DOE/MC/31384 -- 5279
(DE96011381)

Analysis And Control Of The METC Fluid Bed Gasifier

Quarterly Report
April - June 1995

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Work Performed Under Contract No.: DE-FG21-94MC31384

For
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
Morgantown, West Virginia

By
University of South Carolina
Office of Sponsored Programs and Research
Columbia, South Carolina 29208

Disclaimer

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I. Overview of Present Work

This document summarizes work performed for the period 4/1/95 to 7/31/95 on contract no. DE-FG21-94MC31384 (Work accomplished during the period 10/1/94 to 3/31/95 was summarized in the previous technical progress report included in the appendix of this report). In this work, three components will form the basis for design of a control scheme for the Fluidized Bed Gasifier (FBG) at METC: 1) a control systems analysis based on simple linear models derived from process data, 2) review of the literature on fluid bed gasifier operation and control, and 3) understanding of present FBG operation and real world considerations.

Tasks accomplished during the present reporting period include: 1) Completion of a literature survey on Fluid Bed Gasifier control, 2) Observation of the FBG during the week of July 17 to July 21, and 3) Suggested improvements to the control of FBG backpressure and MGCR pressure.

II. Gasifier Run 11

Table 1 below summarizes the steady-state operating conditions for FBG run 11.

Coal Type	Montana #7
Coal Feed rate	70 lb/hr
Reactor Air flow	1025 scfh
Convey Air	1600 scfh
Steam flow rate	52 lb/hr
Cone Nitrogen flow	0 scfh
Cone Steam	9 lb/hr
Nitrogen Underflow	300 scfh
Operating Pressure	425 psi

Table 1: FBG Run #10 Baseline Operating Condition

Note that during the initial part of the run, 50 scfh of cone Nitrogen was fed instead of cone steam. The switch to cone steam was made part way through the run.

Table 2 summarizes the run 11 planned tests, and Tables 3 and 4 give the tests that were made during the run. Note that the run covered two time periods, from July 16 to 22, and July 24 to August 8. The data from run 11 includes changes in reactor air, coal feed rate, underflow N_2, reactor steam, switch from cone steam to cone Nitrogen, and switch from Montana #7 coal to coke breeze.

Test Period (TP#)	Hrs.	Cum. Time	Start Time	End Time	Coal Type	Coal Feed Rate	Reactor Air	Convey Air	Reactor Steam	Cone Steam	Cone N2	UnderFlow N2	Reactor Pressure	Test Parameters
						lb/h	scfh	scfh	lb/h	lb/h	scfh	scfh	psig	
	12	12				0	5000	0	0	0	50	350	425	Startup
	8	20			Montant #7	70	525-825	1600	52	0	50	350	425	Fill bed
1+++	12	32			#7	70	1025	1600	52	0	50	300	425	Stabilization
2*	12	44			#7	70	1085	1600	52	0	50	300	425	Reactor
3*	12	56			#7	70	1025	1600	52	0	50	300	425	Air
4	12	68			#7	70	1025	1600	52	[illegible]	0	300	425	Cone N2 to
5++	12	80			#7	70	1025	1600	[illegible]	[illegible]	0	300	425	Cone Steam
6	12	92			#7	70	1085	1600	45	7	0	300	425	
7***	6	98			#7	70	965	1600	45	7	0	300	425	Reactor Air
8	12	110			#7	70	1025	1600	45	7	0	300	425	
9	12	122			#7	80	1025	1600	45	7	0	300	425	Coal
10	12	134			#7	60	1025	1600	45	7	.0	300	425	Feed
11	12	146			#7	70	1025	1600	45	7	0	300	425	Rate
12	12	158			#7	70	1025	1600	45	7	0	250	425	Underflow
13	12	170			#7	70	1025	1600	45	7	0	300	425	Nitrogen
14	12	182			#7	70	1025	1600	45	7	0	300	[illegible]	Rx Press.**
	10	192			Change Feed	0	1025 - 0 N2	1600 - 100 N2	0	0	100	300	425-0	Hot- Hold
15	10	202			Coke Breeze	70	825-1025	1600	45	7	0	500	425	Coke Breeze
16	8	210			Illinois #6+	70	1200	1600	45	7	0	500	425	Bituminous
17	22	232			Blacksville	70	1200	1600	45	7	0	500	425	Coals
	8	240				0	4200-7000 N2	800 N2	0	0	100	2000	425-0	Cool Down

+++ Old Baseline (94FBG10). ++ New Baseline (95FBG11) *** Will not perform if time is off schedule. + May be replaced by Fort-Martin (Martiki) coal if available.
** Put PCV-713 on "Manual" mode. * Sample UF & OF each per hour during the 2nd half of TP#2 and the entire TP#3; total of 36 samples between TP#2 and 3.

Table 2: Test Matrix Planned for 95FBG11. (July 16 to 26, 1995)

Test Period	Hrs.	Cum. Time***	Start Time	End Time	Coal Type	Coal Feed Rate	Reactor Air	Convey Air	Reactor Steam	Cone Steam	Cone N2	UnderFlow N2	Reactor Pressure	Test Parameters
(TP#)	.					lb/h	scfh	scfh	lb/h	lb/h	scfh	scfh	psig	
	16	16	7/16/95 20:00	7/17/95 11:56		0	5000	0	0	0	50	350	425	Startup
	1.25	17.25	7/17/95 11:56	7/17/95 13:13	Montant #7	30.6	2300+	800-1600	0	0	55	350	425	Combustion/Fill Bed
	5.25	22.5	7/17/95 13:13	7/17/95 18:30	#7	70	525-825	1600	56	0	55	350	425	Stabilization
1a	4	26.5	7/17/95 18:30	7/17/95 22:21	#7	70	1025	1600	56	0	50	300	425	Baseline**
	6	32.5	7/17/95 22:21	7/18/95 4:30	Lost of coal feed due to feeder plug by foreign material in feed									Hot-Hold
1b	9	41.5	7/18/95 4:30	7/18/95 13:30	#7	70	1025	1600	52	0	50	300	425	Baseline*
2#	12	53.5	7/18/95 13:30	7/19/95 1:30	#7	70	1085	1600	52	0	50	300	425	Reactor Air
3#	12	65.5	7/19/95 1:30	7/19/95 13:30	#7	70	1025	1600	52	0	50	300	425	
4	13.25	78.75	7/19/95 13:30	7/20/95 2:45	#7	70	1025	1600	[illegible]	[illegible]	0	300	425	Cone N2 to Steam
5	6.75	85.5	7/20/95 2:45	7/20/95 9:32	#7	70	1025	1600	52	16	0	250	425	Baseline**
6	16	101.5	7/20/95 9:32	7/21/95 1:30	#7	70	1025	1600	66	[illegible]	0	250	425	Rx & Cone Steam
7	12.3	113.8	7/21/95 1:30	7/21/95 13:49	#7	80	1025	1600	66	9	0	250	425	Coal Feed Rate
8	12	125.8	7/21/95 13:49	7/22/95 1:40	#7	70	1025	1600	66	9	0	250	425	
9	12.2	138	7/22/95 1:40	7/22/95 13:52	#7	60	1025	1600	66	9	0	250	425	
	6	144	7/22/95 13:52	7/22/95 20:00		0	4200/7000 N2	800 N2	0	0	100	2000	425-0	Cool Down

* Old Baseline (94FBG10). ** New Baseline (95FBG11) *** Total Gasification Time = 115 hours

\# Sample UF & OF each per hour during the 2nd half of TP#2 and the entire TP#3; total of 19 samples between TP#2 and 3.

Table 3: Test Matrix Completed for 95FBG11 (July 16 to 22, 1995)

Test Period	Hrs.	Cum. Time***	Start Time	End Time	Coal Type	Coal Feed Rate	Reactor Air	Convey Air	Reactor Steam	Cone Steam	Cone N2	Underflow N2	Reactor Pressure	Test Parameters
						lb/h	scfh	scfh	lb/h	lb/h	scfh	scfh	psig	
	22	22	7/24/95 20:00	7/25/95 17:53		0	5000	0	0	0	50	350	425	Startup
	1	23	7/25/95 17:53	7/25/95 18:53	Montana #7	30.6	2300+	800-1600	0	0	55	350	425	Combustion
	1	24	7/25/95 18:53	7/25/95 20:00	#7	70	525-825	1600	66	0	55	350	425	Fill bed
	5.75	29.75	7/25/95 20:00	7/26/95 1:45	#7	70	1025	1600	66->52	0	50	300	425	Stabilization
1	5.5	35.25	7/26/95 1:45	7/26/95 7:15	#7	70	1025	1600	52	0	50	250	425	Baseline*
	12	47.25	7/26/95 7:15	7/26/95 19:13	Shut down due to valve malfunction									Shut Down
	1	48.25	7/26/95 19:13	7/26/95 20:15	#7	30.6	1025	1600	0	0	55	250	425	Combustion
		48.25	7/26/95 20:15		#7	30.6->70	500->1025	1600	60	0	55	250	425	Fill Bed
	5.25	53.5		7/27/95 1:30	#7	70	1025	1600	60->52	0	50	250	425	Stabilization
2	12	65.5	7/27/95 1:30	7/27/95 13:30	#7	70	1025	1600	52	0	50	250	425	Baseline**
3	12	77.5	7/27/95 13:30	7/28/95 1:30	#7	70	1025	1600	52	0	50	[illegible]	425	Underflow Nitrogen
4	12.75	90.25	7/28/95 1:30	7/28/95 14:17	#7	70	1025	1600	52	0	50	280	425	
5	11.3	101.55	7/28/95 14:17	7/29/95 1:30	#7	70	1025	1600	52	0	50	250	425	
6	12.43	113.98	7/29/95 1:30	7/29/95 13:50	#7	70	1025	1600	[illegible]	0	50	250	425	Reactor Steam
7	2.5	116.48	7/29/95 13:50	7/29/95 16:20	#7	70	1025	1600	[illegible]	0	50	250	425	
	3.75	120.23	7/29/95 16:20	7/29/95 20:00	Fixed rupture disc on secondary cyclone									Hot-Hold
	5.5	125.73	7/29/95 20:00	7/30/95 1:30	#7	70	825->1025	1600	52	0	50	250	425	Fill Bed
	13.25	138.98	7/30/95 1:30	7/30/95 14:42	#7	70	1025	1600	52	0	50	250	425	Stabilization
8	10.75	149.73	7/30/95 14:42	7/31/95 1:30	#7	70	1025	1600	57	0	50	250	425	Conc Steam
9	12	161.73	7/31/95 1:30	7/31/95 13:30	#7	70	1025->1400	1600	57	0	50	250	425	
10	12	173.73	7/31/95 13:30	8/1/95 1:30	#7	80	1400	1600	57	0	50	250	425	Coal Feed Rate
11	9	182.73	8/1/95 1:30	8/1/95 10:30	#7	70	1025	1600	52	0	50	250	425	Baseline
12	75.42	258.15	8/1/95 10:30	8/4/95 13:49	#7	70	1085	1600	52	0	50	250	425	Reactor Air
	11	269.15	8/4/95 13:49	8/5/95 0:47	Allowing MGCR to load the sorbent									Hot-Hold
13	4.3	273.45	8/5/95 0:47	8/5/95 5:10	#7	70	1085	1600	52	0	50	250	425	Baseline
	14.3	287.75	8/5/95 5:10	8/5/95 20:00	Loss of coal feed due to clinker									Shut Down
	3.85	291.6	8/5/95 20:00	8/5/95 23:50		0	4000	0	0	0	0	250	425	Heat Up
	1.75	293.35	8/5/95 23:50	8/6/95 1:35	#7	12	4000	800	0	0	0	250	425	Combustion/ Fill Bed
	4	297.35	8/6/95 1:35	8/6/95 5:30	#7	12->70	500->1085	1600	52	0	0	250	425	Stabilization
14	16.5	313.85	8/6/95 5:30	8/6/95 22:00	#7	70	1085	1600	52	0	0	250	425	Sorbent Test
15	12	325.35	8/6/95 22:00	8/7/95 10:00	Coke Breeze	70	1085	1600	55	0	0	400	425	Coke Breeze
	8.83	334.58	8/7/95 10:00	8/7/95 18:50	Loss of Rx Air and Steam due to Clinker									Shut Down
	1.44	336.12	8/7/95 18:50	8/7/95 20:15	Coke Breeze	70	500->1085	1600	0	0	0	400	425	Combustion/ Fill Bed
16	5.4	341.52	8/7/95 20:15	8/8/95 1:35	Coke Breeze	70	1085	1600	55	0	0	400	425	Stabilization
17	0.1	341.62	8/8/95 1:35	8/8/95 1:40	Blacksville Coal	70	1085	1600	55	0	0	400	425	Bituminous Coal
	18.38	360	8/8/95 1:40	8/8/95 20:00	Cool Down									Cool Down/ Shut Down

* Old Baseline (94FBG10). ** New Baseline (95FBG12) *** Total Gasification Time = 261 hours % availability = 82% (319 hrs = 100%)

Table 4: Test Matrix Completed for 95FBG11 (July 24 to Aug 8, 1995)

Gasifier Operation and Control:

In operating and controlling the FBG a number of objectives must be met. They are summarized below:

1. No clinkering
2. High carbon conversion
3. Meet targeted gas make
4. Meet targeted bed density
5. High gas heating value
6. Meet targeted Fuel/noncombustible mole ratio
7. Meet targeted mean bed temperature
8. Maintain HOC balances and inventories.

Presently all of these objectives and more are considered by operators during gasifier operation. All inlet gas flow rates are flow controlled with simple PID-type controllers, gasifier backpressure is controlled via a split range controller, and MGCR pressure is controlled via a PID controller. The backpressure control is critical to steady operation of the gasifier, as fluctuations in backpressure impact inlet gas flowrates and bed density. More detail on backpressure and MGCR control is given in the next section. Typically the maximum bed temperature is maintained by adjusting the air flow setpoint, gas moisture content is maintained at 10% by adjusting the steam flow setpoint.

III. Backpressure and MGCR Control

Good pressure control is critical to successful operation of the FBG. Fluctuations in gasifier pressure affect inlet gas flowrates, gasifier temperatures, and downstream MGCR pressure. Over the last several gasifier runs, the FBG backpressure has been controlled using a split-range automatic controller. Most of the time this controller maintains the pressure within plus or minus 5 psi of setpoint. However, frequently the controller overreacts and the pressure swings dramatically. If the operator does not take the proper intervention steps immediately, the pressure swings will ultimately shut down the gasifier.

This section identifies several sources of problems with the present pressure control system and then suggests modifications to the present scheme.

i. Problems with the present control scheme

Below are summarized some of the major problems with the present backpressure controller.

1. *Split-range control scheme*: A large valve and a small valve operating in parallel are manipulated in order maintain desired FBG pressure. The small valve opens first to control pressure at low to moderate make-gas flowrates, while the large valve remains closed. At high make-gas flows, the small valve is open completely, and the large valve is manipulated to maintain pressure. At the operating condition used in the first four days of 95FBG11, the make-gas flow was such that the split-range controller operated at the crossover point from the large valve to the small valve (that is, the large valve closed, the small valve open). One can not expect good control in this region.

2. *Interactions with the MGCR pressure controller:* The MGCR pressure fluctuates due to a large dead time between the upstream valve and the vessel pressure (V-100). Fluctuations in the valve controlling the MGCR pressure (PV-254) affect the backpressure controller.

3. *Upstream disturbances:* The inlet gas flow controllers interact with the backpressure. Changes in inlet gas flowrates will affect gasifier pressure. Similarly gasifier pressure will affect inlet flow of gases. Most of the time when the gasifier backpressure cycles, so do the inlet gas flows.

4. *Controller tuning:* Optimal controller tuning parameters will change as the operating condition changes. For example, one would expect markedly different tuning parameters in the backpressure controller under conditions where the large valve is adjusted than under conditions where the small valve is being adjusted. In one observed instance, the backpressure loop was swinging rather dramatically. The operator on duty intervened by simply putting the controller in manual and maintaining a constant valve position. Almost immediately, the backpressure stabilized. This points to poor controller tuning.

5. *Buildup of solids at the control valve:* There is evidence to suggest that fine solids particles are accumulating just upstream of the control valve. In one case, backpressure was oscillating continuously with increasing amplitude. Finally, the pressure swings were large enough to force solids out of the gasifier and into the incinerator (and damaging the incinerator). After this 'burp' gasifier control was very good for a long period of time.

ii. Suggested modifications to backpressure and MGCR pressure controllers.

The following modifications are suggested in order to eliminate backpressure control problems:

A. Backpressure controller

1. Replace the split-range configuration with the following: Two valves placed in parallel (similar to the present configuration). One valve should be tied to a PID controller and will directly control FBG backpressure. This valve should be sized to cover the range of desired operating conditions. A second, larger valve will be used to let down system pressure quickly. This valve can only be manipulated manually or through a safety override. With this configuration, under normal, steady operation of the gasifier, the large valve will remain static and the controller will manipulate the other valve to maintain backpressure.

2. Install a purge system to remove solids accumulation in the exit line.

3. Establish good controller tuning guidelines - how controllers should be tuned and who should tune them. An autotuning facility available in most DCS's should be most useful.

B. MGCR pressure control

1. Implement a cascade control arrangement to reduce the large time lag between valve V-254 and vessel V-100. In a cascade arrangement, an inner controller would control the pressure just downstream of the valve V-254 or in the particulate removal vessel,

F-100. The outer or master controller maintains the pressure in V-100 by adjusting the setpoint of the inner controller. The result is a control system that responds much faster and rejects disturbances in upstream pressure.

C. Diagram of suggested backpressure and MGCR pressure control scheme.

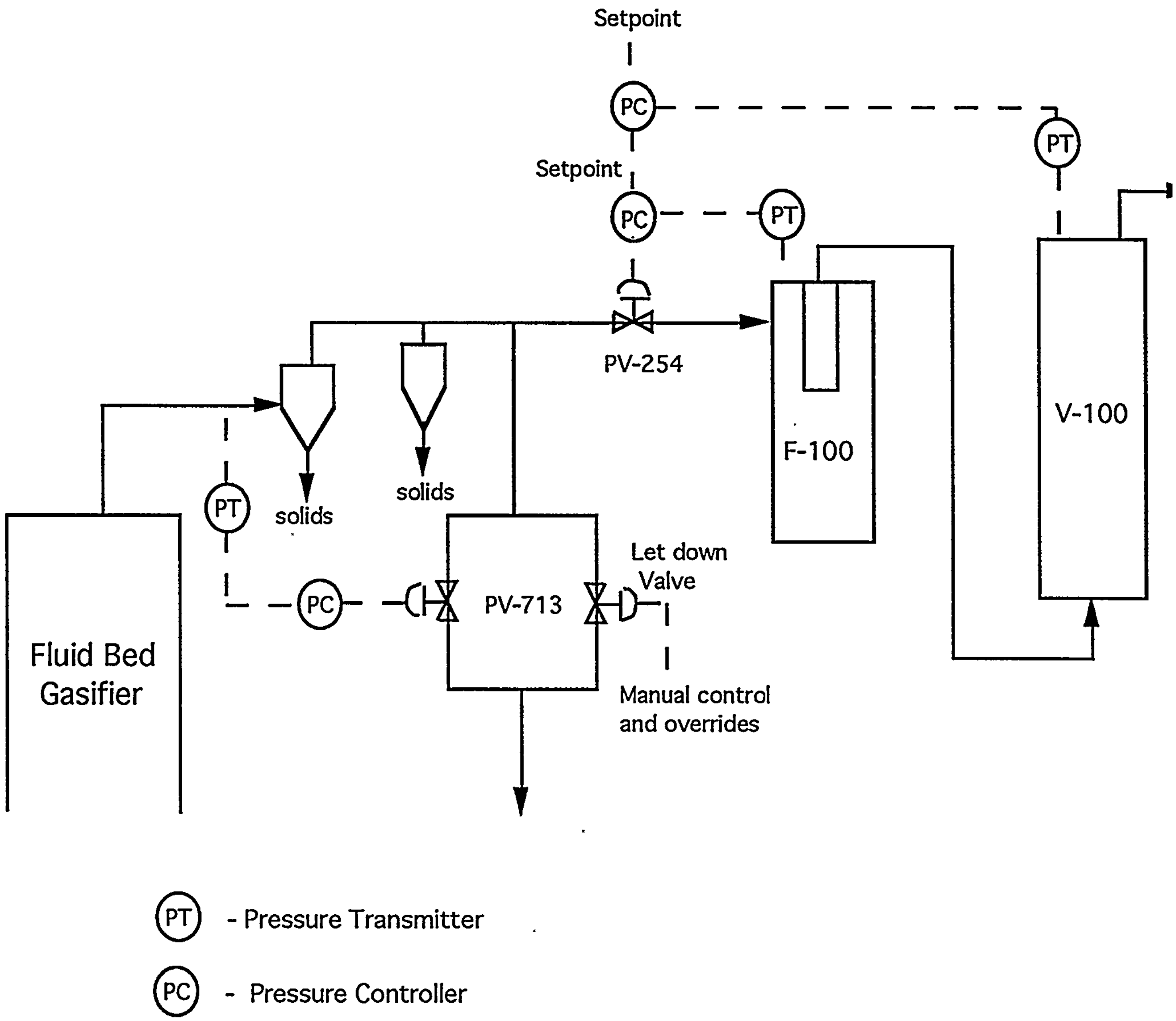

Figure 1 - Suggested FBG backpressure and MGCR pressure control scheme

IV. Plan of Action

This section presents an updated plan of action for the modeling and control effort.

I. Tasks to be Performed

Task 1: Data collection from gasifier operation

FBG data has been collected during FBG run 11. Simple step changes were made in many of the process inputs during these gasifier runs.

Task 2: Development of FBG models

Two models will be developed:

1. A process transfer function matrix will be derived from process data. Process gains, time constants, and deadtimes will be calculated via weighted least squares regression. This transfer function model represents a dynamic, linear model of the gasifier.
2. A backpropagation neural network will be trained with available steady-state data. The model will represent a nonlinear, steady-state model of the gasifier which can be used to predict process outputs such as make gas-flow, gas compositions, maximum bed temps, etc.

Task 3: Steady-state analysis

A steady-state analysis of the FBG will be performed, including:

1. Steady-state gain calculation
2. Relative Gain Analysis
3. Singular value decomposition
4. Neiderlinski index computation

Task 4: Propose improvements to control system design

Improvements to the present control system will be made. These proposed improvements will be made at two levels: 1) designs which include classical single loop PID controllers, feedforward controllers, and ratio controllers (these would include flow controllers, pressure controllers, ratio of flow control, etc.). 2) designs which include advanced multivariable control schemes. These may be linear or nonlinear controllers.

These suggested control designs will be based on 3 elements: 1) Results of the steady-state and dynamic analyses, 2) An understanding of the literature on FBG control (that is, what control schemes have been implement of FBGs and what was their performance?), 3) Observation and understanding of the operation of the METC FBG. Specifically, the designs will address gasifier pressure control (particularly back-pressure control and MGCR pressure control), inlet feed gas control (these flows are coupled to back-pressure), bed temperature control, gas composition control, and coal feedrate control. Many of these suggested improvements (eg. gasifier back-pressure control scheme) will be documented and presented to METC as they are generated (rather than delay their presentation to a final report).

Task 5: Dynamic analysis of the proposed control schemes

The control schemes will be tested dynamically using the transfer function model generated from the process data under task 2. It should be pointed out that the model generated is limited to the range of conditions under which the gasifier data was collected, and that the transfer function model is linear. However, this dynamic analysis is useful in comparing the relative advantages of any proposed control scheme over one presently being used, or to assess the potential gains in using advanced, multivariable control.

Task 6: Provide consultation to METC through throughout implementation of the improved control scheme.

This task includes:

1. Continued updating of the models as FBG data becomes available.
2. Dynamic testing of suggested changes to the control scheme.
3. Visits to METC to discuss topics related to control of the FBG.
4. Consultation during any phase of the control system implementation on the FBG.
5. Travel to METC during the first gasifier runs under the revised control scheme.

II. Tentative Time Table

Task 1: Data collection from FBG	8/15/95
Task 2: Development of FBG models	
a. Transfer function model (updated)	11/15/95
b. Neural network model	11/30/95
Task 3: Steady-state analysis	12/31/95
Task 4: Proposed improvement to control system	
a. Back-pressure and MGCR pres control	8/20/95
b. others - Flow, bed temp, etc.	as available
Task 5: Dynamic analysis of control schemes	2/28/96
Task 6: Consultation on FBG control	as needed

Appendix: Past Progress Report

ANALYSIS AND CONTROL OF THE METC FLUID BED GASIFIER

Technical Progress Report for the Period 1/1/95 - 3/31/95

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Work performed under Contract No.: DE-FG21-94MC31384

For:
US Department of Energy
Morgantown Energy Technology Center
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Table of Contents

I. Overview of Present Work

This document summarizes work performed for the period 10/1/94 to 3/31/95. In this work, three components will form the basis for design of a control scheme for the Fluidized Bed Gasifier (FBG) at METC: 1) a control systems analysis based on simple linear models derived from process data, 2) review of the literature on fluid bed gasifier operation and control, and 3) understanding of present FBG operation and real world considerations. Below we summarize work accomplished to date in each of these areas.

The initial phase of the work focused on developing simple gain matrix and transfer function models of the Fluidized Bed Gasifier (FBG). These models were 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). The transfer function model represents a linear, dynamic model that is valid near the operating point at which data was taken. In addition, a similar transfer function model has been developed using MGAS in order to assess MGAS for use as a model of the FBG for control systems analysis. A steady state gain matrix has also been derived from the GQ Jet spreadsheet model.

The literature on FBG operation and control is rather sparse. However, we have uncovered several articles which should be valuable. This documentation is limited to academic pilot and laboratory scale FBG's. Unfortunately, industrial documentation of FBG (by Shell, Exxon, etc.) is difficult to find. However, the work by Felder at NC State, Fan at Kansas State, and Uemaki in Japan should serve as good starting points.

Both the control systems analysis and an understanding of previous FBG work are extremely important. Just as important are 'real world' considerations. The METC gasifier has its own unique configuration, and will have its own set of operating procedures, limits, and constraints. Understanding the details of how the operators presently run the FBG is critical to designing a safe and effective control strategy.

II. Transfer Function and Steady-State Gain Calculations

The data for which the transfer function model is developed has been 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.

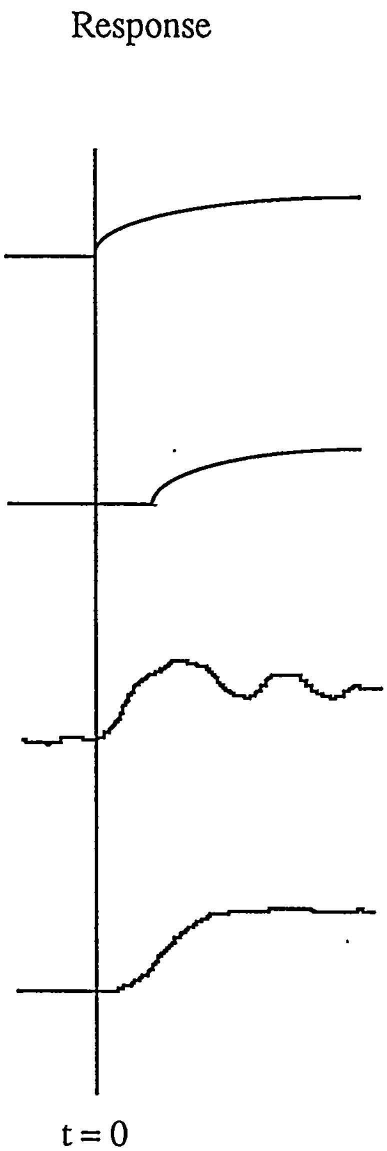
Additional data is available from gasifier runs #8 and #9, however, it is unreasonable to develop a linear model over such a wide range of operating conditions. This additional data will be used in later modeling efforts (see Section VI). The additional data for the transfer function model will be gathered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

IIa. 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.

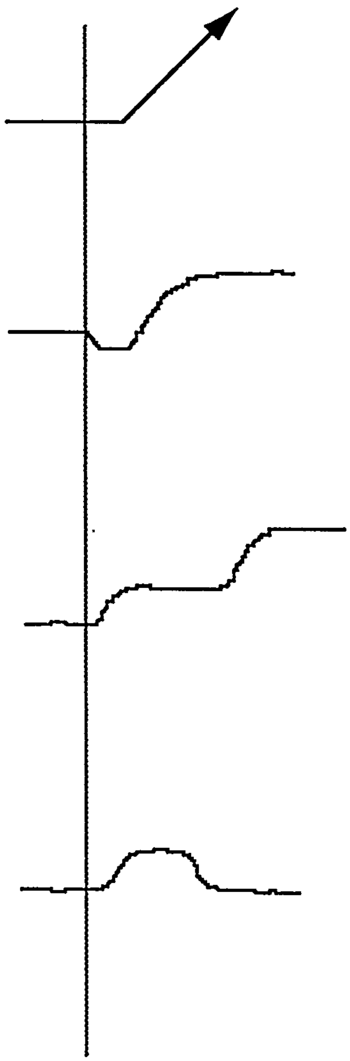
The method for deriving transfer function models involves two steps: first, pose a reasonable form of the model, and second, evaluate model parameters. Defining a reasonable model form is the more important step. In Figures 1 and 2 below, a number of common 'open loop' step responses are shown along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



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 $K_2 > 0$

$t = 0$

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

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.

IIb. Transfer Function Matrix from Process Data and MGAS, Gain Matrix derived from GQ Jet Model

As an initial basis of comparison between the FBG data, and the MGAS and GQ Jet models, Tables 2 and 3 compare outlet gas compositions after cone nitrogen and reactor air have been changed. The actual gas exiting from the FBG is comprised mainly of carbon monoxide (10%), carbon dioxide (10%), nitrogen (60%), and hydrogen (20%). MGAS predicts only carbon monoxide, carbon dioxide and nitrogen in the exit gas. The GQ Jet model predicts some hydrogen (6%) in the product gas, but under predicts the carbon monoxide composition.

Steady-state gains were calculated for important process variables using FBG data, MGAS, and GQ Jet. These results are summarized by Tables 4 and 5. There is some agreement in a few of the gains, however, for the most part gains computed using the models do not consistently match those calculated from the FBG data.

Tables 6 through 11 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.

III. FBG Control in the Literature

Below we summarize the relevant literature relevant to FBG control. All of these articles are based on academic studies made on pilot or lab scale units. Documentation on industrial processes has not been found to this point. The articles mentioned below have been attached in the appendix of this report.

Work at North Carolina State [1][2][3] in the mid 1980's centered on a pilot scale FBG. This reactor was a 15 cm-id stainless steel pipe enclosed in several layers of Fiberfrax bulk ceramic insulation. Steam and oxygen were preheated to 800 K and injected into the bottom of the bed. The jet penetration was estimated to be approximately 10 cm into the bed. The reactor pressure was maintained at 100 psi, typical feedrates of coal, steam, and oxygen were 55 lb/hr, 58 lb/hr, and 14 lb/hr respectively. In their work, NC State develop a working dynamic model of the FBG, studied the effects of process inputs (coal feed, oxygen, steam, reactor pressure) on process outputs (gas composition, average bed temp, etc), examined the dynamics of the process, and looked at FBG control.

L.T. Fan and coworkers at Kansas State University [4] studied a bench scale fluidized bed reactor for gasification of coal with steam as the fluidizing medium. They also developed a mathematical model of their system which could prove useful for scaleup. Their system also contained a mixture of sand and limestone as bed material to prevent agglomeration.

A number of researchers have performed work on spouted bed coal gasifiers [5][6][7][8][9]. In a two stage fluidized spouted bed gasifier, Tsuji and Uemaki examined the effects of oxygen/coal ratio, steam/coal ratio, and coal feed rate on process outputs (gas composition, carbon conversion, maximum bed temperature). Although this work was performed on a spouted bed, the results seem to be consistent with the work at NC State.

We have also included three articles [10][11][12] relevant to the monitoring and control of fluidized bed reactors. Although the fluidized beds here are not specifically coal gasifiers, the techniques are applicable. The work of MacGregor at McMaster University is particularly useful for monitoring a process that is multivariable in nature. His multivariable statistical plots could be a very useful tool on the FBG.

IV. Plan of Action

This is an updated plan of action for modeling and control of the METC FBG. We outline the tasks to be performed in each of the three components of our study. Note that the upcoming gasifier run (July 1995) is critical to the success of this project. Once the tasks in each area have been completed, we will propose a cohesive control system design for the METC FBG. This plan is consistent with the original scope of work in the contract.

I. Modeling and Control Analysis

1. Obtain a complete transfer function model of the FBG from process data. At present, we only have data for changes in nitrogen underflow and reactor air. This task will be completed with data from the July 1995 gasifier run.
2. Perform a control system analysis on the transfer function model developed. RGA, SVD, controllability, observability, robustness indexes, etc. will all be calculated. This information must be interpreted in the context of physical constraints.
3. The data from gasifier runs 8 - 11 will be used to develop a simple neural network model of the gasifier. This neural network will be trained with steady-state data gathered during these gasifier runs. The model will therefore be steady-state, but it will be nonlinear, and can be used to examine the controllability over a range of conditions. It may also be useful for finding an optimal operation condition for the FBG.

II. Literature Search

1. Obtain theses of Russell Rhinehart and of Mark Purdy from NC State. In these documents, they give detail on their modeling effort and on their analysis of a control scheme for the NC State gasifier.

2. In a few conference proceedings, Exxon and Shell have presented some information on their fluid bed coal gasification units. We will continue to seek documentation on those presentations.

III. Real World Considerations

1. We will observe the actual operation of the METC FBG during the week of July 17 - 21. We will identify operational constraints, undocumented procedures, and, in general, the nuts and bolts of operating the gasifier that models can not consider.

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	FBG	MGAS	GQ Jet
CO	10	17	0.9
CO_2	10	18	16.5
CH_4	0	0	0.9
H_2	20	0	6.0
H_2S	0	0	0.3
N_2	60	65	73.8
$C_2H_6 + C_2H_4$	0	0	1.2
NH_3	0	0	0.5

Table 2. Outlet gas compositions (dry basis, mole %) after a change in cone nitrogen from 50 to 100 scfh.

	FBG	MGAS	GQ Jet
CO	9	15	0.9
CO_2	11	18	16.5
CH_4	2	0	0.9
H_2	18	0	6.1
H_2S	0	0	0.3
N_2	60	67	73.7
$C_2H_6 + C_2H_4$	0	0	1.3
NH_3	0	0	0.5

Table 3. Outlet gas compositions (dry basis, mole %) after a change in reactor air from 1060 to 940 scfh.

Temperatures		FBG	MGAS	GQ Jet
Temperature at pyrolyzer outlet (level 4)	(~ TIR 714)	-0.0302	-0.0160	-0.0127
Temperature at jet outlet (level 2)	(~ TIR 703)	-0.0200	-0.0200	-0.0159
Temperature at jet center (level 3)	(~ TIR 702)	0.0805	-0.0220	-0.0160
Pressure Drops				
Pressure drop in jet	(~ PDIR 707 + 708)	------	0.1583	0.00005
Pressure drop in upper pyrolyzer	(~ PDIR 709 + 431 + 710)	------	0.1050	-0.0050
Gas Composition at pyrolyzer outlet				
CO		-0.0400	-0.0109	-0.0007
CO_2		0.0	-0.0187	-0.0029
CH_4		0.0	-0.0005	-0.0005
H_2		-0.05	-0.0012	-0.0019
H_2S		0.0	-0.00002	0.0
N_2		0.12	0.0331	0.006

Table 4. Process gains for reactor temperature, pressure differentials, and gas compositions for a change in cone nitrogen.

Temperatures		FBG	MGAS	GQ Jet
Temperature at pyrolyzer outlet (level 4)	(~ TIR 714)	0.3663	0.0968	0.5742
Temperature at jet outlet (level 2)	(~ TIR 703)	-0.0918	0.1860	0.3180
Temperature at jet center (level 3)	(~ TIR 702)	0.1764	0.2481	0.3180
Pressure Drops				
Pressure drop in jet	(~ PDIR 707 + 708)	0.1800	-0.4417	-0.0030
Pressure drop in upper pyrolyzer	(~ PDIR 709 + 431 + 710)	-----	-0.3681	-0.1106
Gas Composition at pyrolyzer outlet				
CO		0.0873	0.0309	0.0080
CO_2		-0.0087	0.0175	0.0088
CH_4		-------	-0.0018	0.0018
H_2		-0.0407	0.0	-0.0150
H_2S		0.0	0.0	-0.0009
N_2		0.0707	0.0213	0.0071

Table 5. Process gains in reactor temperatures, pressure differentials, and gas compositions for a change in reactor air.

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 6. Transfer functions for the response of reactor temperatures under a change in cone nitrogen.

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 7. Transfer functions for the response of reactor temperatures under a change in reactor air.

FBG Data				MGAS						
T.F: $\frac{P_i(s)}{F_{N_2}(s)} = \frac{K e^{-\theta s}}{\tau s + 1}$				Transfer Function: $\frac{P_i(s)}{F_{N_2}(s)} = \frac{K_1 e^{-\theta_1 s}}{\tau_1 s + 1} + \frac{K_2 e^{-\theta_2 s}}{\tau_2 s + 1}$						
PDIR	K	τ	θ	PDIR	K_1	τ_1	θ_1	K_2	τ_2	θ_2
706	-0.5769	400	-	706	--	--	--	--	--	--
718	-0.2400	400	-	718	--	--	--	--	--	--
707	--	--	--	707	0.0501	25	-	0.0672	875	500
708	--	--	--	708	0.0301	25	-	0.0341	1375	800
709	--	--	--	709	0.0232	25	-	0.0285	1800	1150
431	--	--	--	431	0.0217	25	-	0.0226	2000	1250
710	--	--	--	710	0.0118	25	-	0.0119	2200	1750

Table 8. Process Parameters for the response of Pressure Differentials for a change in Cone Nitrogen

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 9. Process Parameters for the response of Pressure Differentials for a change in Reactor Air

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 10. Process Parameters for the response of Compositions and Outlet Flow for a change in Cone Nitrogen

Compositions:

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

	FBG Data K	FBG Data τ	FBG Data θ	MGAS K	MGAS τ	MGAS 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 K_1	τ_1	θ_1	K_2	τ_2	θ_2	MGAS 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 11. Process Parameters for the response of Compositions and Outlet Flow for a change in Reactor Air