DOE/MC/31384 -- 5278
(DE96011382)

Analysis And Control Of The METC Fluid Bed Gasifier

Quarterly Report
January - March 1995

March 1995

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

DOE/MC/31384 -- 5278
(DE96011382)
Distribution Category UC-103

Analysis And Control Of The METC Fluid Bed Gasifier

Quarterly Report
January - March 1995

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

For
U.S. Department of Energy
Office of Fossil Energy
Morgantown Energy Technology Center
P.O. Box 880
Morgantown, West Virginia 26507-0880

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

March 1995

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).

Table 1

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

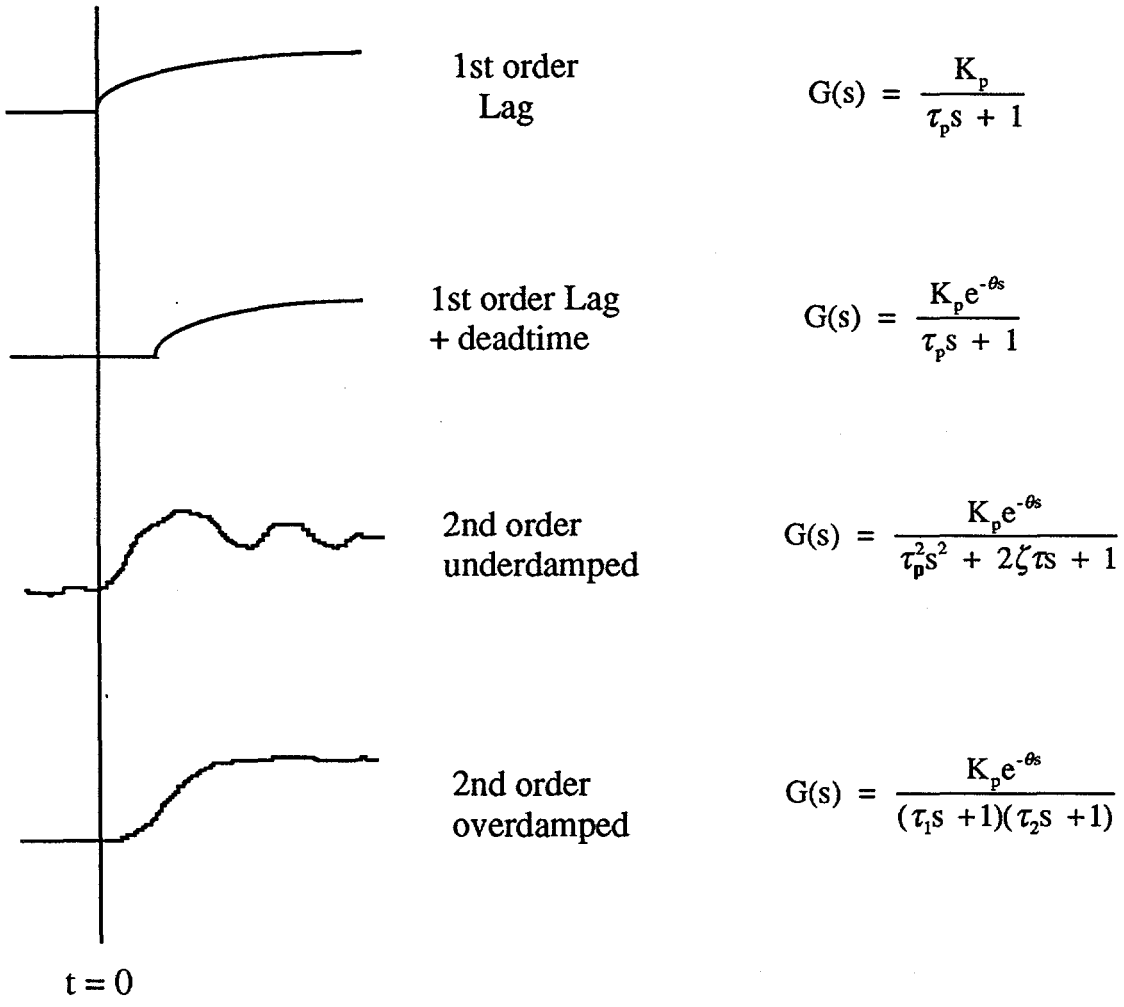

Figure 1: Some common open loop step responses and their appropriate transfer function models

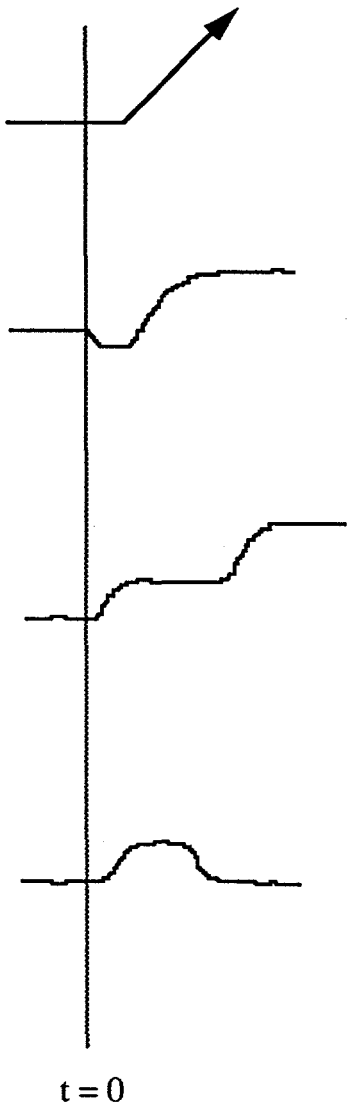

Figure 2: Some open loop step responses and their appropriate transfer function models

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.

References

1. Rhinehart, R.R., Felder, R.M., and J.K. Ferrell, "Dynamic Modeling of a Pilot-Scale Fluidized-Bed Coal Gasification Reactor," Ind. Eng. Chem. Res, 1987, 26, 738-745.

2. Purdy, M.J., Felder, R.M., and J.K. Ferrell, "Coal Gasification in a Pilot Scale Fluidized Bed Reactor. 2. Gasification of a New Mexico Subbituminous Coal," Ind. Eng. Chem. Res, 1984, 23, 287-294.

3. Rhinehart, R.R., Felder, R.M., and J.K. Ferrell, "Coal Gasification in a Pilot-Scale Fluidized Bed Reactor. 3. Gasification of a Texas Lignite," Ind. Eng. Chem. Res, 1987, 26, 2048-2057.

4. Neogi, D., Chang, C.C., Walawender, W.P., and L.T. Fan, "Study of Coal Gasification in an Experimental Fluidized Bed Reactor," AIChE Journal, Vol 32, No 1, Jan 1986.

5. Tsuje, T., and O Uemaki, "Coal Gasification in a Jet-Spouted Bed," The Canadian Journal of Chemical Engineering, Vol 72, June 1994.

6. Haji-Sulaiman, Z., Lim, C.J., and A.P. Watkinson, "Gas Composition and Temperature Profiles in a Spouted Bed Coal Gasifier," The Canadian Journal of Chemical Engineering, Vol 64, Feb 1986.

7. Arnold, M.J., Gale, J.J., and M.K. Laughlin, "The British Coal Spouted Fluidised Bed Gasification Process," The Canadian Journal of Chemical Engineering, Vol 70, Oct 1992.

8. Lim, C.J., Lucas, J.P., Haji-Sulaiman, M., and A.P. Watkinson, "A Mathematical Model of a Spouted Bed Gasifier," The Canadian Journal of Chemical Engineering, Vol 69, April 1991.

9. Kresta, J.V., MacGregor, J.F., and T.E. Marlin, "Multivariate Statistical Monitoring of Process Operating Performance," The Canadian Journal of Chemical Engineering, Vol 69, Feb 1991.

10. Aoufoussi, H., Perrier, M., Chaouki, J., Chavarie, C. and D. Dochain, "Feedback Linearizing Control of a Fluidized Bed Reactor," The Canadian Journal of Chemical Engineering, Vol 70, April 1992.

11. Palazoglu, A. and T. Khambanonda, "Dynamic Operability Analysis of Fluidized Catalytic Cracker," AIChE Journal, Vol 33, No 6, June 1987.

	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

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

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

MGAS

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

PDIR	K_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 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			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 11. Process Parameters for the response of Compositions and Outlet Flow for a change in Reactor Air