac{K_p e^{-\theta s}}{s}$
Inverse response	$G(s) = \dfrac{-K_1}{\tau_1 s + 1} + \dfrac{K_2 e^{-\theta s}}{\tau_2 s + 1}$ where $K_1 < K_2$
Stiff process	$G(s) = \dfrac{K_1}{\tau_1 s + 1} + \dfrac{K_2 e^{-\theta s}}{\tau_2 s + 1}$ where K_1 and $K_2 > 0$
Zero gain	$G(s) = \dfrac{K_1}{\tau_1 s + 1} + \dfrac{K_2 e^{-\theta s}}{\tau_2 s + 1}$ where K_1 and $K_2 > 0$

Figure 2: Some open loop step responses and their appropriate transfer function models

common 'open loop' step responses are shown along with an appropriate model form for each. 'Open loop' means that there are no automatic control systems on-line.

In Figure 1, the most common model transfer function form used to model plant data is the first order lag plus deadtime (FOLPDT). Complex processes are rarely first order and typically higher order terms are lumped into the deadtime term. For example, a distillation column is comprised of a number of first order systems (column trays) in series resulting in a very high order system. These high order systems are often represented as a FOLPDT. Note that the second order overdamped case can often be modeled reasonably well with a simple FOLPDT. The second order underdamped response is one which can occurs frequently in systems such a RC circuits, along with spring and dashpot systems, but is not all that common in chemical processes. It is theoretically possible for such an open loop response to occur in a reactor system. However, more often than not, such a response is the result of an automatic control system somewhere in the process which is controlling some other process variable.

Figure 2 shows system responses which are more interesting as far as control is concerned. The pure integrator is often seen in tank and accumulator levels in addition to system pressures. Variables which exhibit this type of response can become a problem because they are not self-regulating (they increase without bound). It should also be noted that controlling these variables via automatic control systems can become a problem. If controller gain is set too high or too low, an oscillatory response will result. Since these variables are typically not primary process variables, it is best to control them only within certain bounds rather than controlling them tightly.

The inverse response, stiff process, and zero gain responses are typically the result of competing effects. One effect occurs quickly and the other over a much longer time period. For example, when steam flow is increased to a boiler, the boiler level may actually increase initially due to increased bubbling of the liquid. Over the long run, of course, more liquid will vaporize and the liquid level will drop. The inverse response represents a particularly difficult control problem. If the controller reacts to the initial output response, it will move the manipulated variable in the wrong direction.

Once an appropriate model for has been identified, model parameters are evaluated. Typically, this is accomplished through standard linear or nonlinear regression. Traditional graphical fitting techniques should be used as a quick check of nonlinear regression results, particularly in cases where higher order systems are approximated with a first order lag plus dead time.

IV. Gasifier Data and Transfer Function Models

Figures 3 through 8 plot 10 second process data, and demonstrate that the responses presented above are seen in the operation of the gasifier. It should be noted that these plots are given for illustration only. A number of phenomenological and operational effects must be factored in to their interpretation. Such a discussion is beyond the scope of this progress report.

V. Transfer Function Matrix from Process Data and from MGAS

Tables 2 through 7 present the transfer function matrix derived from FBG process data during run #10 and from MGAS. As previously discussed, this represents only part of the desired transfer function matrix.

A comparison of the Transfer Function models derived using MGAS with those from the FBG data shows that MGAS gives reasonable results in some cases. In many areas, however, it does not. This is especially true in predicting process time constants. As it has been run in these studies so far, MGAS is inadequate for control studies on the FBG. However, further studies will reconfigure MGAS to include a recirculation of solids from top to the bottom and some adjustment of model parameters.

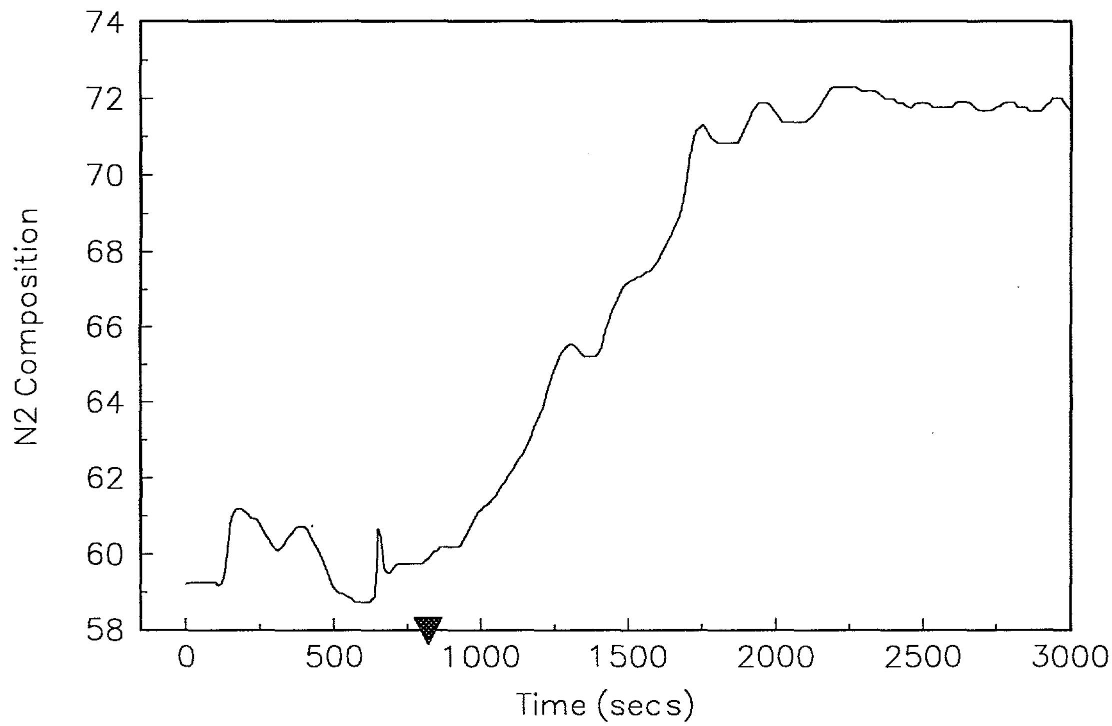


Fig. 3: First Order Response of N2 composition for a change in Cone N2 from 50 to 100 scfh (made at time indicated by ▼).

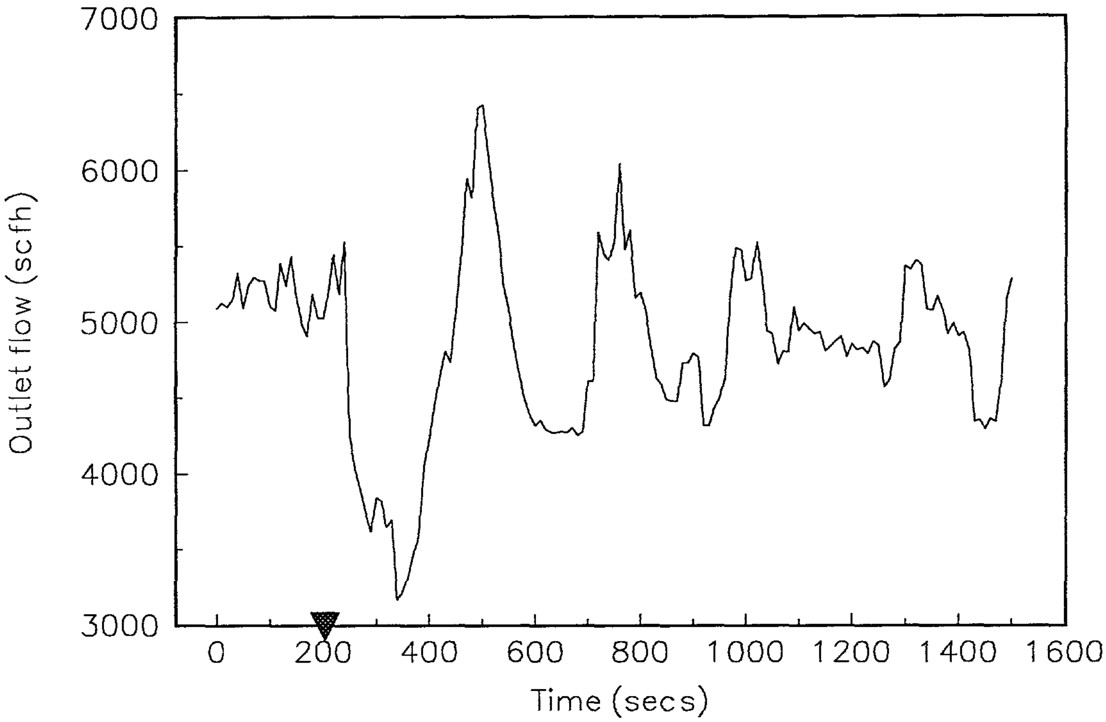


Fig. 4: High order response of Outlet Flow for a change in Reactor Air Flow from 1060 to 940 scfh (made at time indicated by ▼).

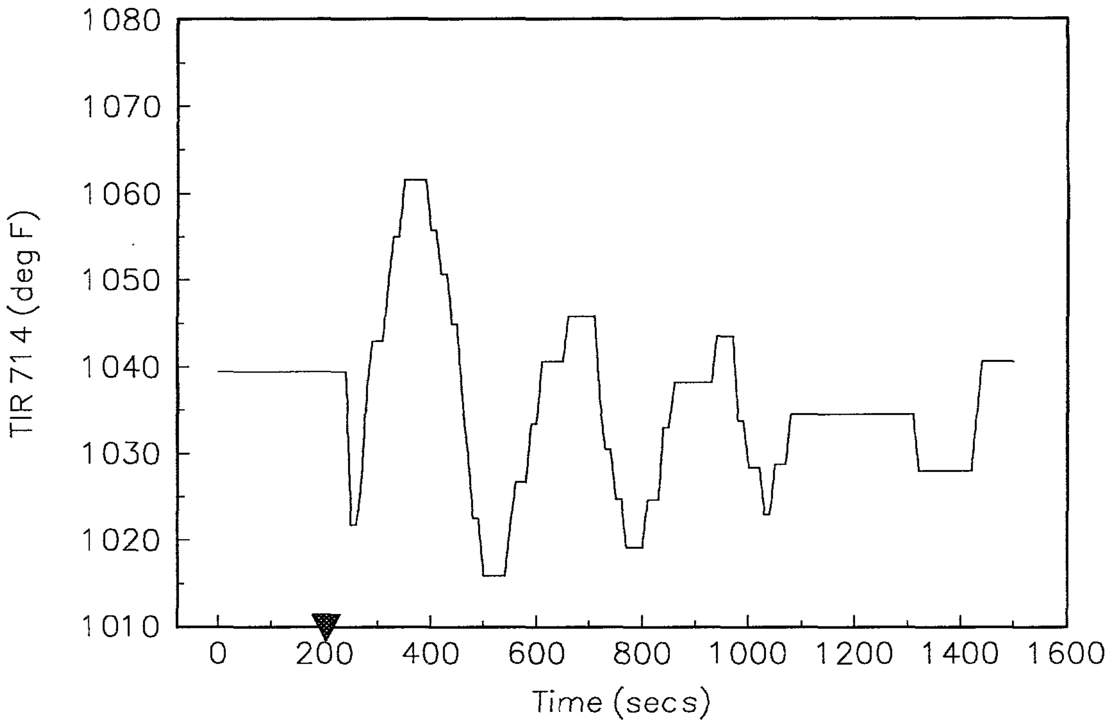


Fig. 5: High Order Response of TIR 714 for a change in Reactor Air Flow from 1060 to 940 scfh (made at time indicated by ▼).

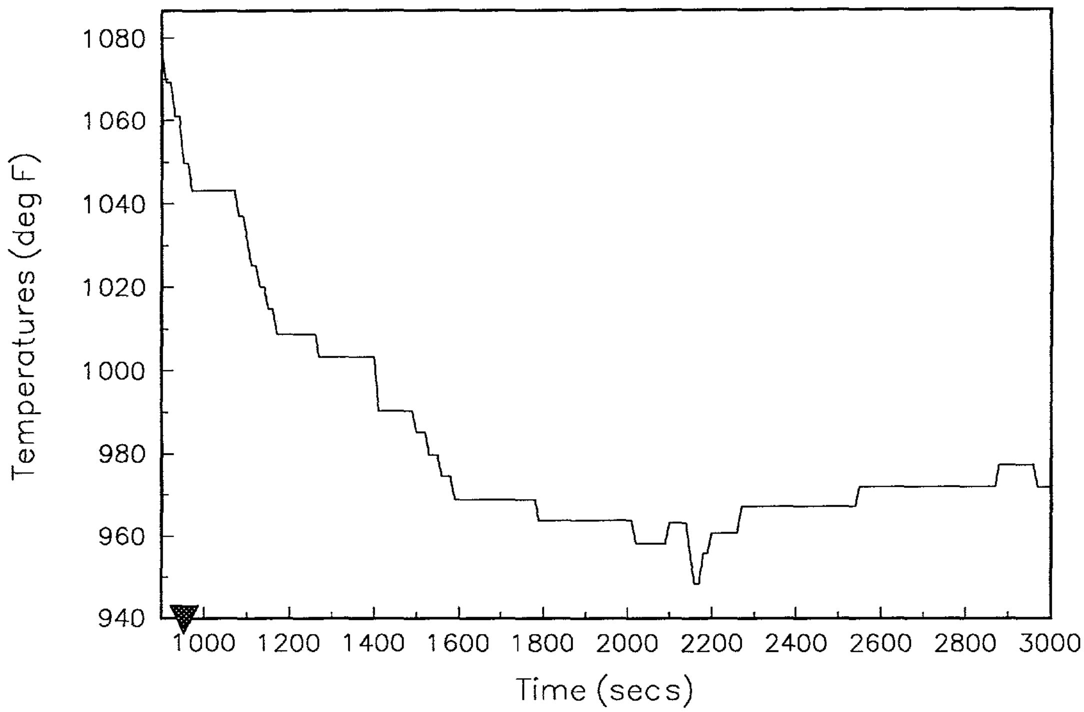


Fig. 6: Response of TIR 700 for a change in Cone N2 from 50 to 100 scfh (made at time indicated by ▼).

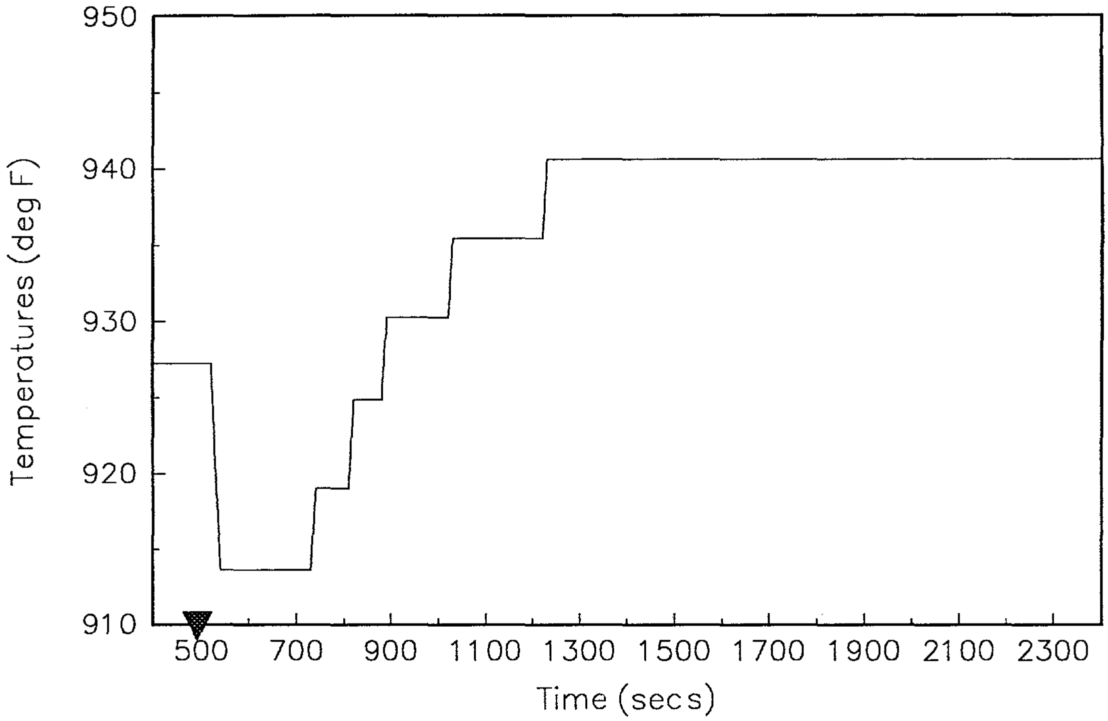


Fig. 7: Inverse Response – TIR 700 for a change in Cone N2 from 150 to 50 scfh (made at time indicated by ▼).

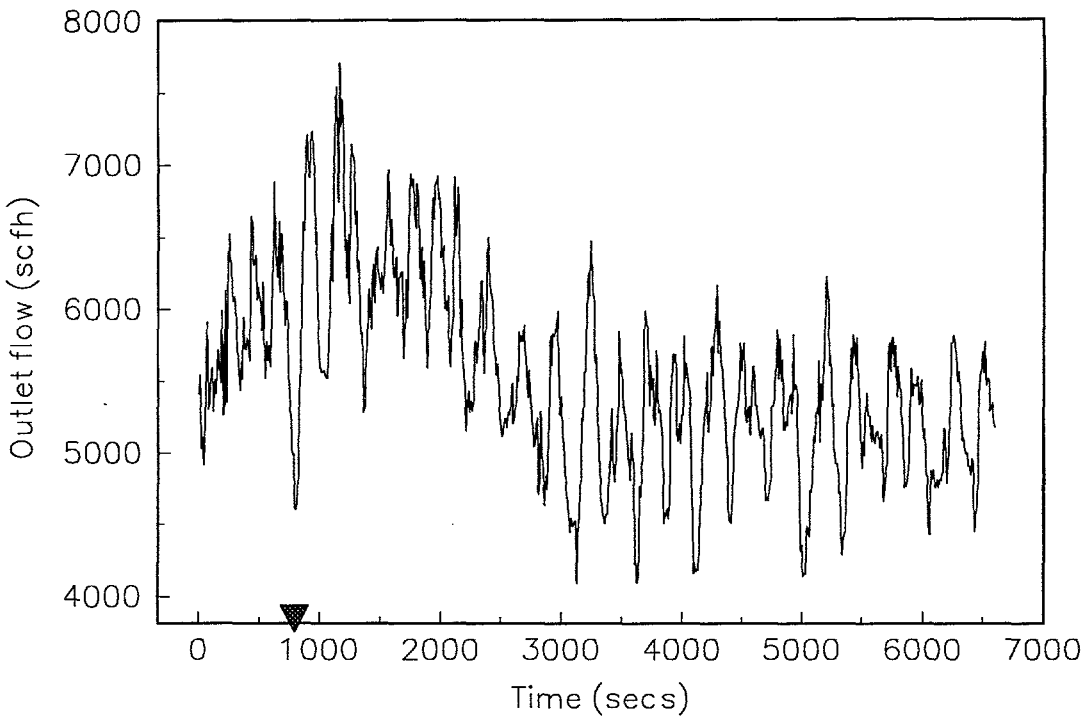


Fig. 8: Noisy Response – Outlet flow rate for a change in Cone N2 from 50 to 100 scfh (made at time indicated by ▼).

Transfer Function: $$\frac{T_i(s)}{F_{N_2}(s)} = \frac{K e^{-\theta s}}{\tau s + 1}$$

	FBG Data			MGAS		
	K	τ	θ	K	τ	θ
TIR 703	-0.0200	500	-	-0.0200	300	-
TIR 702	0.0805	2000	1000	-0.0220	280	-
TIR 707	---------	------	-------	-0.0220	50	-
TIR 701	-0.1051	500	-	-0.0240	75	-
TIR 700	-0.2421	700	-	-0.0231	100	-
TIR 704	-0.0504	600	-	-0.0171	200	-
TIR 705	-0.0298	200	-	-0.0170	120	-
TIR 714	-0.0302	200	-	-0.0160	75	-

Table 2: Process parameters for the response of Reactor Temperatures for a change in Cone Nitrogen from 50 to 100 scfh.

FBG Data

T.F : $\dfrac{P_i(s)}{F_{N_2}(s)} = \dfrac{K e^{-\theta s}}{\tau s + 1}$

PDIR	K	τ	θ
706	-0.5769	400	-
718	-0.2400	400	-
707	--	--	--
708	--	--	--
709	--	--	--
431	--	--	--
710	--	--	--

MGAS

Transfer Function: $\dfrac{P_i(s)}{F_{N_2}(s)} = \dfrac{K_1 e^{-\theta_1 s}}{\tau_1 s + 1} + \dfrac{K_2 e^{-\theta_2 s}}{\tau_2 s + 1}$

PDIR	K_1	τ_1	θ_1	K_2	τ_2	θ_2
706	--	--	--	--	--	--
718	--	--	--	--	--	--
707	0.0501	25	-	0.0672	875	500
708	0.0301	25	-	0.0341	1375	800
709	0.0232	25	-	0.0285	1800	1150
431	0.0217	25	-	0.0226	2000	1250
710	0.0118	25	-	0.0119	2200	1750

Table 3: Process Parameters for the response of Pressure Differentials for a change in Cone Nitrogen from 50 to 100 scfh.

Compositions:

Transfer Function: $$\frac{Y_i(s)}{F_{N_2}(s)} = \frac{K e^{-\theta s}}{\tau s + 1}$$

	FBG Data			MGAS		
	K	τ	θ	K	τ	θ
Y_{CO}	-0.04	700	-	-0.0109	25	-
Y_{CO_2}	0.0	--	--	-0.0187	30	-
Y_{H_2O}	0.0	--	--	-0.0040	175	-
Y_{CH_4}	0.0	--	--	-0.0005	25	-
Y_{H_2}	-0.05	400	-	-0.0012	20	-
Y_{H_2S}	0.0	--	--	-0.00002	30	-
Y_{N_2}	0.12	500	-	0.0331	50	-

Outlet flow:

Transfer Function: $$\frac{F_g(s)}{F_{N_2}(s)} = \frac{K e^{-\theta s}}{\tau s + 1}$$

	FBG Data			MGAS		
	K	τ	θ	K	τ	θ
FGAS	-0.3	1000	-	-0.3	1000	-

Table 4: Process Parameters for the response of Compositions and Outlet Flow for a change in Cone Nitrogen from 50 to 100 scfh.

Transfer Function: $$\frac{T_i(s)}{F_{air}(s)} = \frac{K e^{-\theta s}}{\tau s + 1}$$

	FBG Data			MGAS		
	K	τ	θ	K	τ	θ
TIR 703	-0.0918	25	-	0.1860	275	-
TIR 702	0.1764	50	-	0.2481	175	-
TIR 707	-----	-----	-----	0.1760	30	-
TIR 701	0.1736	150	-	0.2114	60	-
TIR 700	0.2206	150	-	0.2214	100	-
TIR 704	0.2205	75	-	0.1584	225	-
TIR 705	0.2643	100	-	0.1148	175	-
TIR 714	0.3663	125	-	0.0968	125	-

Table 5: Process Parameters for the response of Reactor Temperatures for a change in Reactor Air from 1060 to 940 scfh.

Transfer Function: $$\frac{P_i(s)}{F_{air}(s)} = \frac{K_1 e^{-\theta_1 s}}{\tau_1 s + 1} + \frac{K_2 e^{-\theta_2 s}}{\tau_2 s + 1}$$

	FBG Data						MGAS					
PDIR	K_1	τ_1	θ_1	K_2	τ_2	θ_2	K_1	τ_1	θ_1	K_2	τ_2	θ_2
707	0.1178	50	-	-0.1000	50	20	0.3094	10	-	-0.5890	850	400
708	0.1178	50	-	-0.1200	50	20	0.1913	10	-	-0.4005	1350	750
709	0.2356	50	-	-0.7067	100	50	0.1471	10	-	-0.2944	1750	900
431	-----	---		-----	----	---	0.1177	10	-	-0.3108	2000	1300
710	0.0353	50	-	-0.2356	100	50	0.0442	20	-	-0.0746	2400	1500

Table 6: Process Parameters for the response of Pressure Differentials for a change in Reactor Air from 1060 to 940 scfh.

Compositions:

Transfer Function: $$\frac{Y_i(s)}{F_{air}(s)} = \frac{K\, e^{-\theta s}}{\tau s + 1}$$

	FBG Data			MGAS		
	K	τ	θ	K	τ	θ
Y_{CO}	0.0873	75	-	0.0309	75	-
Y_{CO_2}	-0.0087	50	-	0.0175	400	-
Y_{H_2O}	------	---	---	-0.0530	75	-
Y_{CH_4}	------	---	---	-0.0018	75	-
Y_{H_2}	-0.0407	100	-	-----	----	---
Y_{H_2S}	------	---	---	-----	----	---
Y_{N_2}	0.0707	300	-	0.0213	25	-

Outlet flow:

Transfer Function: $$\frac{F_g(s)}{F_{air}(s)} = \frac{K_1\, e^{-\theta_1 s}}{\tau_1 s + 1} + \frac{K_2\, e^{-\theta_2 s}}{\tau_2 s + 1}$$

	FBG Data						MGAS					
	K_1	τ_1	θ_1	K_2	τ_2	θ_2	K_1	τ_1	θ_1	K_2	τ_2	θ_2
FGAS	3.356	25	-	-3.418	200	75	0.027	10	-	-0.014	200	25

Table 7: Process Parameters for the response of Compositions and Outlet Flow for a change in Reactor Air from 50 to 100 scfh.

VI. Plan of Action

This is a rough updated plan of action for modeling and control of the METC FBG. This plan outlines some of the issues that were discussed during USC's visit to METC on 3/13/95 and suggests actions to be taken to address them. This plan is consistent with the original scope of work in the contract.

1. In this report, we have presented some selected responses meant to show that the gasifier exhibits behavior that is challenging from a control point of view. We have discussed many of these responses with the FBG operations experts at METC to interpret these results. These discussions were very beneficial from our point of view, and will be factored into later versions of the FBG model.

   We will therefore meet in a small group (comprised of S. Noel, J. Rocky, the engineers and technicians responsible for the FBG, and USC) on a more frequent basis and prior to presenting results in a formal seminar at METC.

2. It is possible that the primary cause of premature shutdown during run #10 was due to a poorly tuned pressure controller which manipulates the exit gas flow. It appears that the controller was overreacting to small changes (less than 2 psi) in the gasifier pressure. There is also some uncertainty as to how the pressure control scheme is configured since there are two valves in the loop. It was suggested that a split-range controller may be what is employed.

   We will examine all of the data during (pressure, exit flow, inlet flows, temps) the time period of interest to confirm that the controller was indeed the problem.

The operation of the present pressure control system must be determined (by METC). Once we know what we are dealing with, a general analysis of the control strategy will be made at USC with suggestions for improvements

The control valve(s) manipulating the exit stream should be checked for proper operation. If valves are not working properly, no amount of controller retuning will solve the problem. Once we are certain that there are no hardware problems, the controller can be retuned. This should be done on-line under gasification conditions. A trial retuning should be made during cold start to determine that the controller is acting as expected. Alternatively, one could put the pressure controller in 'manual'. However, this will pose other problems for those actually running the gasifier.

In addition, we will supply references on applied controller tuning and on split-range controllers. We will also send PICLES, a simple controller tuning simulator which will run on a PC.

3. The data from gasifier runs 8, 9, and 10 can all be used to develop a simple gasifier model. The main problem in using all of this data is that the data is spread over a wide range of operating conditions. The initial control modeling plan was to develop a linear model based on small perturbations from a single operating condition (during run #10). A linear model is generally valid only near the operating conditions for which it is developed.

   We will examine the extent of nonlinear behavior exhibited by the gasifier (using data from runs 8, 9, and 10). If it is nonlinear as expected, we can train a neural network model from the steady state data. This model can be used to examine the control at a given operating point and also to find an optimal operating condition (within the envelope of conditions in the process data, i.e.- it won't extrapolate). The accuracy of the neural network model will depend upon the richness of the data from runs 8,9 and 10.

Note that neural network modeling was part of our contract already. This path will be pursued in parallel with the linear transfer function modeling presently underway.

4. Larry Lawson is putting together a control relevant model of PYGAS using TUTSIM. We would like to stay updated on that work as it appears to be the beginnings of a useful model for control purposes. We would even like to obtain a copy of the model at various stages in its development.

5. As for the present Transfer Function modeling:
   S. Reddy will check the present model (there appeared to be some inconsistencies). Appropriate data from the May gasifier run will be added to this model. We will load and run the GQjet spreadsheet model and compare the gains with the transfer function model and also the neural network model. He will also continue with MGAS, adjusting some model parameters and adding a recirculation loop in an attempt to obtain better agreement with the data. Of particular interest is the large discrepancy between the actual time constant and that predicted by MGAS.

6. The success of the modeling and control studies depends upon coupling the process data with the expertise and experience of those running the gasifier. The process data does not always tell the real story. So much is going on during the gasifier run that an important event may be completely missed by simply looking at the sensor data.

We will look more closely at the daily log sheets. More importantly, we will keep in contact (on a weekly basis) with the FBG group. Modeling results will be presented more frequently. This should promote a more frequent exchange of information and ideas.