PARTIAL CONTROL OF COMPLEX CHEMICAL PLANTS

by

Arnon Arbel

A dissertation submitted to the Graduate Faculty in Engineering in partial fulfillment

of the requirements for the degree of Doctor of Philosophy,

The City University of New York.

1996

UMI Number: 9707065

UMI
300 North Zeeb Road
Ann Arbor, MI 48103

This manuscript has been read and accepted for the Graduate Faculty in Chemical Engineering in satisfaction of the dissertation requirement for the degree of Doctor of Philosophy.

2/20/96
Date

Prof. Reuel Shinnar, Chair of Examining Committee

2/13/96
Date

Prof. Irven H. Rinard, Co-Chair of Examining Committee

3/8/96
Date

Prof. Gerard G. Lowen, Executive Officer

Prof. Herbert Weinstein

Prof. Roberto Mauri

Dr. Ajit V. Sapre

Supervisory Committee

THE CITY UNIVERSITY OF NEW YORK

Abstract

PARTIAL CONTROL OF COMPLEX CHEMICAL PLANTS

by

Arnon Arbel

Advisors: Prof. Reuel Shinnar and Prof. Irven H. Rinard

Two of the control problems common to many chemical processes are that they are nonlinear and, that there are more outputs to be controlled than manipulated inputs to control them with. In this latter situation it is desirable to choose a control structure that, while controlling only few variables at their set points, keeps the whole vector of specifications and constraints within prescribed limits. This is called Partial Control.

This research deals with of how to choose the basic control structure so that the two above objectives are met. First, the availability of manipulated variables has to be considered. In most cases this is a function of the design and is given for operation processes. Second, the outputs to be controlled with the available inputs need to be chosen. This work outlines a systematic approach for choosing the best control structure for a process. A set of criteria is suggested by which to compare the different possible structures.

The approach of partial control is demonstrated on a Fluidized Bed Catalytic Cracker (FCC). The FCC is one of the most important units in the refinery train. It converts heavy gas oil to lighter gasoline and gas products. It is a highly nonlinear

process with a large dimension vector of specifications and constraints but very few manipulated inputs, hence a good candidate for partial control.

First, a new model of a generic FCC was developed. It was compared to other available models and commercial data. It was found to give good predictions of FCC operating conditions. Next, the nonlinearities of the process where studied. It was found that an FCC has, for a wide range of input conditions, both multiple steady states and input multiplicities. Several control structures that are in industrial use or discussed in the literature where investigated and evaluated based on the criteria suggested. This was done for both partial and complete CO combustion. It is shown that while for control at a given steady state linear methods are adequate, understanding the nonlinearities of the process is essential for choosing the basic control structure that will ensure stability and partial control.

Acknowledgments

I wish to gratefully acknowledge the patient guidance of my advisors, Prof. Reuel Shinnar and Prof. Irven Rinard, throughout the work on this project. I cannot imagine this work done without the love and support of Riki. Thank you.

Table of Contents

List of Tables

List of Figures

Nomenclature

Numbers in ( ) are values at the base case.

α	Fraction of C_{rgc} that enters the riser bottom	(1)
β	Fraction of CCR to go directly to coke	(1)
β_c	CO/CO_2 ratio at the catalyst surface in the regenerator	
β_{c0}	Pre-exponent constant in β_c expression	(2,512)
γ	Unstripped hydrocarbon [lb/lb cat]	
ΔH_{evp}	Heat of oil feed evaporation [Btu/lb]	(150)
$\Delta H_{f,CO}$	Heat of formation of CO [Btu/lb]	(47,549)
$\Delta H_{f,CO2}$	Heat of formation of CO_2 [Btu/lb]	(169,294)
$\Delta H_{f,H2O}$	Heat of formation of H_2O [Btu/lb]	(52,018)
ΔH_{ri}	Heat of cracking of lump i [Btu/lbmole]	(see table)
ΔT	Temperature rise from the regenerator dense bed to stack-gas (T_{sg}-T_{rgn}) [°F]	
ΔT_{st}	Stripper temperature drop [°F]	(25)
ε	Riser bed void fraction, or regenerator bed void fraction (position dependant)	
ε_{den}	Void fraction in regenerator dense bed	
ε_{dil}	Void fraction in regenerator dilute phase	
ρ_c	Catalyst density [lb/ft^3]	(68)
ρ_{den}	Catalyst density in the regenerator dense bed [lb/ft^3]	
ρ_{dil}	Catalyst density in the regenerator dilute bed [lb/ft^3]	
ρ_g	Molar gas density in the regenerator [lbmole/ft^3]	
ρ_v	Oil vapor density [lb/ft^3]	(0.57)

ρ_w	Regenerator wall material density [lb/ft^3]	(65)
τ_r	Residence time in the riser [sec]	
ϕ	Deactivation function [(lb coke/lb cat)$^{(1-1/b)}$]	
Ψ	Feed coking tendency function	
A	Relative catalyst activity	(1)
a_{ik}	Stoichiometric coefficient between lump i and k	(1)
Air/Oil	Air flow rate to Feed flow rate ratio ($F_{air} \cdot MW_{air}/F_{feed}$) [lb air/lb feed]	
A_{rgn}	Regenerator cross-section area [ft^2]	(590)
A_{ris}	Riser cross-section area [ft^2]	(6.4)
A_w	Regenerator wall area [ft^2]	(4300)
b	power in coking rate expression	(1/3)
C	Coke on catalyst during cracking [lb coke/lb cat]	
Cat/Oil	Catalyst flow rate to Feed flow rate ratio (F_{cat}/F_{feed}) [lb cat/lb feed]	
CCR	Conradson Carbon Residue in feed [wt%]	(0)
C_H	Weight fraction of hydrogen in coke	(0.075)
Conv.	Conversion of oil feed to gasoline, coke and wet gas [wt%]	
Cp_c	Catalyst heat capacity [Btu/lb°F]	(0.31)
Cp_{CO}	CO heat capacity [Btu/lb°F]	
Cp_{CO2}	CO_2 heat capacity [Btu/lb°F]	
Cp_{coke}	Coke heat capacity (graphite carbon) [Btu/lb°F]	
Cp_{fl}	Liquid oil feed heat capacity [Btu/lb°F]	(0.82)
Cp_{fv}	Vapor oil feed heat capacity [Btu/lb°F]	(0.81)

Cp_{H2O}	Water heat capacity [Btu/lb°F]	
Cp_{N2}	Nitrogen heat capacity [Btu/lb°F]	
Cp_{O2}	Oxygen heat capacity [Btu/lb°F]	
Cp_{sg}	Stack gas heat capacity [Btu/lb°F]	
Cp_w	Regenerator wall heat capacity [Btu/lb°F]	(0.31)
C_{rgc}	Coke on regenerated catalyst [wt% lb coke/lb cat]	
C_{sc}	Coke on spent catalyst [wt% lb coke/lb cat]	
d_w	Regenerator wall thickness [ft]	(1/3)
E_β	Activation energy for CO/CO_2 ratio at the surface[E/R, °R]	(12,232)
E_c	Activation energy for coke combustion [E/R, °R]	(34,000)
E_{cc}	Coking rate activation energy [Btu/lbmole]	(1000)
E_k	Cracking rate of lump k activation energy [Btu/lbmole]	(see table)
E_{k3c}	Activation energy for catalytic CO combustion [E/R, °R]	(25,000)
E_{k3h}	Activation energy for homogeneous CO combustion [E/R, °R]	(64,000)
F_{air}	Air flow rate to the regenerator [lbmole/sec]	
F_{cat}	Catalyst circulation rate (when F_{rgc}=F_{sc}) [lb/sec]	
f_{CO}	CO molar flow rate in gas [lbmole/sec]	
f_{CO2}	CO_2 molar flow rate in gas [lbmole/sec]	
F_{ent}	Entrained catalyst flow rate [lb/sec]	
F_{feed}	Oil feed flow rate [lb/sec]	
F_{feed}	Oil feed flow rate to the riser [lb/sec]	
f_{H2O}	Water molar flow rate in gas [lbmole/sec]	

f_{N2}	Nitrogen molar flow rate in gas [lbmole/sec]	
f_{O2}	Oxygen molar flow rate in gas [lbmole/sec]	
F_{rgc}	Regenerated catalyst circulation rate [lb/sec]	
F_{sc}	Spent catalyst circulation rate [lb/sec]	
F_{tf}	Oil feed flow rate [lb/sec]	(131)
f_{tot}	Total gas molar flow rate [lbmole/sec]	
G_{cd}	Primary dynamic control strategy	
G_s	Steady state control strategy	
h	Dimensionless riser height	
H_{ris}	Riser height [ft]	(120)
k	stripping steam coefficient	(0.08)
k_1	Coke combustion rate coefficient in reaction I (to CO)	
k_2	Coke combustion rate coefficient in reaction II (to CO_2)	
k_3	CO reaction rate coefficient to CO_2 in reaction III	
k_{3c}	Catalytic rate constant in reaction III [lbmole CO/lb cat sec psi^2]	
k_{3c0}	Pre-exponent for k_{3c} [lbmole CO/lb cat sec psi^2]	(0.54)
k_{3h}	Homogeneous rate constant in reaction III [lbmole CO/ft^3 sec psi^2	
k_{3h0}	Pre-exponent for k_{3h} [lbmole CO/ft^3 sec psi^2]	($1.465 \cdot 10^{11}$)
k_4	Hydrogen reaction rate coefficient to H_2O in reaction IV	
k_c	Total coke combustion rate coefficient [1/psi sec]	
k_{c0}	Pre-exponent for k_c [1/psi sec]	($7.275 \cdot 10^6$)
k_{cc0}	Coking rate pre-exponent coefficient	(0.64176)

k_h	Adsorption of heavy aromatics rate constant	
k_h	Absorption of heavy aromatics rate constant	(12.8)
K_{k0}	Cracking rate of lump k pre-exponent coefficient	(see table)
k_w	Thermal conductivity of the regenerator wall [Btu/ft^2 sec(°F/ft)]	($2.194 \cdot 10^{-4}$)
M	Reduced order nonlinear model of a process	
M^*	Full order unknown nonlinear model of a process	
MW_c	Coke molecular weight [lb/lbmole]	(12)
MW_H	Hydrogen molecular weight [lb/lbmole]	(2)
N	Unknown and unmeasured inputs (Disturbances)	
$O_{2,sg}$	Oxygen in stack gas [mole%]	
P_{CO}	CO partial pressure in the regenerator [psi]	
P_{O2}	Oxygen partial pressure in the regenerator [psi]	
P_{rgn}	Regenerator pressure [psi]	(29.6)
Q_{air}	Heat flow rate with air [Btu/sec]	
Q_c	Heat released by the carbon combustion [Btu/sec]	
Q_{ent}	Heat flow rate with entrained catalyst [Btu/sec]	
Q_{gas}	Heat flow rate with gas [Btu/sec]	
Q_H	Heat released by the hydrogen combustion [Btu/sec]	
Q_{lost}	Heat flow rate from the regenerator wall to the atmosphere [Btu/sec]	
Q_{rgc}	Heat flow rate with regenerated catalyst [Btu/sec]	
Q_{sc}	Heat flow rate with spent catalyst [Btu/sec]	
Q_{sg}	Heat flow rate with stack gas [Btu/lb]	

Q_{wall}	Heat flow rate to the regenerator wall [Btu/sec]	
R	Gas law constant	
r_c	Coke combustion rate [lbmole coke/lb cat sec]	
r_{co}	CO combustion rate [lbmole CO/ft^3 sec]	
s	Steam flow rate to the stripper [lb steam/sec]	(7)
t	time [sec]	
T	Position dependent temperature [°F]	
T_0	Riser bottom mix temperature [°F]	
T_{air}	Temperature of air to the regenerator [°F]	(300)
T_{atm}	Temperature of the atmosphere [°F]	(70)
T_{base}	Base temperature for heat balance calculations (T_{rgn}) [°F]	
T_{feed}	Oil feed preheat temperature [°F]	(670)
T_{mix}	Temperature at the riser bottom after feed and catalyst mix [°F]	
T_{ref}	Reference temperature for heats of formation [°F]	(77)
T_{rgn}	Regenerator dense bed temperature [°F]	
T_{ris}	Riser top temperature [°F]	
T_{sc}	Temperature of spent catalyst (T_{st}) [°F]	
T_{sg}	Stack gas temperature [°F]	
T_{st}	Stripper temperature [°F]	
T_w	Average temperature of regenerator wall [°F]	
u	Linear velocity in the riser or the regenerator [ft/sec]	
U	Vector of manipulated inputs	

U_{cd} Sub-vector of manipulated inputs entering the primary control

U_d Sub-vector of manipulated inputs varied that can be varied fast

U_{in} Heat transfer coefficient to the regenerator inside wall [Btu/ft^2 sec°F] $(8.333 \cdot 10^{-3})$

U_{out} Heat transfer coefficient from the regenerator outside wall [Btu/ft^2 sec°F] $(2.778 \cdot 10^{-3})$

U_s Sub-vector of manipulated inputs that can be varied slow

U^{ss} Vector of settings for the inputs at steady state

V Volume of each CSTR in gas flow modeling in the regenerator [ft^3]

W Vector of measured but not manipulated inputs

W_{rgn} Regenerator catalyst inventory [lb] (274,000)

W_{st} Stripper catalyst inventory [lb] (100,000)

X Reduced order vector of state variables

X* Full order vector of state variables

x_{pt} Relative catalytic CO combustion rate

Y Vector of outputs

Y_{cd} Sub-vector of outputs entering the primary control

Y_d Sub-vector of outputs measured frequently

Y_g Gasoline yield [wt%]

y_i Wt. fraction of lump i in the oil feed or products (in model eq. only)

Y_p Sub-space of outputs defining specifications and constraints

Y_s Sub-vector of outputs measured less frequently

Y^{ss}_{cd}	Vector of set points for Y_{cd}	
Y_{wg}	Wet gas yield [wt%]	
z	Relative catalyst coking rate	(1)
z_{bed}	Dense bed height in the regenerator [ft]	
z_{cyc}	Cyclone inlet height [ft]	(49)

Additional subscripts:

cyc	at cyclone inlet
in	flowing in
out	flowing out
sg	at stack gas

1. Introduction

In recent years it has become evident that control offers an important tool for improving the profitability and competitiveness of the chemical industry in general and refineries in particular. The huge growth in the power of computers makes it possible to implement advanced control strategies much easier. Modern methods such as Model Predictive Control are more and more common today. However, in most cases even these new and advanced algorithms result with a set of individual control loops for adjusting set points. Thus, the success of the control depends strongly on the proper choice of these loops.

Any control system has two main goals. One is called dynamic control which is to reject disturbances and keep the process at its chosen steady state. The other is called steady state control which is to move the system safely and smoothly from one steady state to another by changing the set points. Tuning controllers for dynamic control is a well addressed problem. In most cases using some kind of a linear method is sufficient as long as the work is at a given steady state point without too large a disturbance.

However, the problem of steady state control is more complex. First, most of the chemical processes are nonlinear and if the steady state change is large, the linearized models based on which the loops were tuned will change as well. Second, exact nonlinear models which are valid over a wide range are seldom available. Third, even when a good model is at hand, any system can undergo fundamental changes during its operation or because of process modification such as new equipment or new catalysts. The last and maybe most important problem in steady state control is that the number of variables that need to be controlled is often much larger than the number of manipulated variables available. While adding measurements to a working plant is relatively easy, adding manipulated variables might be very complex. Therefore, understanding how the choice of manipulated variables affects the efficiency of the control is important for better design.

There is a large body of theory that deals with the design of control algorithms given a set of observed and manipulated state variables [Astrom and Wittenmark (1984), Box and Jenkins (1970), Bristol (1966), Brosilow and Tong (1978), Edgar (1980), Grossman and Morari (1984), Joseph and Brosilow (1978), Kalman and Bertram (1958), Luyben (1990), Morari and Zafiriou (1988),Morari, Arkun and Stephanopolous (1982), Morshedi, Cutler and Skrovaned (1985), Prett and Morari (1987), Ray (1983), Safanov and Athans (1977), Stephanopolous (1983a,b)]. Most present research in process control continues to focus on these problems, especially on the effect of nonlinearities. While there are many important and unsolved problems in this area, there is another important area which has received less attention, namely, the question as to how to choose the control structure, in particular, the set of variables that comprise this structure. In a large number of applications such as, for example, the control of heat exchangers, this choice is obvious. The more complex the process to be controlled, the less obvious is this choice. It all too often becomes a question of experience and intuition.

If the number of output variables that need to be controlled is larger than the number of manipulated variables, then it is impossible to keep all the outputs at exact set points. At best, it might be possible to keep them within an acceptable space. This can be done in two ways. One is to use a full nonlinear model employing all available information from all measurements to adjust the manipulated variables directly. The other is to use a set of specially chosen output variables in a square control structure with the manipulated inputs to keep the system stable in the face of disturbances. The information from all other measured variables is then used to adjust the set points of this structure to keep the required specification variables in an acceptable space and to move it to a new steady state if required. The second method is much more frequently used because direct control requires a much more accurate model, one which is seldom available.

Controlling a set of variables to keep them in a defined space is different from the conventional concept of controllability which deals with problems in which control can be applied in a more rigorous way and variables brought to a desired steady state. Shinnar [1981,1986] called this type of problem *partial control.* Partial control is important not only in control of refineries and complex plants, but also in many other complex systems. The present concepts of economic control are based on controlling a very limited set of dominant variables such as interest rates, money supply and taxes and thereby keep the total economy on a desired course.

This work has a twofold goal. The first is to define and formulate the concepts of partial control. The second is to apply them to a Fluidized Catalytic Cracker (FCC) and so to develop a better understanding of the control problems of an FCC in the context of model based control. The next section will deal with the theoretical formulation of partial control. It will try to present a methodical approach to investigating a control problem from partial control point of view. The following sections will follow this approach using the FCC as an example. First a model will be presented and evaluated. Next the nonlinear features of the system will be studied, and last, different control strategies for different operating modes will be considered.

This is a good place to point out that the approach outlined here is of course not the only way to deal with partial control. Another well known way is Model Predictive Control (MPC). In theory, if a good overall nonlinear model is available it is possible to use it to directly control all manipulated variables, both slow and fast, to monitor the system in the desired operating space (see for example Balchen, Ljungquiest and Strand, 1991). In some cases this has been applied quite satisfactorily. Another MPC approach is known as Dynamic Matrix Control (DMC) (Cutler and Ramaker, 1980) that uses linearized model obtained by step perturbations. However, all the MPC methods assume that the control structure is given while this work is an effort to show how to choose this structure. Therefore the approach taken here is essential to properly understand and operate a MPC controller.

The problem of enforcing constraints in excess of the number of manipulatable variables is mentioned briefly in the context of Quadratic Dynamic Matrix Control (QDMC) by Prett and Garcia (1988). A third approach bearing some similarities to partial control is Modular Multivariable Control (Meadcroft, Stephanopoulos, and Brosilow 1992). While all of these approaches deal with the enforcement of constraints, none addresses the problem that must be solved in addition, namely, the stabilization of highly nonlinear systems.

2. General Principles of Partial Control

The formulation of partial control was outlined for a chemical reactor by Shinnar (1981, 1986). A short summary of the pertinent aspects follows.

2.1. Definition and Concept

Any nonlinear system, such as a chemical reactor, can be described by a nonlinear model in the general form:

$$X^{*}=M^{*}(X^{*},U,W,N) \tag{1}$$

where X^{*} is a vector of the state variables, U a vector of manipulated inputs, W a vector of inputs that can be measured but not manipulated, and N a vector of unknown and unmeasured inputs.

In general, the state vector may be very large. Some of the states may not be known, or more often may not be measurable. Moreover, as the system becomes more complex less is known about M^{*}. A simplified model M can be identified by a combination of identification and mathematical modelling (Aris, 1978; Denn, 1986; Shinnar, 1978). This model will be in the sub-space:

$$Y=M(X,U,W,N) \tag{2}$$

where: Y is a vector of outputs that are measured and may contain some of the state variables that can be measured. X is a reduced vector of the state variable that can be recognized. However, not all of Y and U are used equally in the control system. Some are integrated into the dynamic control, some are used to adjust set points, and some are used mainly for data logging and diagnostic purposes. Y and U are divided therefore into two subsets. Y_d (U_d) contains the process variables monitored (manipulated) either continuously or with small sampling intervals compared to the overall time scale of the process. Y_s (U_s) contains the infrequently measured (manipulated) process variables. From a formal point of view this division of Y into Y_d and Y_s could be carried out by technique of singular perturbations used in modeling and control of multitime-scale systems (Kokotovic et al., 1976; Martinez and Drozdowicz, 1989). However, in many cases (including the FCC), the decomposition is obvious.

There is, however, another way to divide Y into sub-spaces that is important for deriving the control strategy. Y_p is defined as the set of process variables defining the product and process specifications and process constraints. The elements of Y_p may lay in both Y_d and Y_s and its dimension is generally smaller than Y. For chemical reactors Y_p is seldom completely contained in Y_d and part of it lies in Y_s. Next, Y_{cd} , a subset of Y_d, is defined as the set of process variables on which the dynamic control strategy is based. Although Y_d may be quite large, a subset of much smaller dimensions is normally chosen as the basis for the dynamic control. If integral control is to be used, the control matrix must be square, and its dimensions are limited by the number of manipulated variables available. The elements of Y_{cd} are not necessarily part of Y_p, but usually are.

The distinction between partial control and exact control comes in the definition of the control objectives. The objectives for exact control are to keep the controlled variables at their set points. Reactor partial control in general is a strategy to develop a sufficiently large subset of U and use it to maintain Y_p within prescribed limits:

$$Y_{pj\,min} < Y_{pj} < Y_{pj\,max} \tag{3}$$

To achieve the goal of Eq. 3, a primary dynamic control system, G_{cd}, is defined first which uses a small set of manipulated variables, U_{cd} to keep a small set of measured variables, Y_{cd} at their set points:

$$U_{cd} = G_{cd}(Y_{cd}, W) \tag{4}$$

The set points Y^{ss}_{cd} for this primary matrix are determined by a steady state control system, G_s, which also determines the steady state values for the rest of the controlled inputs U:

$$(Y^{ss}_{cd}, U^{ss}|_{U_{cd} \epsilon U}) = G_s(Y|_{Y_{cd} \epsilon Y}, U|_{U_{cd} \epsilon U}, W) \tag{5}$$

When the superscript 'ss' denotes steady state values. Note that it is very important to use all the information available to determine the steady state values for all those manipulated inputs that do not enter U_{cd} and the set points for the dynamic control G_{cd}.

This is a good place to point out that the approach outlined above is of course not the only way to deal with partial control. Another well known way is Model Predictive Control (MPC). In theory, if a good overall nonlinear model is available, it is possible to use it to directly control all manipulated variables, both slow and fast, to monitor the system in the desired operating space (see for example Balchen, Ljungquiest and Strand, 1991). In some cases this has been applied quite satisfactorily. Another MPC approach is known as Dynamic Matrix Control (DMC) (Cutler and Ramaker, 1980) that uses linearized model obtained by step perturbations. However, all the MPC methods assume that the control structure is given while here the purpose is to show how to choose this structure. Therefore the approach taken in this paper is essential to properly understand and operate a MPC controller.

The problem of enforcing constraints in excess of the number of manipulatable variables is mentioned briefly in the context of Quadratic Dynamic Matrix Control (QDMC) by Prett and Garcia (1988). A third approach bearing some similarities to partial control is Modular Multivariable Control (Meadcroft, Stephanopoulos, and Brosilow 1992). While all of these approaches deal with the enforcement of constraints, none addresses the problem that must be solved in addition, namely, the stabilization of highly nonlinear systems.

2.2. Criteria for the Choice of the Primary Control Structure

Any design, including that of control structures, is based on compromises. Before embarking on evaluating the different control structures, some design criteria must be defined. A list of the proposed criteria is given in Table 1 where the order is not any indication of importance. The criteria are very much tied together and it is almost impossible to discuss them separately. However, an attempt is made to define each one by itself.

Table 1: Criteria for evaluating possible primary control structures

●	Modellability
●	Dominance
●	Nonlinearities and Stability
●	Time scales of responses to set point change and disturbances
●	Sufficiency

1. Modellability. While accurate models are seldom available, in most cases laboratory data are. It is essential to know how the measured variables relate to the specifications and constraints. While accurate predictions are not needed as feedback is used, the least one needs is an idea of what are reasonable set points and what the signs of the gains are. Further, the model sensitivity of the estimate for individual variables varies considerably. For example, in FCC reactor temperature is strongly constrained by heat and mass balances and is much less sensitive to kinetic or hydrodynamic parameters as compared to afterburn in the reactor's cyclones or flue gas. Also, laboratory experiments are seldom adiabatic and most of the data available on reactions are at isothermal conditions. As the measured variables in themselves are not the goal, but serve for set point control, their modellability has a strong impact on their suitability for the primary control matrix. It is needed to be able to obtain information about how the desired output variables Y_p (specification and constraints) related to the set points Y_{cd}.

2. Dominance. In a reactor with a complex reaction network, temperature has a strong impact on the relative and absolute rate of all reactions as different reactions have different activation energy. Thus, temperature is a dominant

variable and it is this feature that allows one to keep a large set of variables within a limited space by manipulating only a smaller number of set points. Variables other than reactor temperature may have similar properties. In partial control it is therefore necessary to choose dominant variables for the set points. This is important both for disturbance rejection and change of operating conditions (or in other words intentional change of the output vector).

3. Nonlinearities and stability considerations. The first nonlinear consideration is multiple steady states. That is, one input setting may result in two or more sets of outputs. For a given set of inputs there might be only a range of set points where there are steady states or where the steady states are desirable. For a system with multiple steady states all desirable steady states are unstable in the following sense. Even if a choice of set points is stable, a strong perturbation in an uncontrolled input might cause the system to drift to another steady state. Further, if operating conditions change, that stable steady state might change to be unstable. For both cases it is desirable for the control structure to keep the process stable. A second nonlinear consideration is that multiplicities in the inputs lead to multiplicities in the outputs. That is, one set outputs can be reached by two or more sets of inputs. It is important that the control structure is unique in the sense that it will not lead to a different operating point by choosing the wrong input settings although the output set points is correct. Although some of the output multiplicities might not be practical, each case has to be checked. A third nonlinear feature is the change of the gain around the operating point. For any

control loop to be effective, there must be a finite gain between the input and the output. In nonlinear systems this gain changes from operating point to operating point and may even change sign. Thus, the operating point must be chosen such that the gain stays reasonable. Since the gains at different operating points may change their magnitude as well as their sign, their relation to modellability is an important consideration in choosing a control structure.

4. Time scale of responses. While fast response is an important criterion in the design of the control loops, it is not the dominant criterion in the choice of the structure. Rather, it is one of several often conflicting criteria which the designer must take into account. Very often in the literature there is an emphasis on choosing a variable that has a fast and smooth response. However, fast response is only useful if the variable chosen is a critical specification of the process or it is a dominant variable. For example, an FCC has a large thermal holdup. As a result the regenerator temperature responds very slowly. This is beneficial as the system is damped and the thermal inertia protects it by giving the operator more time to take action. Slow response is often intentionally introduced by the designer, and is essential for the controllability of uncertain systems. On the other hand, a hydrocracker is sensitive to very small perturbations in inlet temperature and therefore fast control of the inlet temperature is essential (Silverstein and Shinnar, 1982).

5. Sufficiency. A very important criterion in the design of any control system is to match the capability of the control to the desired specifications (of both process

and products) and constraints. This criterion is not an independent one but based on the Criteria 1 to 4, nor is it easily definable. Specification are a result of both market demand and process capabilities. Process capabilities are strongly determined by the plant design as well as by the number and nature of manipulated variables. These are determined during the design of the plant. Many times plants encounter severe control problem due to changed specifications that are inconsistent with the capabilities of the plant. The minimum requirement for any complex reactor is that the control should able to keep it stable in the face of reasonable perturbations, but in most cases much more than this is required. This is a highly neglected area both in research and practice. Estimating these capabilities and their relation to initial design or needed modifications is a critical area for the design of the control structure as well as of the process.

Note that all five criteria have both linear and nonlinear aspects. All but Criterion 3 can be defined by using linear examples. In practice the importance of nonlinear considerations changes from process to process. However, when nonlinear considerations are important they are more critical in the choice of the control structure and much less so in the tuning and choice of the control algorithms for control of a specific steady state.

3. Modelling a Fluidized Catalytic Cracker

3.1. Introduction

Worldwide, the Fluidized Catalytic Cracker (FCC) is the workhorse of the modern refinery. Its function is to convert heavy hydrocarbon petroleum fractions into a slate of more useable products such as gasoline, middle distillates, and light olefins. First commercialized over a half century ago, the FCC is still evolving. Improvements in the technology as well as changing feed stocks and product requirements continue to drive this evolution.

Control of the FCC has been and continues to be a challenging and important problem. As will be seen, its steady state behavior is highly nonlinear, leading to multiple steady states, input multiplicities and all that this implies. In earlier years before the development of zeolite catalysts, the major control problem was one of stabilization, of just keeping the unit running. Later with zeolite catalysts the emphasis shifted to increasing production rates in the face of unit constraints and to handling heavier feeds. The requirements for reformulated gasoline have added the need to control product composition.

Before undertaking the study of the dynamic and steady state control of any complex system, an adequate model is needed. In the case of the FCC, the literature abounds with such. Before initiating this control study a thorough evaluation of available models was done and it was found that none of them was detailed enough to be suitable for this control study.

Their range of applicability was either too limited, or their behavior was not consistent either with published data or with available knowledge of the behavior of industrial units. It was decided therefore to develop a better, much more detailed model based on the full extent of published knowledge, described here in detail. Also, previous work on FCC modeling was utilized whenever it was found to be applicable. The model is built in a modular form such that a more detailed description of the kinetics in both the reactor and regenerator (such as more lumps and relations predicting octane and gas composition, or SO_2 and NOx in the regenerator) can be easily incorporated.

3.2. Description of the System

For a detailed description of the FCC, its historic development, the different types of units in current use, and the design philosophy underlying them, one can refer to Avidan and Shinnar (1990), Avidan et al. (1990), Ewell and Gadmer (1978) and McDonald and Harkins (1987). The following is a short description of the different types of current units. A schematic general description is given in Fig. 1, which is a more general generic description of an FCC. However, not all elements are presented in all units and furthermore not all control and design features are present in all units.

The two basic units comprising an FCC are a reactor in which hot catalyst is brought in contact with feed oil, and a regenerator in which the spent catalyst is regenerated by burning of the coke formed during the cracking. The heat of combustion raises the temperature of the catalyst recycling from the regenerator. In turn this hot catalyst supplies the heat required for heating and evaporating the feed as well as for the endothermic heat of the cracking reaction. The hot product vapor is fed to a distillation

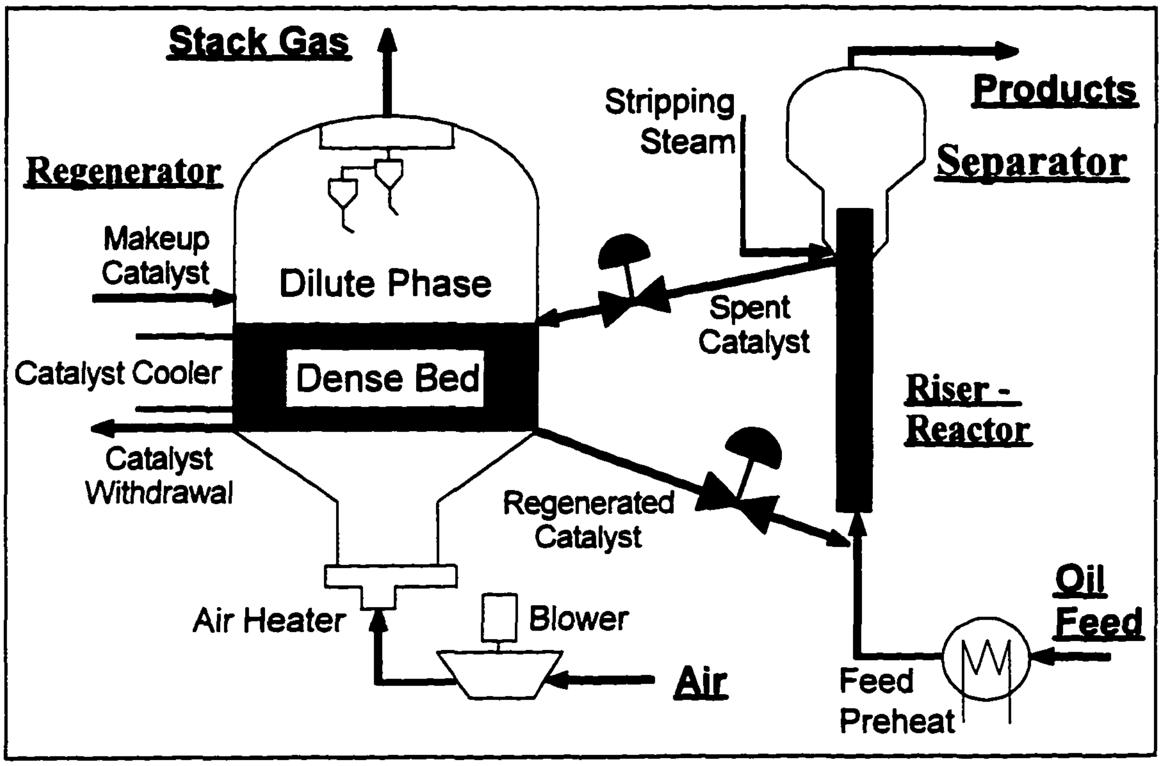


Figure 1: Schematic description of FCC

column where it is separated into a gasoline fraction , a light cycle oil in the diesel boiling range and a heavy bottom. The uncondensed vapor is compressed to about 200 psi which allows recovery of lower boiling gasoline fractions as well as butanes, butenes, propane and propene by absorption in oil. The C_2^- fraction containing some unrecovered C_3's is used as fuel gas in the refinery. In today's operation $C_4^=$ and $C_3^=$ formed in the cracker are used together with isobutane as alkylation feed. Therefore it is important to include their yields in the product description.

The design of the reactor varies but almost all reactors presently in use are risers with relatively short contact times of catalyst and feed. A main difference between designs is the way the catalyst circulation to the reactor is controlled. In most modern units it is adjustable by a slide valve in the pipe feeding the hot catalyst to the reactor.

This allows a variation of the catalyst flow over a relatively wide range (factor of 3). In some older units catalyst circulation rate can only be adjusted over a narrow range by the pressure balance, the key variable being the catalyst holdup in the regenerator. The slide valve has a strong impact on the design of the FCC control, as it allows for fast and accurate control of the reactor temperature. Catalyst residence time in the stripper as well as the ratio of the steam to catalyst can be varied adding another control to the heat balance.

Most modern reactors also have a feed preheater. Feed should not be heated above 750°F to prevent coking of the heating coils. Unlike other reactors, preheat is often just a control feature. Very few FCC's need additional heat; most can use heat removal. But the ability to adjust feed temperature by a feed preheater gives an additional steady state variable one can manipulate to adjust the heat balance. This allows increasing throughput despite constraints in the maximum catalyst flow rate and air flow rate. Controlling feed temperature also allows some dynamic control of reactor temperature in old units where the slide valve was used for emergency shutdown only.

Regenerator designs have a wider variation. There are two basic designs. One is a fixed fluid bed operating at a high velocity (3 ft/s) and therefore a high recirculation rate through the cyclones. (Catalyst inventory circulates through the cyclones approximately every five minutes). This gives a high catalyst density (0.5 - 1 lb/ft^3) in the dilute phase which is important for the regenerator operation. The other design uses a circulating fluid bed regenerator with multiple air inputs as shown in Fig. 2. The division between fixed and circulating fluid beds is somewhat arbitrary since in the fixed bed there

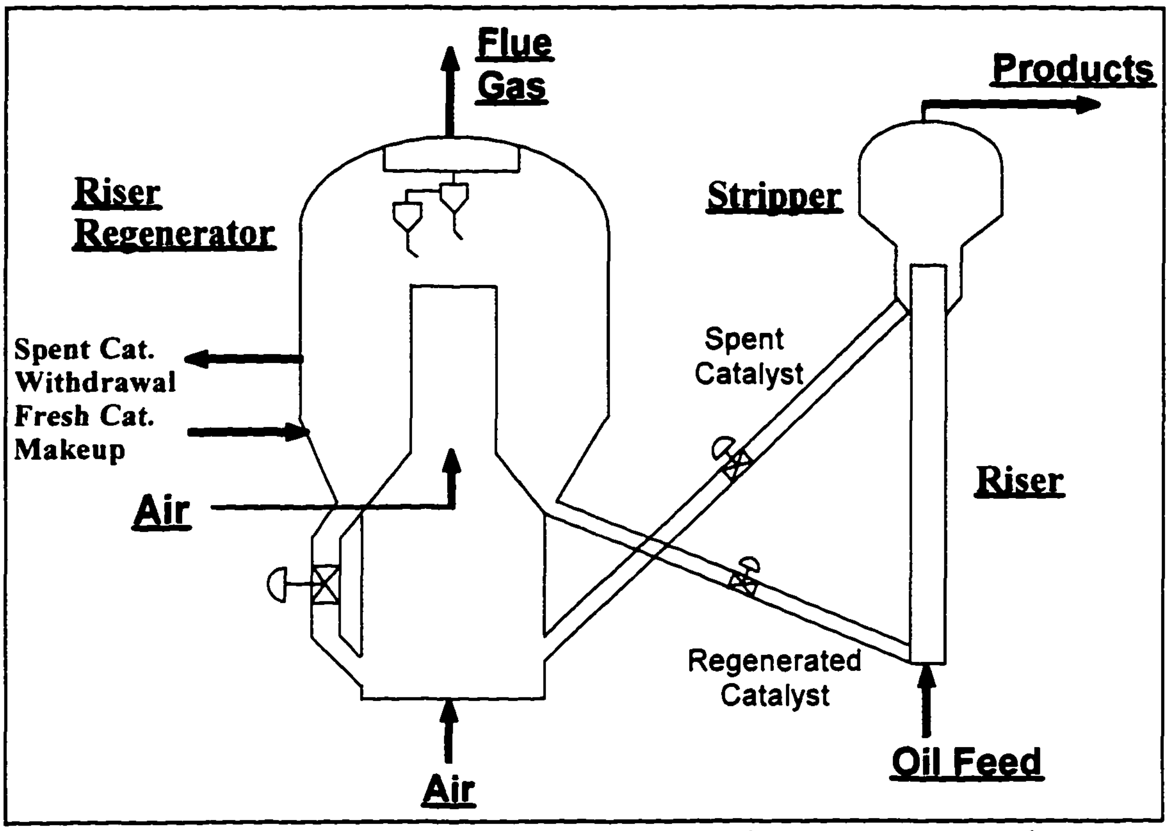


Figure 2: Schematic diagram of high-efficiency FCC with riser reactor and riser regenerator.

is also a large catalyst circulation. However, in the circulating bed catalyst densities in the riser section are much higher (10 lb/ft^3 in circulating bed vs. 1 lb/ft^3 in fixed bed) and the catalyst recirculation is independently adjustable.

Some units also have various types of catalyst coolers which allow removal of excess heat from the unit. Most large units of this type also have a CO boiler. Operating in the partial combustion mode (below 1250°F), about 50% of the carbon ends up in the flue gas as CO. This has to be combusted in a boiler before being emitted to the atmosphere. The combustion of coke to CO_2 generates three times as much heat (and uses twice the air) compared to its combustion to CO. Therefore, operating in partial combustion reduces both the heat generated in the regenerator and the air required for a specific amount of coke. This allows processing of heavier feeds which tend to generate more coke. Smaller units without such a boiler have to use a promoter to achieve full CO combustion in order to avoid CO emissions.

One unique feature of the FCC reactor is the behavior of the catalyst. The catalyst loses its activity very rapidly (within seconds) due to coking. This loss of activity can be recovered by combusting the coke in the regenerator. The catalyst also loses activity more slowly but permanently due to steaming and exposure to higher temperatures. The first problem is solved by a design in which the sojourn time of the catalyst in the reactor between regenerations is measured in seconds. The permanent deactivation is taken care of by daily removal of old and addition of fresh catalyst. The average total sojourn time of the catalyst in the system is two to three months.

In fixed bed reactors catalyst activity loss can only be compensated for by changing reactor temperature. In a fluidized bed, activity can be adjusted by catalyst management (addition and withdrawal) independent of reactor temperature. This gives the FCC control capabilities that few other reactors have. Furthermore, various catalyst additives can be added to modify some of the actions of the main catalyst. Some, like HZM5, act on the primary products of the cracking reactions, cracking higher boiling olefins and paraffins resulting in higher octane and a higher C_4 and C_3 olefin make. Another additive is used to modify the reaction in the regenerator promoting the combustion of CO to CO_2. Other additives are used to capture SO_2 in the regenerator releasing the sulfur as H_2S in the reactor. There is no apparent reason why there modifiers can not be incorporated into the main catalyst. Adding them separately however, increases the flexibility of reactor control.

3.3. History of FCC Modeling - A Short Literature Survey

Numerous papers relating to the FCC process can be found in the published literature. They present various aspects of mathematical modeling and simulation, stability, optimization and optimal control. As the investigation of the dynamics and control of a system requires a model to work with, this is the subject of many of the FCC papers. Some papers include integrated models for the regenerator and reactor coupled [Luyben and Lamb (1963), Kurihara (1967), Iscol (1970), Lee and Kugelman (1973), Seko et al. (1978, 1982), Lee and Groves (1985), McGreavy and Isles-Smith (1986), Bozicevic and Lukect (1987), Zhao and Lu (1988), Felipe and Richard (1991), Arandes and de Lasa (1992), Elnashaie and Elshshini (1979, 1990, 1993), McFarlane et al. (1993) and Zheng (1994)]; some have only regenerator models [Ford et al. (1976), Errazu et al. (1979), de Lasa et al. (1981), Guigon and Large (1984), Krishna and Parkin (1985) and Lee and Cheng (1989)]; and others reactor or cracking models [Weekman and Nace (1970), Paraskos et al. (1976), Jacob et al. (1976), Shah et al. (1977), Takatsuka et al. (1987), Lee et al. (1989), Larocca et al. (1990), Shnaider and Shnaider (1990) and Farag et al. (1993)].

Some of these papers present useful approaches but insufficient information to allow to their use. For the purposes of this work it is sufficient to discuss the properties of two of the most recent models used by investigators as these incorporate most of the previous work. These are the model published by McFarlane et al. (1993) and the model proposed by Lee and Groves (1985). The latter is used with slight modifications both by Hovd and Skogestad (1992) and by Balchen et al. (1992).

McFarlane et al. (1993) published their model in the context of the AIChE industrial challenge problem. This model describes an old Exxon Model IV unit which has no independent way to control catalyst flow and is rather obsolete today. Most of the model is dedicated to computing the complex pressure balance and catalyst circulation rate in this kind of units. The model is valuable because it gives very detailed and realistic description of the fluid dynamic behavior of the regenerator (bed density, distribution, flow and catalyst density in the dilute phase) as well as a detailed description of the catalyst circulation in that specific reactor. McFarlane's model lacks a detailed description of the reactor kinetics. It does not predict feed conversion, the main purpose of the FCC. Further, the heat balance is strongly affected by the endothermic heats of the cracking reactions. As a result, one does not obtain a proper heat balance and this masks some of the major interactions in the FCC. As a result, the reactor part of the model is useless for steady state control and of limited validity for studying dynamic control.

Another critical deficiency for the purposes of this work is the lack of detailed kinetic information for the combustion reactions in the regenerator which is based on Ford et al. (1976). While Ford et al. (1976) describe a detailed kinetic scheme for the combustion reactions in the regenerator, they give no rate constants. The rate constants given by McFarlane et al. correspond to a highly promoted case, and always lead to full CO combustion. This prevents any modelling of operation in partial combustion which is still the practice of industry in many units. In fact, the majority of present units under construction are equipped with a CO boiler to allow them to operate in partial combustion. Even for units operating in full CO combustion, the model is not really adequate as it does not predict either afterburn or the effect of changing the activity of the combustion promoter.

The model used by Lee and Groves (1985) differs in the fluid dynamic description of the regenerator. Based on Errazu et al. (1979), it uses a simple stirred tank with no dilute phase. Actually, Errazu started with a two phase model but showed that with realistic parameters, this is very close to a stirred tank. The main problem of the model is a lack of detailed kinetics for the combustion of CO and CO_2 which occurs both at the solid catalyst surface [Weisz (1966)] and in the homogeneous phase [Upson et al. (1993)]. Errazu used the CO_2/CO ratio suggested by Weisz as a correlation with temperature only. This does not allow modelling of the impact of catalytic combustion promoters and it neglects the impact of coke on catalyst on the relative kinetics. Most important, since Errazu's correlations are valid only up to 1240°F [Errazu et al. 1979)] it does not allow to study the transition to full combustion (or the reverse), a control problem in present units.

On the reactor side, Lee and Groves use the three lump model proposed by Weekman (1979). For many purposes three lumps are adequate, though one has to be careful. If the feed changes, the rate constants change both absolutely and relatively. However, Weekman's original model had one problem as pointed out by Krambeck (1991). The coking rate and catalyst deactivation are not properly coupled.

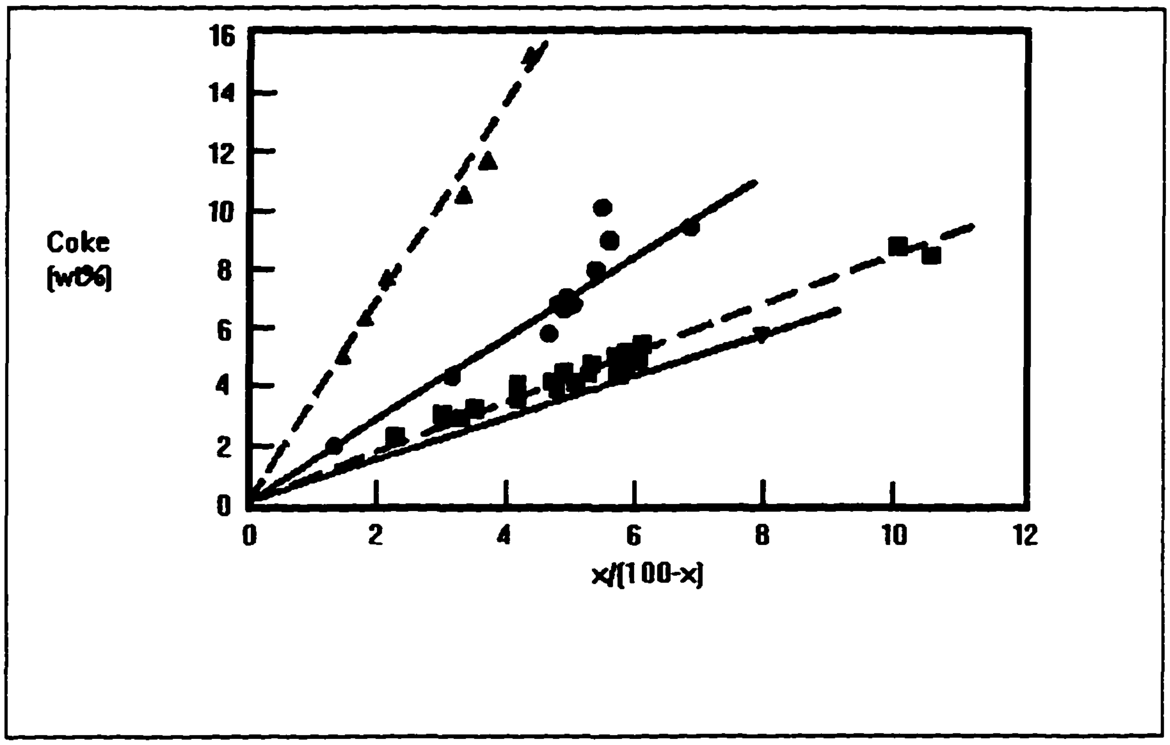


Figure 3: Coke yield vs. conversion (plotted as *x/(100-x)*). Predicted (lines) vs. actual data (symbols) for four different feedstocks. [From Krambeck (1991)].

As shown by Krambeck (1991) and Sapre and Leib (1991), in an isothermal reactor for a given catalyst and feed, conversion is a unique function of total coke produced, as shown in Fig. 3 taken from Krambeck (1991). Thus, coke selectivity (coke produced versus oil converted) is not a function of the cat to oil ratio or coke on regenerated catalyst, but is only a function of temperature, catalyst and feed properties. The model of Lee and Groves (1985) as used by Hovd and Skogestad (1992) does not have this property as can be seen from Fig. 4. This is a critical property which can lead to wrong steady state and dynamic responses to changes in feed properties or operating conditions. This requires a modification of the reactor model. A more detailed ten lump model based on Jacobs et al. (1976) which automatically takes care of the changes in rate constants due to feed properties, together with the catalyst deactivation function proposed

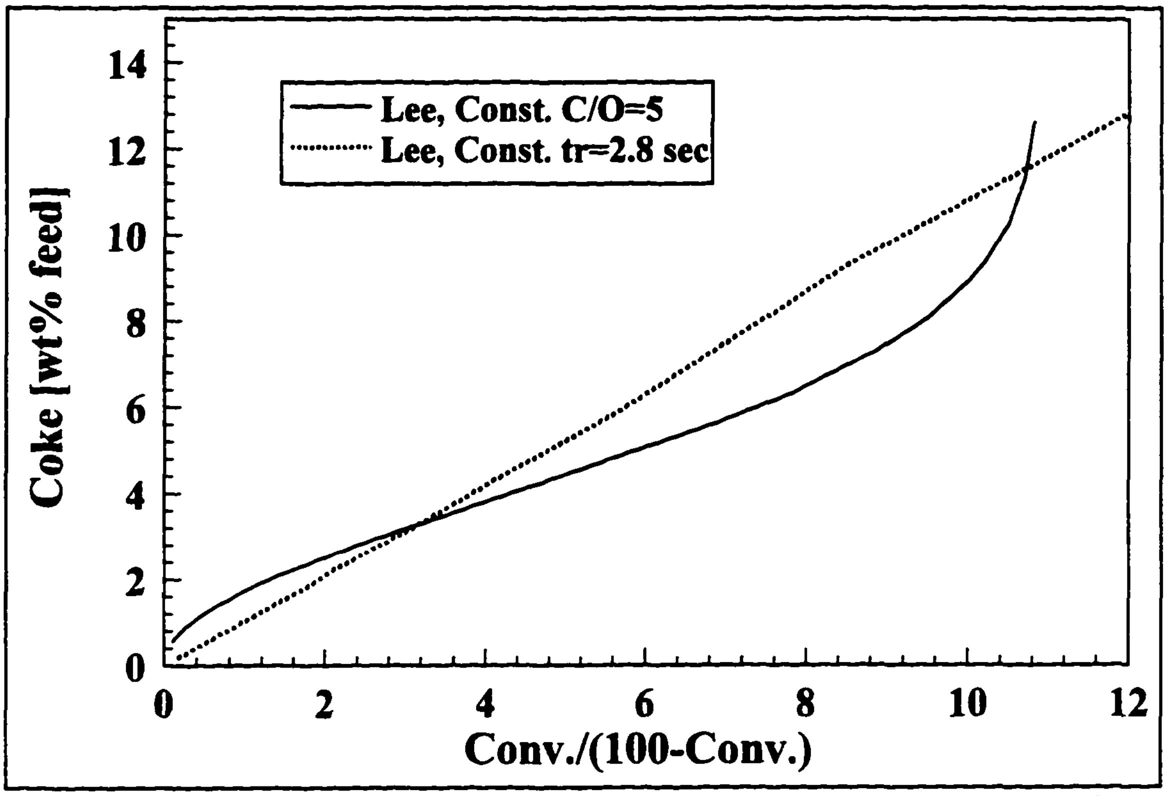


Figure 4: Lee and Grove (1985) riser model at 1000°F. Effect of changing Cat/Oil and residence time.

by Krambeck (1991) is used in the model presented here.

There is another set of papers that is just as crucial to the purposes of this work, those dealing with the kinetic processes in the reactor and regenerator and giving actual results for the FCC. The experimental and theoretical basis for describing the reactions in the regenerator is the work of Paul Weisz and associates (1966) and of Dwight Prater and associates (1985). Their work on TCC regenerators is also relevant [Prater et al. (1983)]. They provide rate expressions for coke combustion and for the CO to CO_2 reaction. Data for the homogeneous reaction of CO to CO_2 in the presence of a catalyst are given by Upson et al (1993). The impact of catalytic combustion promoters is discussed by Chester et al. (1981). Data on coking and cracking are given by Sapre and

Leib (1991) and in many papers describing FCC operation by catalyst vendors. The latter are normally not in a form that allows deriving actual rate data but are useful for checking trends and overall rates.

3.4. The Major Features of the New FCC Model

Based on the needs described above, a new model for the FCC has been developed. The exact model equations are summarized in the Appendix.

3.4.1. Regenerator

The behavior of the regenerator dominates both the dynamic and steady state behavior of the system. This is due to the adiabatic nature of the system in which the need to balance coke formation and combustion is the overriding driving force. The heat of combustion released in the regenerator is therefore the most critical item in any simulation. For a given amount of air, the heat released is a strong function of gas composition. All hydrogen in the coke will be converted to steam. Carbon can be converted to either CO or CO_2. As the heat of combustion to CO_2 is almost three times the heat of combustion to CO, it is very important to model correctly the impact of operating conditions on the ratio of CO_2 to CO as this controls the heat balance.

Kurihara (1967) neglected this problem by using a fixed CO_2/CO ratio. Arthur (1951) found that at the catalyst surface the coke burns to CO_2 and CO in a ratio which decreases with increasing temperature (Fig. 5). But the CO continues to react to CO_2 both on the catalyst surface [Weisz (1966)] and in the gas phase. Based on Weisz's data

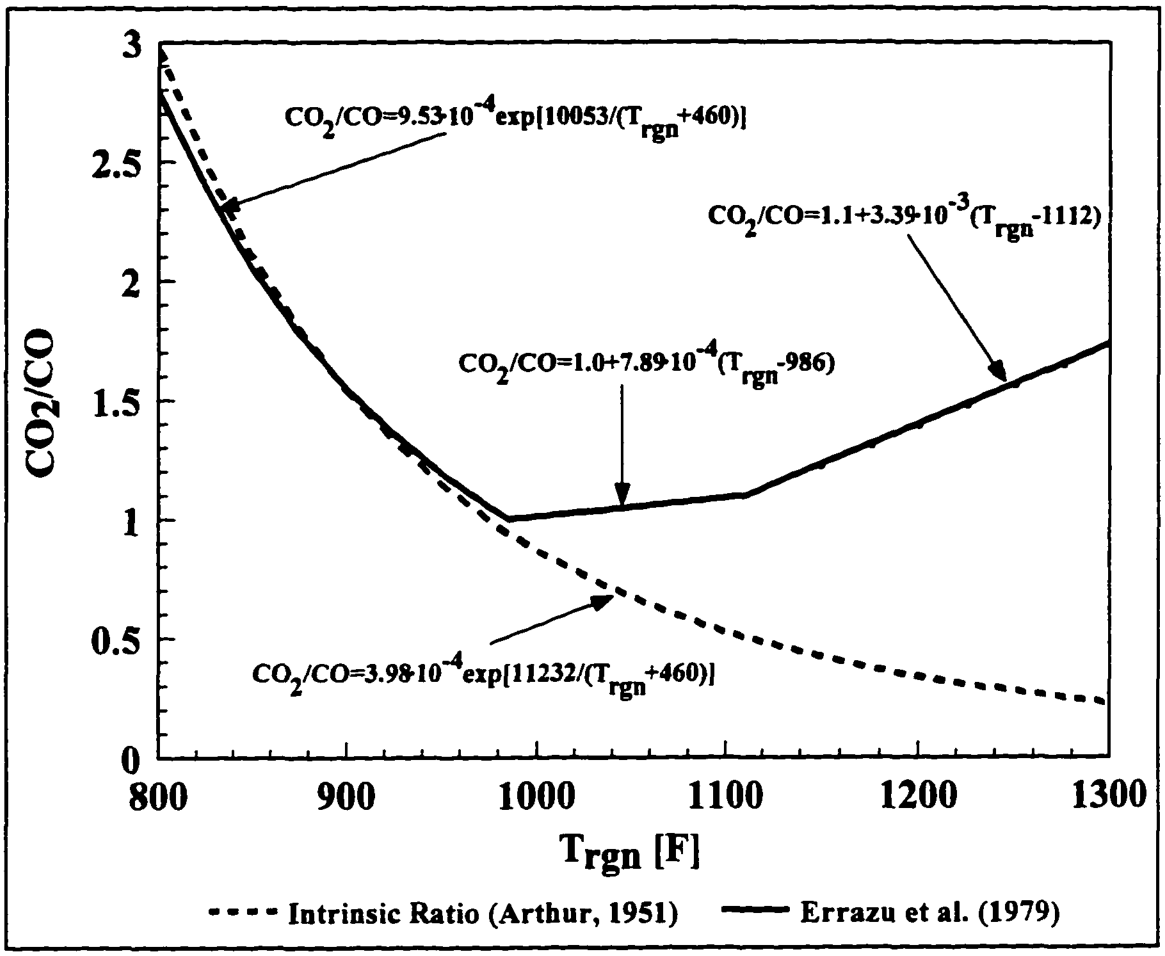


Figure 5: CO_2/CO ration vs. regenerator temperature.

Errazu et al. (1979) neglected the homogenous reaction and used an empirical correlation valid up to 1240°F, also shown in Fig. 5. Errazu's ratio has positive slope with increasing temperatures above 980°F. Both CO to CO_2 reactions are rather complex. The catalytic surface reaction is promoted (1) by combustion promoters such as platinum [Chester (1981)] which are widely in use today, (2) by unpromoted zeolites and, (3) by metals such as nickel and vanadium deposited on the catalyst. Different catalysts may therefore have varying unpromoted activity. However, the catalytic activity of a catalyst to promote the reaction of CO to CO_2 can be measured accurately in the laboratory and used for modelling specific cases. In this model it is an adjustable parameter. As a base case, the rate reported by Weisz (1966) is used.

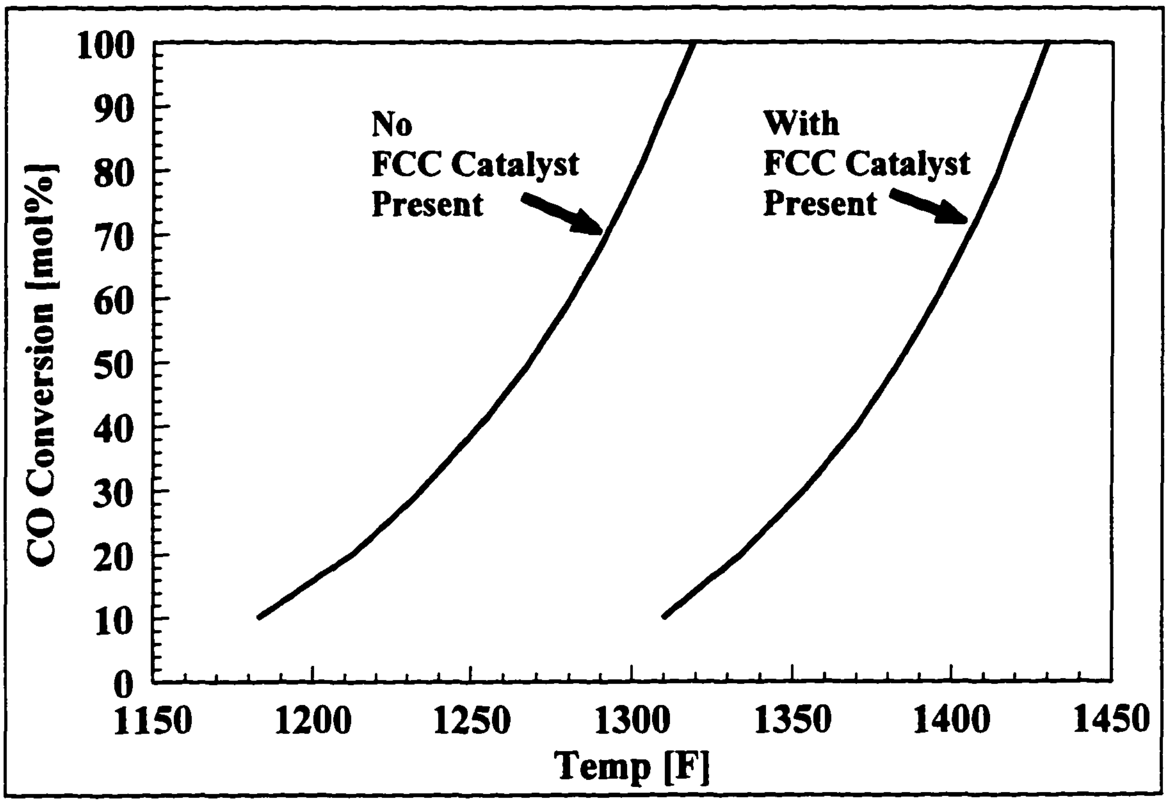


Figure 6: Effect of catalyst on CO burning rate. [From Upson et al. (1993).]

Describing the homogeneous reaction is more complex, as can be seen in Fig. 6 from Upson et al. (1993). It is a free radical reaction which is inhibited by the presence of surfaces such as the catalyst. This can be seen from the fact that when the catalyst is removed in the second cyclone, the gas ignites if both O_2 and CO are present. This effect, known as afterburn, is important in control. The degree of inhibition is a function of temperature and catalyst size and concentration. It is hard to measure accurately in the laboratory due to the controlling effect of reactor surfaces in small isothermal reactors. Krishna and Parkins (1985) did give some data for the combustion rates, but not enough details to compare their results to ours. The rate equation reported by Avidan and Shinnar (1990) is used.

Since the coke combustion and the CO oxidation are competitive reactions (all compete for the same oxygen), the relative rates are effected not only by temperature but also by the level of coke on the catalyst as well as catalyst density and CO concentration in the gas. It is important therefore that the model describes these effects. As Shinnar (1978) pointed out, relying on empirical correlations is risky as these are only valid over limited ranges. It is always recommended to use a consistent kinetic description even if it is simplified and not completely accurate. As the amount of coke per unit volume is substantially different in the bed and in the dilute phase above it, it is important to describe both regions separately. The relations from King (1989) to predict the dense bed density and from McFarlane's model to predict the dilute phase density and elutriation rates are used.

However, the solid phase in dense bed is described as a stirred tank, the gas flowing through it as a series of three stirred tanks, and the dilute phase in an analogous way. This approximates back-mixing. As the residence time in the dilute phase (both for catalyst and gas phase) is very short compared to the response time of the system, quasi steady state in the dilute phase can be assumed which simplifies the dynamic analysis.

An adjustable heat loss term has also been added to the model.

3.4.2. Reactor

Present reactors operating with modern active catalysts are riser reactors with relatively short residence times. As this time is much shorter (2-5 seconds) compared to the response time of a regenerator (one half to an hour), one can at any instance describe the riser reactor by a set of steady state relations, which again simplifies the dynamic analysis.

It assumed to be an adiabatic plug flow reactor with a radically uniform temperature distribution. As at the present stage thermal reactions are not included, the impact of slip can be taken into account by changes in catalyst residence time or, in other words, catalyst inventory in the riser. Assuming uniform temperature at any cross section is a stronger and less realistic assumption, especially in the mixing section at the bottom of the riser, but still leads to useful results.

The main impact that the riser has on both the dynamic and steady state behavior of the total system is in the production of coke and in the removal of heat from the system. The latter is a function of conversion and gas yield or, in other words, the composition of the product stream. For steady state control and optimization of the system, prediction of product composition is the main goal. But it is much less critical for predicting the operating conditions of the system itself, as the only impact is due to the heat balance, which is affected by conversion but is not very sensitive to actual product composition.

As the interest is in steady state control and reactor performance, a ten lump model as suggested by Jacobs et al. (1976) is used. The constants have been updated based on more recent results from Sapre and Leib (1991) and can vary from catalyst to catalyst. Users can substitute values for the reaction rates based on their own data for a specific catalyst. The model is also configured such that a more detailed kinetic model can easily be substituted.

In generalized terms, the rate equations for cracking can be written (see Appendix for details):

$$\frac{dy_i}{dh} = \tau_r A \rho_c \frac{(1-\epsilon)}{\epsilon} \phi \frac{1}{1+k_h y_{Rh}} \sum_{k=1}^{9} a_{ik} y_k K_{k0} e^{-\frac{E_k}{RT}} \tag{6}$$

The term describing the catalytic cracking rate can be divided into three terms. A rate constant $K_{oi}e^{Ei/RT}$ for the lump i, and two constants describing the catalyst activity. A is a constant describing the catalyst activity of the clean burned catalyst relative to the base case included in K_{oi}. It is assumed that K_{oi} does not change as catalyst ages, and overall activity changes can be represented by a single factor. ϕ is a factor given the impact of coke on catalyst on the cracking rates. It is assumed to be $C^{1-1/b}$, where C is coke on catalyst and b is a function of the catalyst type and varies from 0.2-0.5. Thus, in this model catalyst activity is infinite for a completely clean catalyst [Krambeck (1991)].

The coking rate is given by the overall rate equation (based on a Voorhies relation, Krambeck (1991)):

$$\frac{dC}{dh} = \tau_r A z k_{cc0} e^{-\frac{E_{cc}}{RT}} \left(\frac{\Psi}{100}\right)^{\frac{1}{b}} b C^{1-\frac{1}{b}} \qquad (7)$$

A and ϕ have the same meaning as in Eq. 6. z is a factor characterizing the coke selectivity of the catalyst. The deactivation of the catalyst for coking and cracking are coupled, which leads to the results described by Krambeck, i.e., the amount of coke produced for a given degree of conversion is independent of cat to oil ratio and is solely a function of catalyst state, feed composition and temperature.

Some heavy fuels contain compounds that deposit on the catalyst as coke almost immediately by thermal decomposition. As this coke is not formed at a catalytic site, its effect on activity is much smaller. The feed properties leading to it are characterized by giving a Conradson Carbon Residue (CCR) concentration. It is assumed that a fraction β of the carbon converts to coke and this CCR coke only affects the value of coke entering the stripper. It has no deactivating effect on the rate of coking or cracking which is reasonable for small levels of CCR (less than 5% in feed or even more for modern catalysts).

This model includes a very simplified stripper which is assumed to be a well stirred tank. The amount of unstripped hydrocarbon, γ (wt/wt cat), is solely a function of steam to catalyst ratio and is given as:

$$\gamma = 0.0002 + 0.0018(1 - ks) \qquad (8)$$

This assumes an unstripped content of 0.2% (wt/wt cat) of hydrocarbons, and a maximum stripped content of 0.02% (wt/wt cat). These numbers can be adjusted by the user. ***s*** is steam to the stripper, and ***k*** is a constant that characterizes the impact of the steam to catalyst ratio and is a function of stripper design. The effect of holdup on the stripper efficiency has been neglected. But as the holdup is significant, its impact on the dynamics of the system is included.

3.4.3. Afterburn

For control of the regenerator, it is possible to use the air rate to control the afterburn in the dilute phase and in the cyclones. This was a common practice in many FCC's, especially when it was important to prevent damage to the cyclones. The afterburn can be measured as the temperature rise from the dense bed to the entrance to the cyclones, from the dense bed to the flue gas, or across the cyclones only. Following other control papers the second option which gives the most reliable measurement will be designated as ΔT in this work. The other combinations were and can be used for control as well. All three values are computed in this model. They give a measurement of oxygen conversion in both the dense and dilute phase. But in partial CO combustion this conversion is very high and hard to predict accurately. Therefore it is hard to model ΔT exactly and it is assumed that any remaining oxygen reacts with any CO available stoichiometrically in the cyclones. This is approximately correct for cyclone temperatures above 1150°F. Below about 1000°F ignition does not occur. The limit for ignition is a function of residence time in the cyclone and gas composition, and no data are available to predict it. Therefore, complete combustion in the cyclones is assumed, but this limits the validity of this prediction to regenerator temperatures above about 1120°F. As ΔT is not a state variable, this has no impact on the steady states or on the dynamic behavior of the FCC. It enters the equations only if used in a control structure.

3.5. Model Properties and Comparison with Previous Models

This Section concentrates on some overall properties of the model, and on the difference in comparison to previous models.

One way to look at the properties of the model is, at given feed and catalyst properties, to study the impact of catalyst flow rate and air flow rate keeping all other inputs constant. This is a multi-dimensional plot and therefore only cross sections are used.

Fig. 7 shows the riser top temperature (T_{ris}) and the regenerator temperature (T_{rgn}) as functions of catalyst circulation rate and of air rate. Both these are normalized to the oil feed rate (Cat/Oil and Air/Oil). In one case Air/Oil is kept constant while Cat/Oil varies and in the other Cat/Oil is fixed while Air/Oil varies. For each set of inputs there are three steady states, this being an adiabatic unit. Only the two higher temperature states are shown, as the lowest one is outside the range of interest for operation. The stability of these states and the multiplicity problems are be discussed in the following sections. It is shown that for this model and in the range of interest there are always one or three steady states. If there is one steady state, it is a trivial state as the unit remains cold. If there are three steady states, the steady state with the highest regenerator temperature is stable and the intermediate one unstable. A detailed investigation of

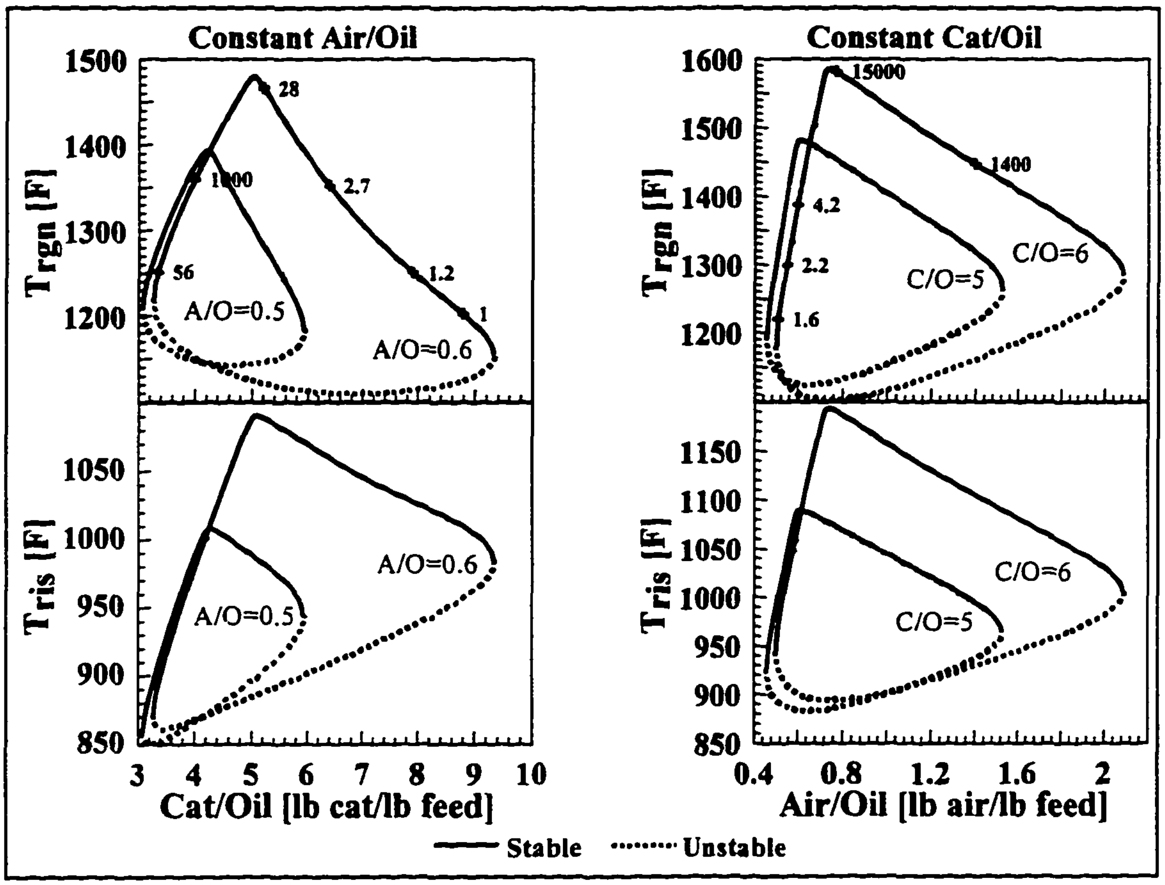


Figure 7: Cross-cuts of T_{rgn} and T_{ris} at constant Air/Oil and Cat/Oil. (Number in graphs are CO_2/CO at cyclones inlet.)

dynamic behavior as well as stability of the steady states is essential for control.

For any operating condition (feed properties, catalyst properties, feed temperature, etc.) this model will always give a plot similar in overall form to Fig. 7. Different operating conditions will just move it and change the slopes. The upper part of the oval represents the high temperature stable steady state. The lower part of the oval represents the intermediate unstable steady state. The low temperature stable steady state is not plotted as it has no operational meaning. Note that for a given Air/Oil there is a maximum and minimum Cat/Oil above and below which there is only a trivial steady state. The same holds for Air/Oil in the case where Cat/Oil is constant.

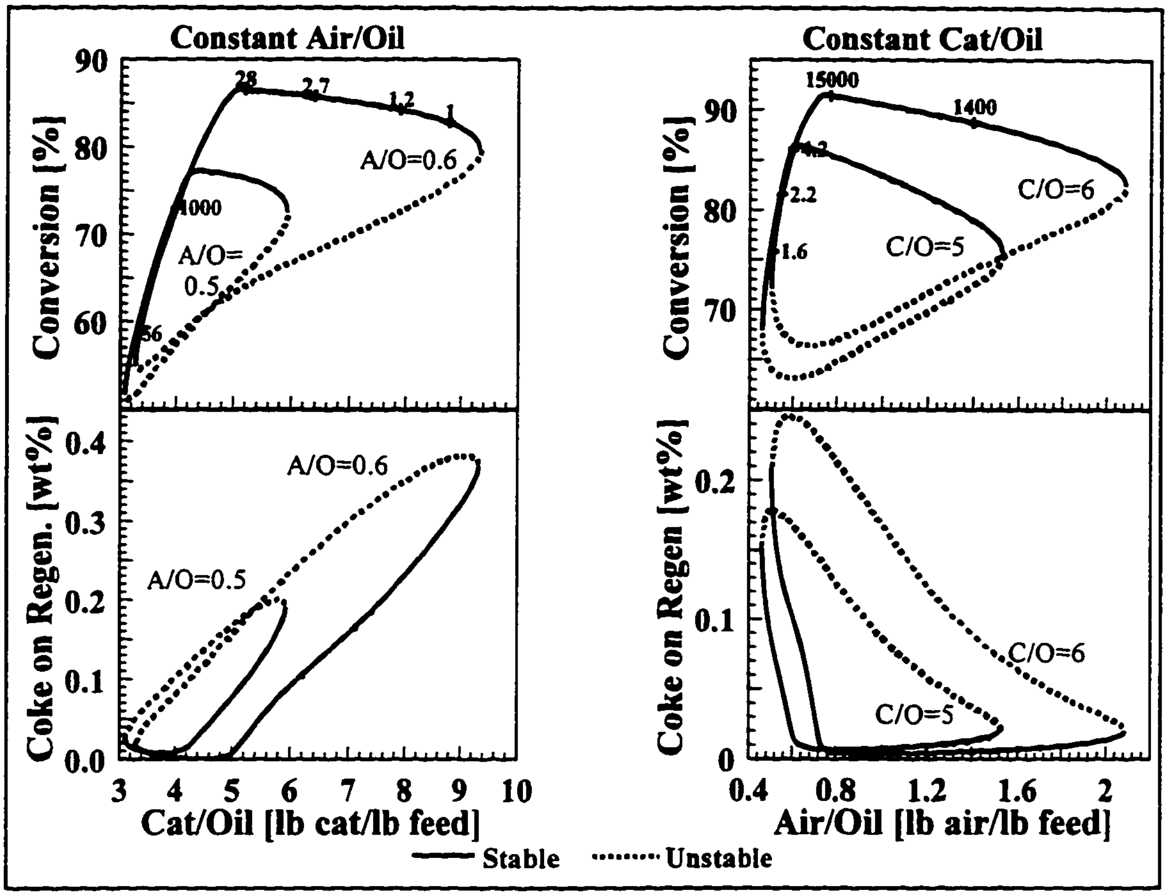


Figure 8: Cross-cuts of Conversion and C_{rgc} at constant Cat/Oil and Air/Oil.

Fig. 7 also gives the CO_2/CO ratio at selected operating points. The maximum reactor and regenerator temperatures occur in complete CO combustion. At lower Cat/Oil ratios the upper steady state is in complete CO combustion. At high Cat/Oil ratios the unit is in partial combustion. Note that the slope of both riser and reactor temperature with increasing Cat/Oil ratio (at fixed Air/Oil) is positive in full combustion and negative in partial combustion. The opposite holds for increasing Air/Oil at constant Cat/Oil.

The model realistically describes the transition of operating conditions from partial to full combustion. Operators will be more familiar with plant behavior when controlling regenerator temperature with air rate while riser temperature is kept constant. This is a problem in closed loop control to be dealt with later. The present concern is open loop behavior. To better understand the implications of Fig. 7, Fig. 8 shows the corresponding plots for conversion and coke on regenerated catalyst and, Fig. 9, those for gasoline and wet gas yields. Plots like Fig. 7-9 are very useful in understanding the impact of different input parameters (feed temperature and properties, catalyst properties, additives, etc.).

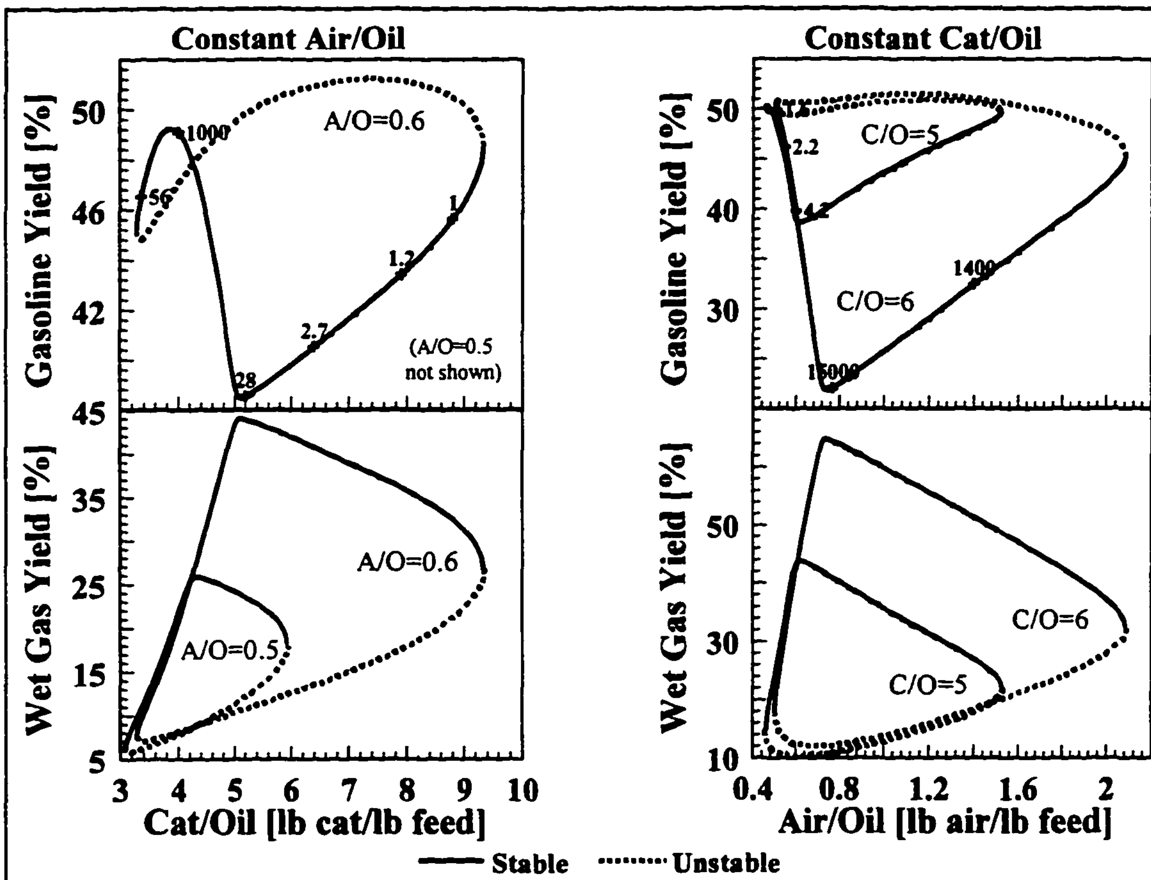


Figure 9: Cross-cuts of gasoline yield and wet gas yield at Air/Oil and Cat/Oil.

In presenting another FCC model it is also important to understand the differences in the results compared to those of previous models. This is rather difficult as many of the papers on modelling present useful approaches but insufficient information on reaction rates and other crucial information to allow any comparison. Several authors have presented regenerator models with a kinetic scheme similar or identical to that presented here [Ford et al. (1976), Krishna and Parkin (1985) de Laza et al. (1981)] but again, they give too little information to allow use of their models. However, some comments on the two main models discussed above are given, models which do allow comparison, namely, that of Lee and Groves (1985) and that of McFarlane et al. (1993).

The regenerator model used by Lee and Groves, that of Errazu et al. (1979), is limited to partial CO combustion up to 1240°F, while the new model permits operating at higher temperatures and is capable of a smooth transition to full CO combustion. However, the results in the range where Lee and Groves' model does apply are quite similar, since the CO_2/CO ratio is practically the same. Thus, for representing the flexibility of modern FCC units the model presented in this paper is superior.

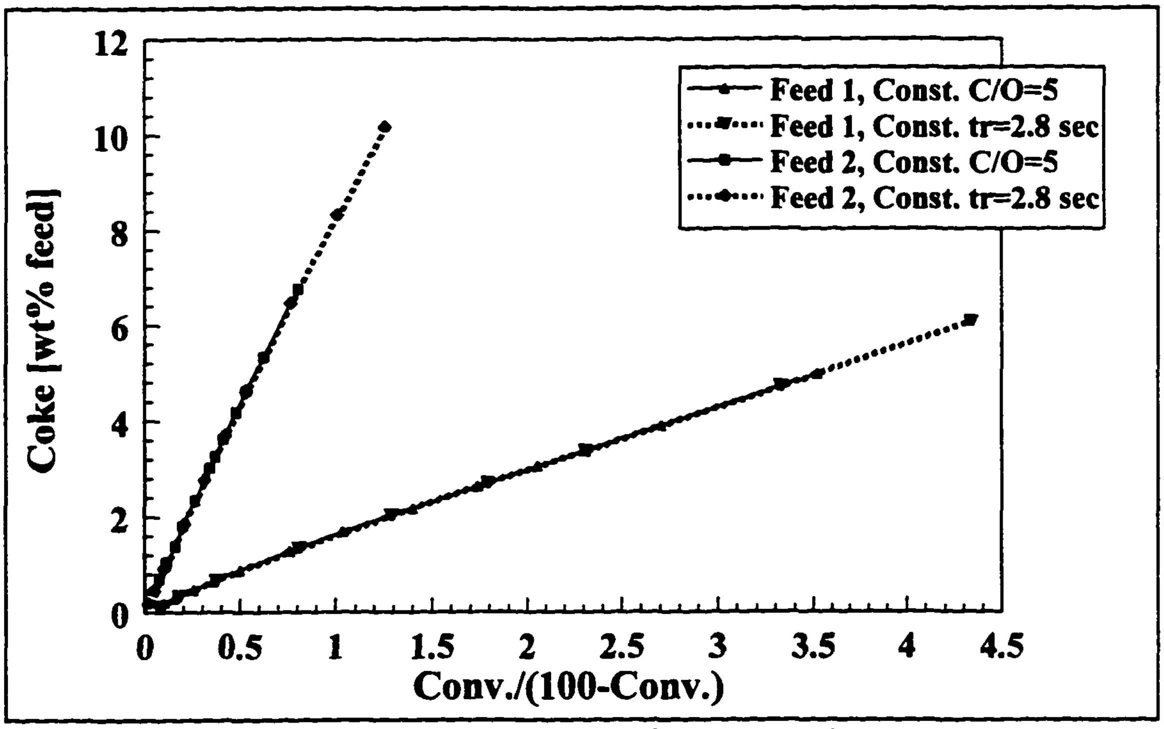


Figure 10: New riser model at T_{ris} = 1000°F. Effect of Cat/Oil, residence time and feedstock.

As to the assumptions used by Lee and Groves for modelling the reactor, it was shown in Fig. 4 that this model is not consistent with the data presented by Krambeck (1991) and Sapre and Leib (1991). The same is true for the ten lump model published by Jacob et al. (1976) and Gross et al. (1976). While a similar scheme is assumed here, the deactivation function was modified based on Krambeck's (1991) suggestion. Fig. 10 shows that this model has the appropriate property which can also be seen from the equations themselves. The model by Lee and Groves coupled with this regenerator model will give a curve similar to Fig. 7. The slopes will however be different and the impact of increasing Cat/Oil or activity will not be properly predicted. Krambeck (1991) gives a similar three lump model which has the proper properties.

It was noted before that the regenerator model given by McFarlane et al. (1993) is based on using a catalytic promoter for the CO to CO_2 reaction, as commonly used in full CO_2 combustion. Fig. 7 is based on an unpromoted catalyst. Fig. 7 shows however a region of full CO combustion. This conforms to experimental experience as full CO combustion was introduced before CO combustion promoters became available. For this regime this model will give similar results. To understand the action of combustion promoters, Fig. 11 shows a constant Air/Oil plot equivalent to that in Fig. 7. (All data in Fig. 11 are based on the model in the appendix.) Fig. 11 gives two levels of promotion. In the model the level of promotion is defined by a ratio of the catalytic combustion rate to a base level in an unpromoted catalyst. This allows direct measurement in a laboratory unit. At the lower level it is still possible to operate at a regenerator temperature of 1250°F in partial combustion. At higher levels of promotion

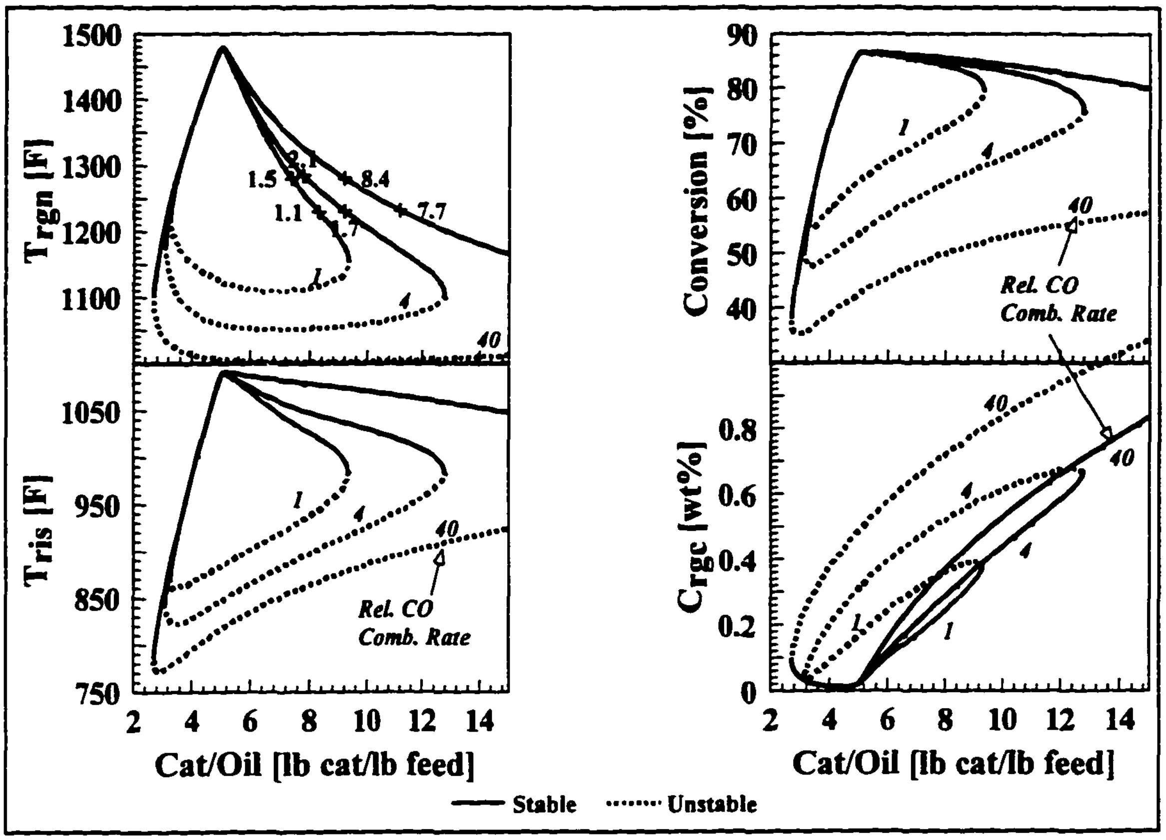


Figure 11: Cross cuts at constant Air/Oil=0.6. Effect of CO combustion promoter. (Numbers in T_{rgn} graph are CO_2/CO at cyclones inlet)

full CO combustion occurs over most of the practical range. Some CO_2/CO ratios are indicated in the plot. Moderate promotion is sometimes used in partial combustion to reduce the impact of flow non-uniformities which may cause excess temperature in a single cyclone. Combustion promoters increase the operable range of Air/Oil and Cat/Oil, and increases stability. The penalty is a lower conversion in the partial CO combustion range as shown in Fig. 11.

The reactor model presented by McFarlane et al. is too oversimplified to enable any comparison. Modelling of the cracking was not one of McFarlane's targets. It is important here, especially in the concept of partial control.

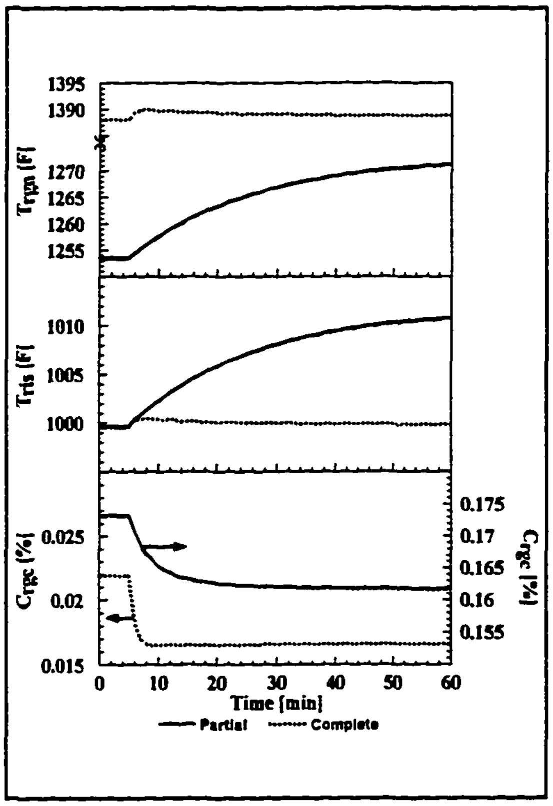


Figure 12: Dynamic response to step change in Air/Oil (+2% in partial combustion;+5% in complete combustion).

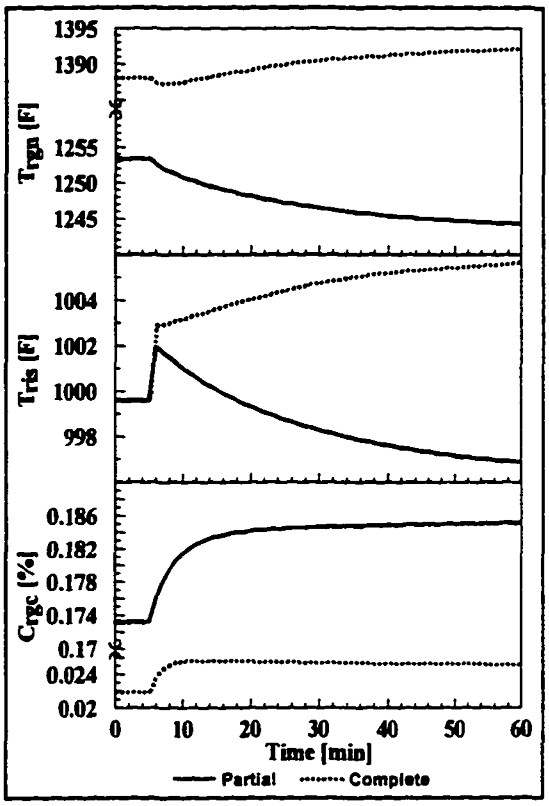


Figure 13: Dynamic response to step change in Cat/Oil (+2%).

Testing the dynamic properties of the model both open loop and under control is an important part of model verification. Here, only open loop dynamic properties are presented, focusing on the important state variables that dominate performance, T_{rgn}, T_{ris}, and C_{rgc}. The results are given in Fig. 12 and 13 for two cases, one in partial CO combustion and one in full CO combustion. First consider the effect of an increase in Air/Oil ratio keeping Cat/Oil constant (Fig. 12). For both conditions T_{ris} and T_{rgn} increase monotonously and C_{rgc} decreases. But the time scales are different by an order of magnitude, that in complete CO combustion being much shorter. Both the magnitude of the time scale and the large difference between complete and partial CO combustion

correspond to experience in commercial units, this
is important if one wants to apply the model to control studies of real cases.

The long time scale of the response in partial combustion (1/2 hour to one hour or 2 to four hours for complete line out) is at first puzzling. The residence time in the riser is measured in seconds. Gas residence time in the regenerator is about 20 seconds. Catalyst residence time is 2-5 minutes. The long response time is partially due to thermal holdup and partially due to the interactions between reactor and regenerator. It is typical of may systems with internal recycles. The coupling agent is C_{rgc}. In full CO combustion, C_{rgc} is negligibly small, which reduces the interaction and thereby reduces the time scale.

The impact of this on steady state control will be discussed in later sections. Fig. 13 gives the analogous impact of an increase in catalyst circulation rate. The main time scale is obviously similar as the overall response is dominated by regenerator behavior and the same coupling effects. However, the model exhibits some inverse responses which are also observed in practice. It is important to understand this difference.

When one supplies more air to the system, one supplies the ability to combust more coke and generate more heat. Therefore the system has the ability to increase both T_{rgn} and T_{ris}. This is true for both partial and complete CO combustion. However when one keeps Air/Oil constant, there is an inherent difference between partial and complete combustion. In partial combustion excess air is very small. If one increases Cat/Oil then at first T_{ris} increases, making more coke. But there is not enough air to sustain a higher coke make and a higher T_{rgn}. The only conditions at which the system will find a new steady state are at lower T_{rgn} and T_{ris}. Lower T_{ris} decreases the coke make which compensates for the increase in Cat/Oil. Lower T_{rgn} decreases the CO_2/CO ratio and thereby increases the amount of coke that can be combusted. This explains both the inverse response in T_{ris} and the ultimate decrease of T_{ris} and T_{rgn}.

Whereas in partial CO combustion both T_{rgn} and T_{ris} will line out at a lower value for an increase in Cat/Oil at constant air, in full CO combustion increasing Cat/Oil increases both T_{rgn} and T_{ris}. In full CO combustion there is excess O_2 (0.6 % for this case). For a small increase in Cat/Oil this allows to burn more coke, and therefore both T_{ris} and T_{rgn} ultimately increase. However, due to the thermal inertia T_{reg} first decreases as more heat is removed by circulating catalyst. For a large increase in Cat/Oil, the response will be nonlinear and the system will go to partial combustion (see Fig. 7). Both T_{rgn} and T_{ris} could decrease reversing the response observed for a small increase.

The same is also true for partial combustion. At constant Cat/Oil an increase in air rate will lead to complete combustion (see Fig. 7). However, the tolerance is larger in partial combustion, as normally operation is further away from the maximum of the plot in Fig. 7. For large changes in Cat/Oil or Air/Oil the response is highly nonlinear for both cases such that the sign of the steady state changes reverses. This is important for steady state control.

The model has the proper time scales and directional dynamic behavior. For studying actual operation of FCC, the complex interactions are sometimes counter-intuitive and have a strong implication for control. It is therefore important to understand them, and here this model should be useful.

3.6. Range of Applicability for the Model

Any kinetic model of the type given here has a limited range of applicability. As the model is based on kinetic equations, it remains valid over a wider range than one based on correlations. Even so, kinetic relations of the type used are only valid over a certain range, also determined by the range of conditions over which the model was obtained.

On the reactor side, the model is valid in the temperature range 850-1050°F. Below 850°F there are no data as most FCC's operate above 900°F. Above 1050°F thermal reactions become important. For very low residence times the model will give reasonable predictions up to 1100°F.

The reactor model allows the prediction of the impact of riser top and bottom temperature, catalyst type and activity, feed composition and C_{rgc} on coke make, conversion, gasoline yield and wet gas yield. The constraints can be obtained from isothermal laboratory experiments. It cannot predict gasoline composition and wet gas composition, which have become important for overall refinery optimization. The reason is a lack of nonproprietary data. The way the model is structured, the user can substitute a more complex lumping scheme either by increasing the number of lumps or using a structure oriented lumping scheme [Quann and Jaffe(1992)]. However no non-proprietary data are available. Alternatively the steady state solution of this model can be used as a basis for correlations based on laboratory data.

Another caveat - the reactor model assumes near plug flow with uniform temperatures across a cross section. The bottom of the riser however involves a nonisothermal zone due to a finite mixing time. This is a strong function of nozzle and reactor design. More complex models would require more information than is available. Still one gets reasonable predictions this way, but one should remember these limitations.

On the regenerator side, the model is valid at all temperatures above 1000°F. However, afterburn may not be used (for control) at any temperature below 1120°F, as the CO will no longer self ignite. The actual limit depends on cyclone design and gas composition. Above 1500°F regenerator temperature, the zeolite catalyst may deactivate rather rapidly. As a result predicting steady state operating conditions at higher regenerator temperature is rather meaningless.

The model is, however, valid over the whole range of operating conditions used in current FCC operation both in partial and full combustion. There is one problem in the results as presented, especially when dealing with operating conditions in open loop such as Fig. 7. In such a figure the unstable intermediate steady state may be in a range where the model is not applicable. In this case it cannot accurately predict this unstable steady state. For other operating conditions the same may hold for the stable steady state. This is not important, as there is no chance that an operator in closed loop would choose such a setting. From a dynamic point of view it still gives a valid indication of how close the stable point is to unstable conditions. Analogously, the stable upper point in Fig. 7 could be at conditions outside the model range limits. While this model still predicts the slope correctly, it will not be useful for predicting operating points outside of this range.

There is another limitation that should be pointed out. Fluid beds involve complex hydrodynamics that also depend on unit design. One has to understand the price one pays for making the simplifications. This model predicts reasonably well regenerator temperature, C_{rgc} and CO_2/CO ratio. It is rather inaccurate in predicting T_{sg} and therefore ΔT, the predictions being better in some operating ranges than others. ΔT depends strongly on oxygen conversion. A difference between 99.5% conversion and 99.0% conversion doubles ΔT. There are no models for an unbaffled fluid bed reactor that are that accurate. However, trends of T_{sg} are predicted quite well. More complex models are of no help. A difference in oxygen conversion of 1% or even 2% has very little impact on T_{rgn} or T_{ris} which are therefore more accurately predicted. However, even here this model will work well only for the limited range of conditions where most FCC's operation. It will break down for lower gas velocities, shallow beds, and nonconventional catalysts with different fluidization behaviors. The difference in accuracy in predicting T_{sg} or ΔT versus T_{rgn} impacts the choice of the control structure as will be discussed in the follwoing sections.

4. Nonlinearities and Stability in FCC

4.1. Introduction

The premise is that a knowledge of the steady-state behavior of the system is a necessary prerequisite to the design of its control system. That knowledge is based on the operating map which shows the relationship of important state variables to the key manipulated variables over the feasible range of operation. For nonlinear systems the operating map will reveal the existence of multiple steady states and input multiplicities, both of which must be taken into account in the design of the controller. In the case of chemical reactors, for instance, linear controllers are generally adequate for dynamic control around a specified steady state, providing that this steady state is properly chosen. The nonlinearities must be considered because changing set points can radically alter the linearized transfer functions and the stability of the system.

Both the existence and stability of multiple steady states are explored over the range of current operating conditions for FCC's. This is done first for a base case representative of normal operation. Next the impact of design as well as the effects of feed composition, catalyst properties, and combustion promoters are studied. Finally, despite all efforts to develop an accurate model of the FCC, the effects of the many remaining uncertainties are considered.

The exploration of multiple steady states and input multiplicities requires choosing a set of manipulated and controlled variables. For the present study, those control strategies most commonly used in industry and report in the literature have been used.

There is a large and valuable literature on multiple steady states and bifurcation analysis. Most of it deals with a rigorous theoretical analysis of some generalized examples. Very few papers deal with detailed industrial examples. The purpose of this section is to show how to approach such a problem in actual design, what information is crucial , and how to deal with model uncertainties.

There is also a substantial literature on the subject of stability and the existence of multiple steady states in fluid beds [Edwards and Kim (1988), Elnashaie and El-Hennawi(1979), Arandes and de-Lasa (1992), Elshishini and Elnashaie (1990), Iscol (1970), Lee and Kugelman (1973), Seko, Tone and Otake (1982)]. None of these issues have been addressed in a thorough theoretical way. Nor has the impacts of design, operating conditions, and catalysts on the existence of multiple steady states and stability been considered. In some papers [Edwards and Kim(1988) and Arandes and de-Lasa(1992)] the term multiple steady states is applied to input multiplicities. Clarification of these matters is one of the goals of this section.

The section is divided into four parts, each dealing with an important aspect the designer has to face and investigate. In Section 4.2 the FCC and its features are described in brief. Section 4.3 provides an evaluation of the number and nature of the steady states. First, the behavior of the base case is explored. Then the impacts of model uncertainties , of new catalysts, and of operating conditions are considered. The potential input multiplicities which dictate the ultimate choice of the control matrix are analyzed in Section 4.4. The problem from the viewpoint of the operator is discussed in Section 4.5. The operator chooses set points for the control loops and then changes these set points to compensate for changes in inputs or in the desired output specifications. One needs to know (1) what combinations of set points are permissible and (2) how likely a given set point combination corresponds to an unstable steady state.

It will later be shown that in design of controllers for nonlinear systems, the analysis outlined in this section is essential even if the design of the control loops themselves can be based on locally linearized design methods.

4.2. Outline of the Problem

The problem in general is that of exploring the steady-state behavior of a complex, highly nonlinear system in such a way that its impact on operations and control becomes clear. The FCC is used as an example for the approach to this problem. The strength of this approach is that it is based on a system which is both realistic and of practical interest. The results, in addition to being illustrative of the approach, are also of direct utility. However, every system of the complexity of FCC has its idiosyncracies which complicate generalization. The drawback of the this approach, if there is one, is that some of the results may appear to be directly related to these idiosyncracies and of little general value. The intent is to minimize this aspect of the problem. However some of these details must be included to fully understand the problem and to lay the basis for the subsequent parts in this work.

All FCC's are autothermic and require heating to start up the system. As a result they exhibit multiple steady states just like any combustor or autothermic reactor [Aris (1965), Balakotaiah and Luss (1984)]. In current designs, there is a cold steady state that is always unconditionally stable, since present FCC's lack facilities to preheat the feed sufficiently to have a single hot steady state. Because the feed can not be sufficiently preheated to achieve self-ignition, a start-up burner is needed. Therefore, if there is only one stable steady state, this has to be the cold one. Whenever there is a useful hot steady state, there must be at least three steady states, the middle of which is always open loop unstable [Aris (1965), Balakotaiah and Luss (1984)]. An interesting question arises. Is this intermediate unstable steady state in an operating range that can be selected by the operator? It is therefore important to know for what operating conditions the hot steady state is stable and how changes in process conditions affect stability. It is quite possible for an operator to change operating conditions in such a way that the upper steady state is lost. The unit may wind down to the cold steady state or else reach an unstable steady state that needs stabilization by a control circuit.

Actually there is no practical cold steady state in an FCC. Although one can compute a cold state, the temperature is so low that the feed will not be vaporized. There will be no reaction and no catalyst flow. The unit will merely fill up with unvaporized feed. To prevent this, an operator must intervene and shut down the unit. In this sense, talking about a cold steady state is simply a convenient convention. The equations show an attractor to which the system moves. If the system enters that region, it cannot return without intervention. The fact that the model equations fail to hold in that region is not important. This is not an isolated example. In many reaction systems one can compute a high-temperature steady state which is also physically undesirable. For example, the catalyst may sinter or, if the temperature is high enough, the reactor may melt. However, it is convenient to refer to these impractical steady-states. In the case of the FCC, the low temperature steady state will be referred to as the cold steady state.

Even the simplest autothermic reactor has the potential to exhibit three steady-states. However, when multiple steady states in FCC operation are discussed in the literature, the reference is not to this basic question of three steady states. It is, rather, to the existence of multiple steady states in the high temperature regime, i.e., the existence of more than three steady states. About this there is some confusion. For example, Lee and Kugelman (1973) discuss the stability of FCC reactors, showing that for their operating conditions the system is linearly stable. They talk about multiple steady states and give an example for one unstable steady state without pointing out that this is simply the intermediate unstable steady state that exists for all FCC operations. Their example has a conventional stable steady state at higher riser and regenerator temperatures. The results presented here show that operation at an open loop unstable steady state can occur at reasonable operating conditions. Furthermore, at the stability limits, the unstable and the stable steady states converge.

There is another confusion due to nomenclature. In referring to multiple steady states, a reaction engineer normally assumes all inputs to be constant. Edwards and Kim (1988) as well as Arandes and de-Lasa (1992) assume that riser top temperature is kept constant by varying the catalyst flow and obtain two steady states at the same air flow. This involves a control circuit and is what a control engineer would consider an input multiplicity.

It is also quite obvious that a system of the complexity of the FCC, given arbitrary kinetics, has the potential for more than three steady states. The regenerator has consecutive reactions. This alone has a potential for five steady states. Interactions with complex coking kinetics could increase this number to an arbitrary level. Furthermore, some of these states may have complex stable limit cycles. However, the authors' model does not show this behavior, at least when using the constants given in the Appendix.

4.3. Multiple Steady States: Nature and Number

Both in design of chemical reactors and their operation, it is important to understand the topology of any multiple steady states or other instabilities. A potential approach to this problem (useful also for other systems) is illustrated using the FCC as an example. When the model is known, e.g., the model given in the appendix, this is straightforward. Keep in mind that not all the uncertainties of the model have been resolved. This and the changes that could occur as new catalysts are developed must be taken into account.

4.3.1. Base Case Behavior

The behavior of the system using the base case parameter values for the model given in the appendix are examined first. A typical plot of heat generation versus heat removal is shown in Fig. 14. The full properties for the two high temperature steady states are given in Table 2. Fig. 14 is very similar to the case of a simple exothermic CSTR with or without external heat removal. A linearized stability analysis shows (see Table 2) that analogous to the single exothermic reaction, the upper steady state is stable and the intermediate unstable. In present FCC's, the lower steady state always occurs when there is no self-ignition and it is linearly stable. It is well-known that in a system with three steady states, the intermediate is always linearly unstable. Only the upper steady state can have a more complex bifurcation structure and can be stable or unstable. All comments on stability in the following will therefore relate only to the upper steady state.

An FCC is much more complex than a CSTR. To plot Fig. 14, all the inputs

Table 2: Steady state data for the upper and middle steady states in Fig. 14.
Constant inputs: Cat/Oil=6.45, Air/Oil=0.545, T_{feed}=670°F

Variable	Upper Steady State	Middle Steady State
Riser Top Temperature, T_{ris}, [°F]	1000	924
Regenerator Dense Bed Temperature, T_{rgn} [°F]	1253	1134
Stack Gas Temperature, T_{sg}, [°F]	1247	1993
Temperature Rise, T_{sg}-T_{rgn} [°F]	44	859
Oxygen in Stack Gas, $O_{2,sg}$, [%]	0	0
CO_2/CO at Cyclones Inlet	1.6	1.2
Coke on Regenerated Catalyst, C_{rgc}, [wt%]	0.173	0.266
Coke on Spent Catalyst, C_{sc}, [wt%]	0.924	0.932
Conversion [%]	80.3	70.2
Gasoline Yield, Y_g [%]	47.5	51.0
Eigenvalues [1/min]	-0.642	-0.612
Note: The five eigenvalues are for the five state variables: C_{sc}, T_{st}, C_{rgc}, T_{rgn}, T_{wall}. See Appendix for details.	-0.563	-0.586
	-0.279	-0.262
	-0.051	0.034
	-0.017	-0.019

were fixed. Then, for each regenerator temperature, the reactor was treated adiabatically and the unit was mass balanced taking into account all the reactions. The heat balance shown in Fig. 14 is centered on the regenerator. The heat removal line is based on the difference in the heat content between the flue gas and the air, as well as between the catalyst flow from the regenerator to the reactor and back to the regenerator. The heat generated is simply the sum of all of the heats of combustion. A breakdown of the

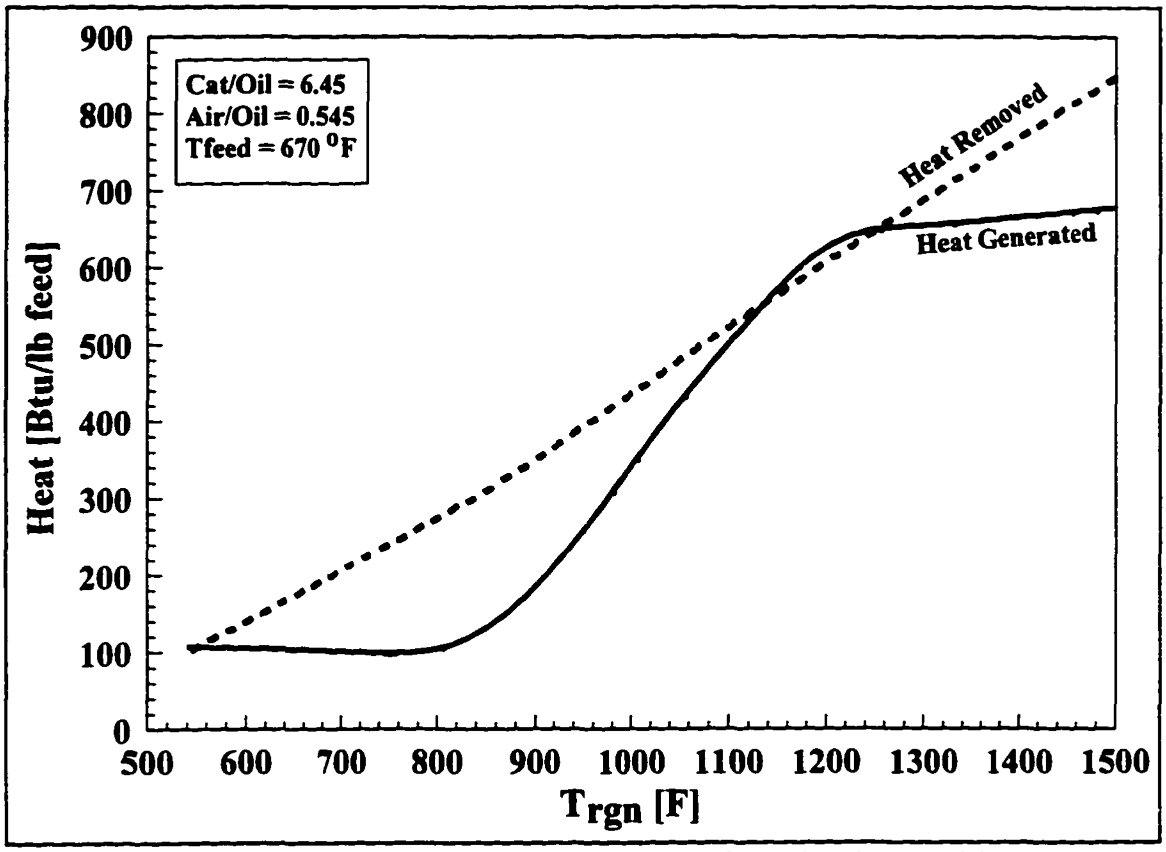


Figure 14: Heat generation and heat removal in partial combustion. All inputs are constant.

different elements of the heat balance is given in Fig. 15. Total heat removal is almost linearly proportional to the regenerator temperature (the slope is a function of catalyst and feed properties and preheat). This is due to the fact that the endothermic heat of the overall cracking reaction is significantly less than needed to vaporize the feed. The complex equations of the riser can be approximated for the purpose of the heat balance as being proportional to the regenerator temperature. This produces an exothermic combustor with consecutive reactions and heat removal. This is somewhat oversimplified since coke generation is not constant, but is a useful simplification for discussion.

The method in plotting Fig. 14 assumes implicitly that for a given ratio of catalyst circulation rate to oil feed rate (Cat/Oil) and given regenerator and feed temperatures,

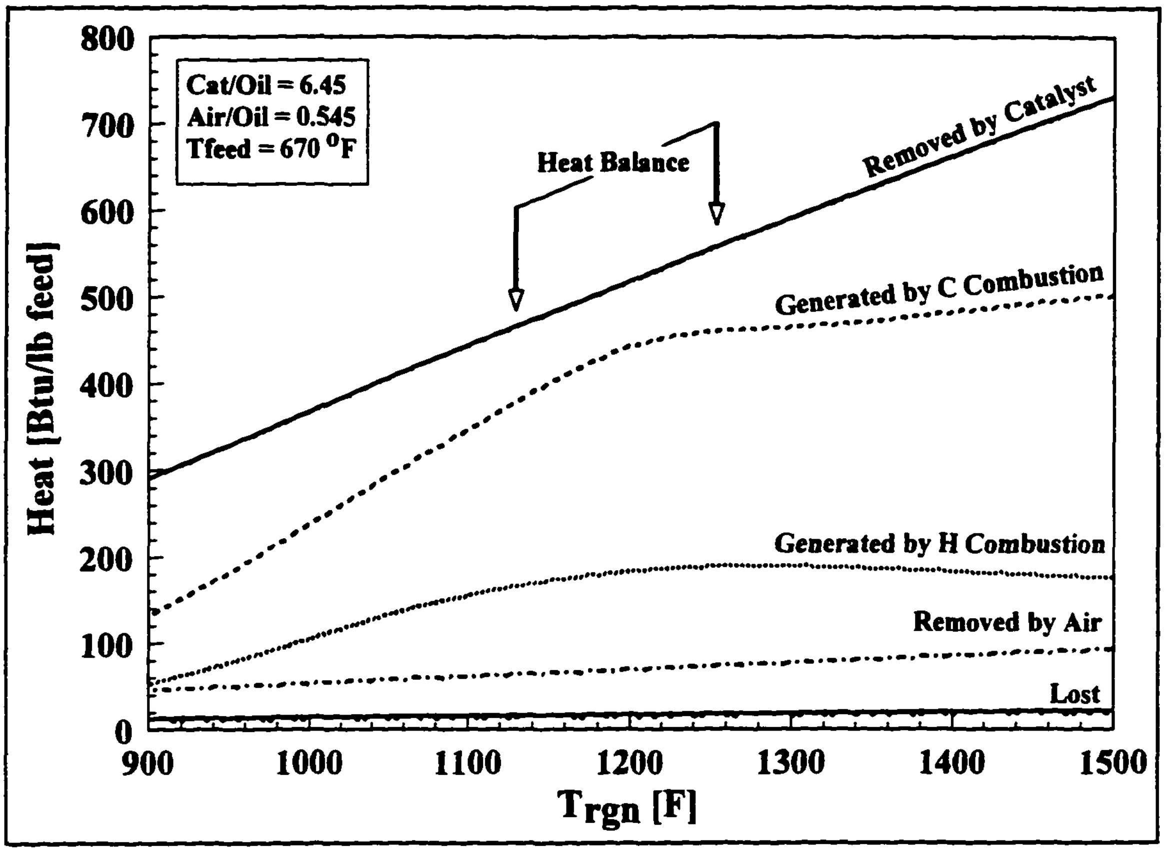


Figure 15: Heat generation and heat removal in partial combustion. Breakdown by source of Fig. 14.

the riser reactor itself has a unique stable steady state. For this model, this is true for all conditions.

The upper steady state in Fig. 14 is linearly stable. The same approach (stability analysis based on linearization and determining all the eigenvalues) was applied to a wide range of operating conditions, including different feed stocks and catalyst properties. It was found that there are always either one steady state or three steady states. In the case of three steady states, the upper is always linearly stable. It should be pointed out, however, that this investigation was limited to reasonable operating conditions and parameter changes. For a simple exothermic CSTR it is a necessary but not a sufficient condition for stability that the slope of the heat removal line at the intersection is larger than that of the heat generation curve. In the FCC case the slope condition works but in other complex systems, the outcome may not always be the same.

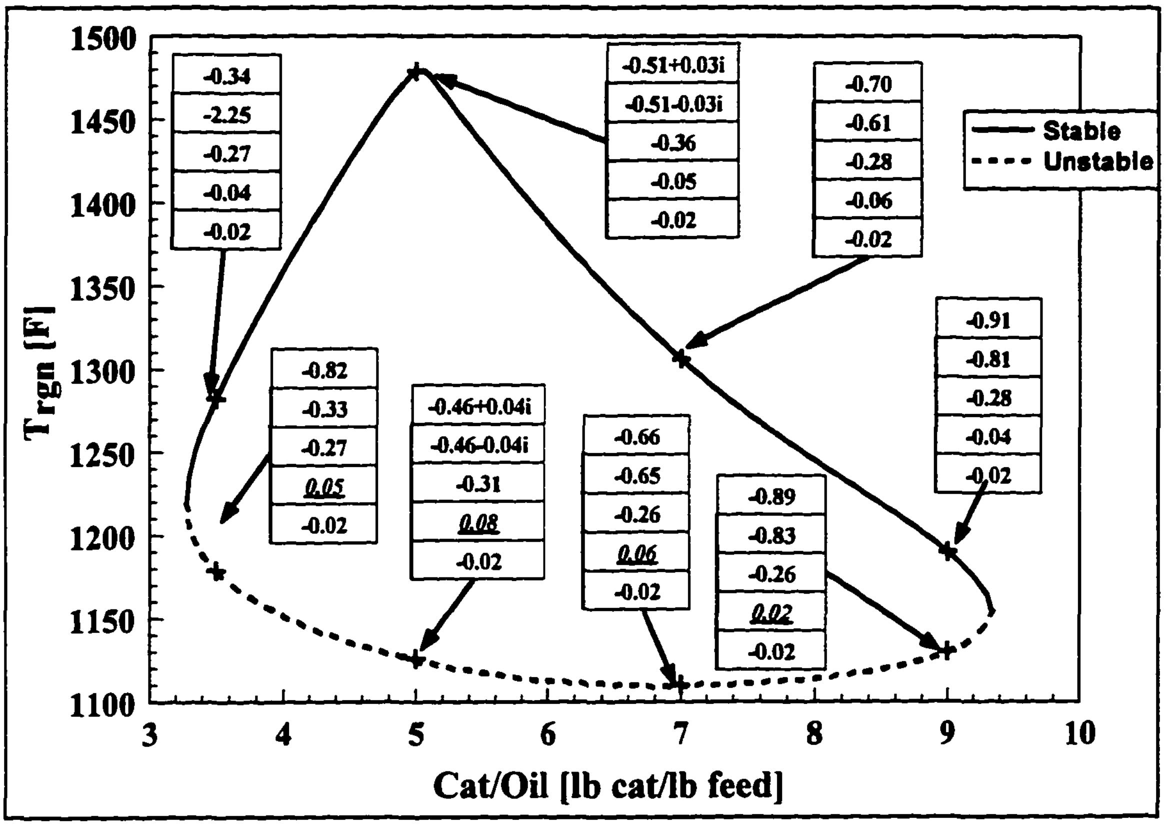


Figure 16: Eigenvalues for some points along a cross-cut at constant Air/Oil =0.6.

Next, the problem of stability is examined in some more details. In Fig. 16 are given some typical sets of eigenvalues of the state space matrices for the different steady states described in Fig. 7. To obtain the results in Fig. 16, twenty-four points along the curve were checked. The lower steady state is not shown in Fig. 16. There is one interesting feature in Fig. 16 that also applies to the other cases studied. Over almost all of the range of the conditions in Fig. 16, all of the eigenvalues are real. Close to the value of Cat/Oil where either the maximum or the minimum reactor temperature occurs, two eigenvalues merge into a complex pair that always has a negative real part. In both cases, however, the FCC would not be controllable by measuring reactor temperature and adjusting catalyst flow rate because of the zero process gain. This will be considered in more detail in the next section which deals with control.

The situation may be quite different in closed loop control. Limit cycles may occur for even the simplest CSTR [Aris (1965)]. Others have shown a rich structure of bifurcations, period doubling and chaos for such reactors [Giona and Palidino (1994), Pellegrini and Biardi (1990)].

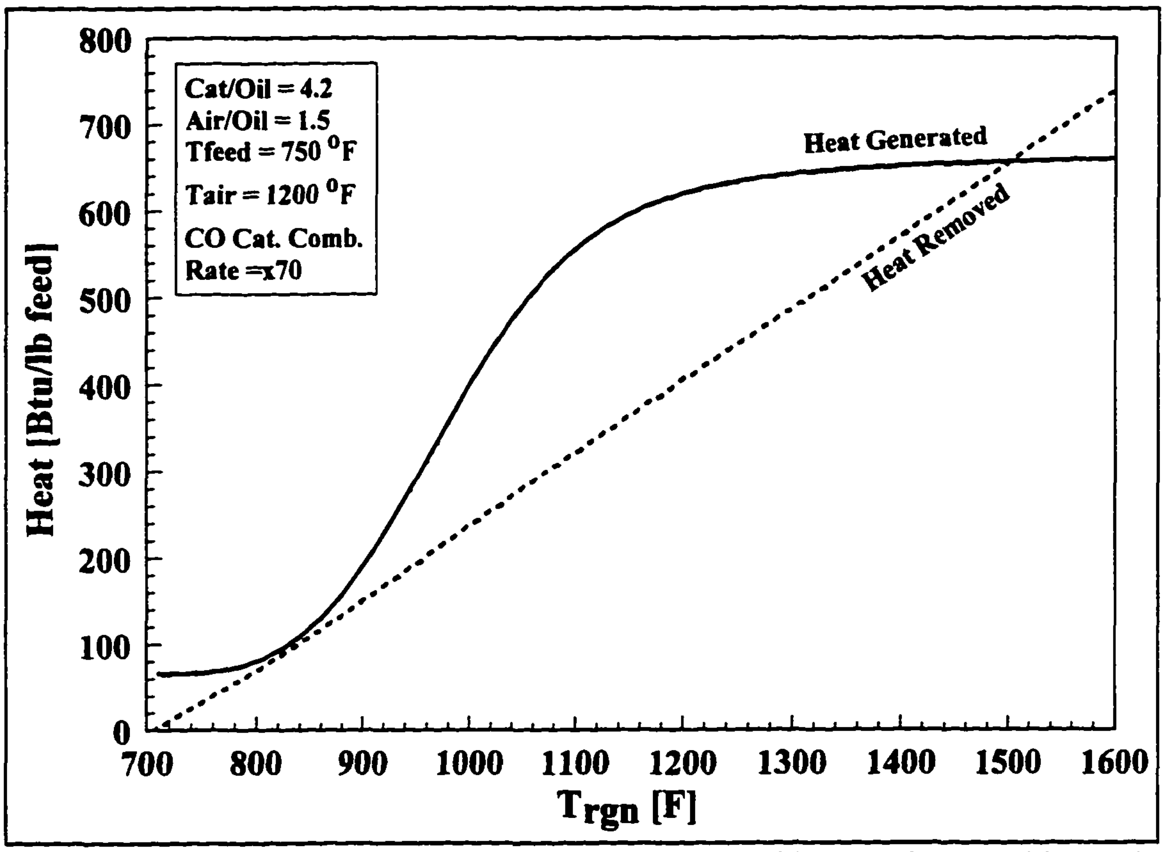


Figure 17: Heat generation and heat removal for a case with air preheat resulting with only high temperature stead state. All inputs are constant.

Fig. 7 shows clearly that there are operating conditions in terms of Air/Oil (Air flow rate to Oil feed rate) ratio and Cat/Oil ratio for which the FCC, with the given inputs, has no hot steady states. This is expected and predicting these conditions is of great interest to the operator. It was argued in Section 4.2 above that the FCC cannot have a single hot steady state due to limitations on preheating the feed. However it is possible to preheat the air to 1200°F or even higher. A typical example is given in Fig. 17. Such a practice is not economical as is evidenced by the fact that no FCC's have air preheaters.

To summarize, this model has a rather simple operating topology, at least in the range of practical operating conditions. There is either one or there are three steady states. The middle one is always linearly unstable and the upper one always stable. The range of existence of these steady states and the permissible operating space are explored in more detail later in this paper.

4.3.2. Effects of Uncertainties and New Catalysts

Next, the uncertainties of the model and the impact of new catalysts are considered, starting with the possibility of five steady states. It is important to know if there are five steady states, since it is more likely for the operator to move to an unstable steady state then if there are only three steady states.. Normally operation in partial combustion at high regeneration temperatures is stable; this is not the case should there be five steady states.

In this model with arbitrary kinetic constraints, there is a potential for five steady states as there are two consecutive exothermic reactions in the regenerator (C --> CO --> CO_2). An example is given in Fig. 18. The catalytic burning rate of CO was decreased and the activation energy of the homogeneous combustion increased to 210 kcal/gmole, the pre-exponential term being reduced so that the rates are equal at 1250°F. This gives five steady states. However, the third one at 1150°F has a significantly lower CO_2/CO

ratio than is normally observed, 0.44 instead of 1.0 t o 1.1. Again, a linear stability analysis shows that the third and fifth states are stable.

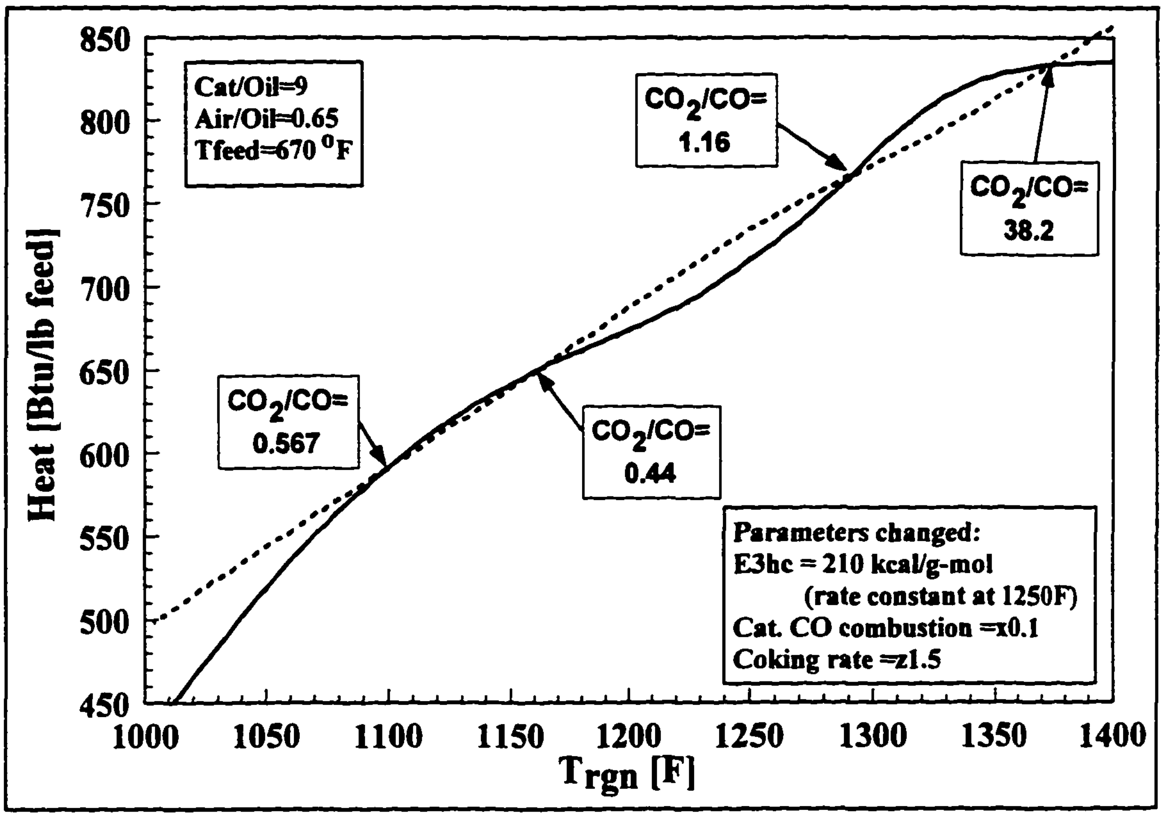


Figure 18: Heat generation and heat removal with parameters changed leading to five steady states. (Low temperature steady state not shown.)

While these kinetic constants predict normal operating conditions quite well, there is at least one large uncertainty. Homogeneous combustion of CO to CO_2 in the presence of solids is a complex and not well understood process. At temperatures below 1300°F, the presence of catalyst particles with large surfaces inhibits the reactor by inhibiting free radical propagation. At a sufficiently high temperature this inhibition ceases. This was shown by Upson et al. (1993) and the data were reproduced in Fig. 6. The inhibition in Upson's case ceased at 1400°F, but the reduction in the efficiency of inhibition is a function of catalyst loading and reactant concentration and is hard to measure experimentally. For modelling today's FCC operation, this is not very important as operation in partial combustion is mostly below 1300°F. In full combustion catalytic CO combustion promoters are used. Thus, a higher homogeneous combustion rate would have no impact.

In order to have five steady states a takeoff in the CO to CO_2 combustion rate must take place at a temperature higher than the temperature of the third steady state. In Upson's data the takeoff occurs at about 1400°F. In Fig. 18 the takeoff is in the 1300°F range, but there is no data to support the takeoff at lower temperatures. Further, to lead to five steady states, the takeoff has to occur in the dense bed because there is not enough oxygen left in the upper dilute phase for it to occur there. Transferring Upson's takeoff to the 1300°F's range would result in five steady states as shown in Fig. 19. These five steady states were obtained using Upson's data while maintaining the kinetic constants for the other combustion reactions that result with the observed CO_2/CO ratio at low temperatures (the conditions of Fig. 14 do not give five steady states even with this modification).

Unlike the examples in Fig. 18 and 19, the conditions of Fig. 14 did not lead to a second inflection point. Other conditions in this model give an inflection point but the slope is not steep enough to allow for five intersections. The possibility of a pronounced inflection point (in five steady states) is, in this case, more limited than in the general case of consecutive reactions such as in a hydrocracker. In partial combustion, present units operate with very high oxygen conversion in the regenerator (above 95%). Otherwise ΔT will be excessive. Thus, at the third stable steady state, there is only small excess of oxygen. Otherwise an upper steady state would be achieved simply by combusting more oxygen. With present catalysts this is unlikely because a significant

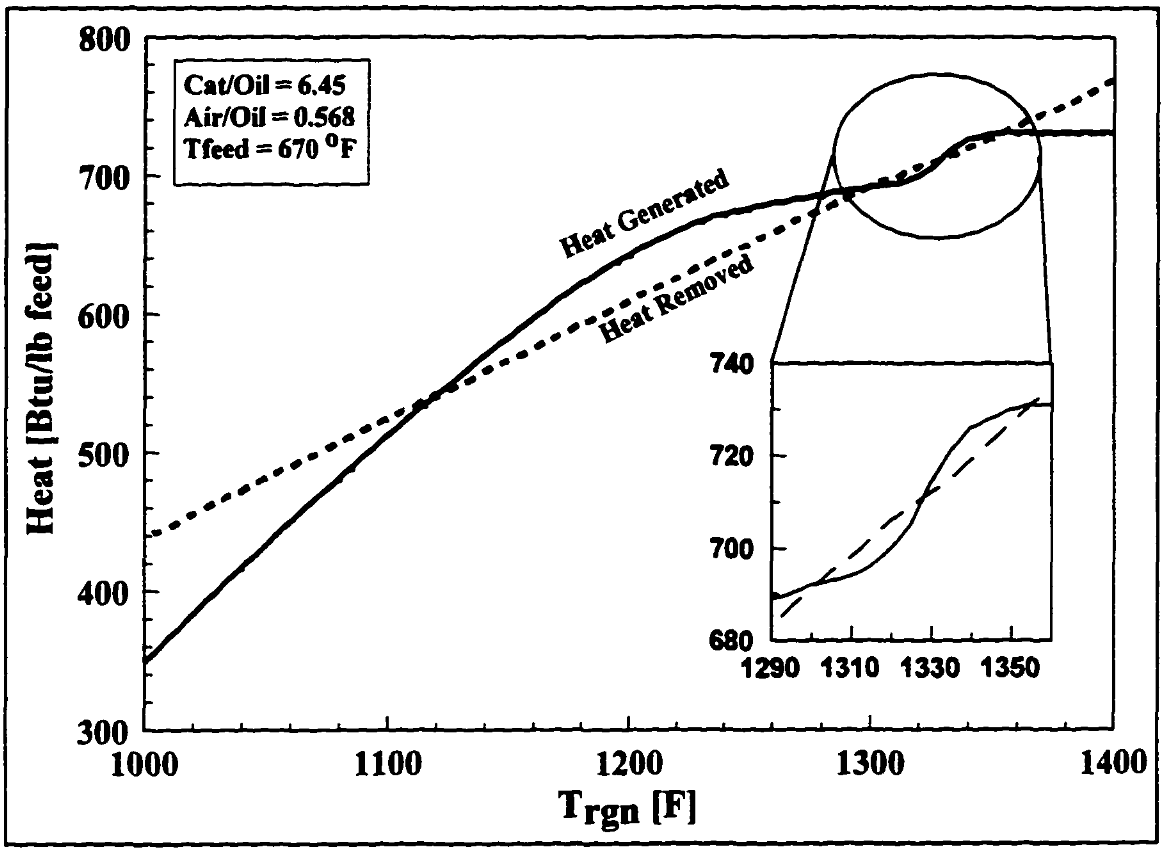


Figure 19: Heat generation and heat removal using hypothetical kinetic data for CO combustion (see text) leading to five steady states. (Low temperature steady state not shown.)

excess of oxygen would lead to higher combustion rates of CO and therefore a higher CO_2/CO ratio at the third steady state.

Therefore, the existence of an inflection point requires a significant difference in the relative reaction rates of coke and CO combustion between the regenerator temperatures of the third and fifth intersection. The maximum possible difference in reaction rate for this case is a function of the maximum temperature rise and the difference in activation energies. In a hydrocracker operating with a large excess of hydrogen it is not the difference but the value of the activation energy of the consecutive reaction that counts. At constant air and no excess oxygen, the maximum temperature rise is a function of the CO_2/CO ratio. At a CO_2/CO ratio of 1.0 the rise in T_{rgn} is about

120°F or less. In the model, the difference in activation energies between coke and homogeneous CO combustion is 35 kcal/gmole. This is too small to cause a real significant increase in the ratio of the rate of CO combustion to that of coke combustion with a change of only 120°F in T_{rgn}. The method of changing parameters to achieve five steady states resulted in an unrealistically low CO_2/CO value of 0.44. This raises the potential T_{rgn} rise to 200°F. The difference in the activation energies was also increased. If the CO_2/CO ratio would have been left at 1.0, which is the lowest value actually observed at partial combustion, increasing the activation energy of the homogeneous CO combustion to 210 kcal/mole would not be sufficient to give five steady states as shown in Fig. 20.

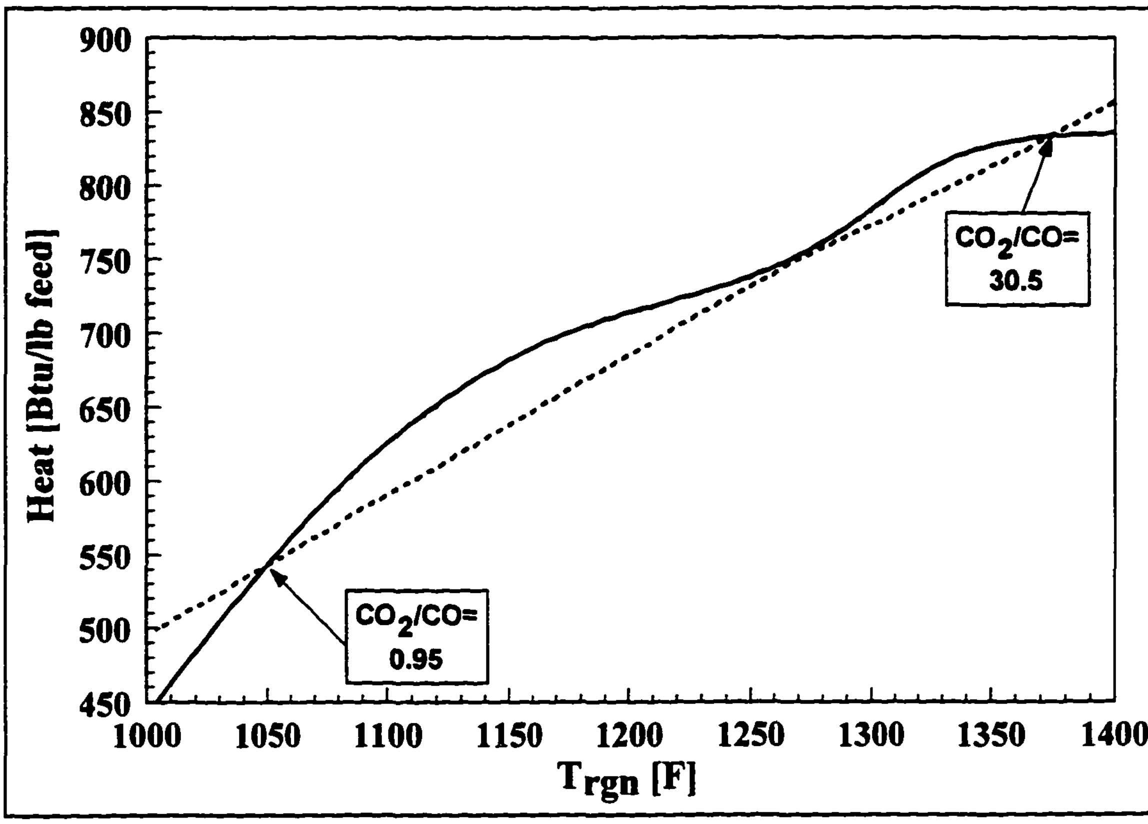


Figure 20: Heat generation and heat removal, re-plot of Fig. 18 with catalytic CO combustion rate increased back to base rate, resulting with only three steady states. (Low temperature steady state not shown.)

It can be concluded that five steady states in the regenerator due to the nature of the consecutive reactions are feasible but not likely. Again, it should be pointed out that all the above arguments refer to present catalyst and operating conditions. At high reactor temperatures coke kinetics change [Avidan and Shinnar (1991)]. This may affect the results.

Theoretically there is another potential for additional steady states in a regenerator. If there is a significant heat transfer resistance between the solid catalyst particles (or the dense catalyst phase) and the gas phase, there is a chance for excess temperatures in the particle phase. Ross and Fong (1981) observed such differences in a fluid bed coal combustor but only for much larger particles (>3 mm). For particles of the size used in FCC's (50-100 microns) this does not happen.

Elnashaie and El-Hennawi (1979) have published a paper showing a possibility for two steady states using a two-phase model for the fluid bed regenerator. Two phase models with a transfer resistance between the dense bed catalyst phase and a bubble phase have been published in the literature of fluidized bed reactors [Kunii and Levenspiel (1969)]. Avidan and Shinnar (1991) have shown that such models do not apply to FCC regenerators. But the problem with these models is more fundamental. The only data for the bubbling model that justify the use of the two phase model are based on either tracer studies or kinetic results and thus relate to mass transfer only. While for a single particle in the gas phase the ratio between the heat transfer and mass transfer coefficients is close to unity, the same ratio for interchange between the bubble and dense phase is very large, as heat transfer occurs by particles showering through the bubbles. There is a large body of experimental evidence in different kind of fluid bed reactors that the temperature of the gas leaving the bed is identical to the temperature of the bed itself. Thus, there is no experimental or theoretical basis for any model that predicts a significant difference between the mixed solid phase and the gas phase.

Another possibility of additional steady states or a more complex bifurcation map is in the interaction between coking in the reactor and the combustion in the regenerator. In the narrow range of temperatures for which there are reliable data (900-1050°F) and for the range of residence times studied (larger than one second), the coking relation is quite simple and monotonous in temperature with a low activation energy. The same applies to the cracking reactions which, at present conditions, have a quite simple kinetic form [Krambeck (1993)]. At higher temperatures the situation becomes much more complex. The same would apply for a radically different catalyst. For a complex system like the FCC, the result of a bifurcation analysis for such a case could be much more complex [Morbidelli et al. (1987), Razon and Schmitz (1987), Balakotaiah and Luss (1984), and Uppal and Ray (1974)]. The results presented here are therefore limited to the operating conditions studied and are based on the known kinetics of present catalysts in the operating range.

Last, there is a possibility in any design involving complex flows, such as in the FCC, for cyclic instabilities due to fluid mechanic reasons. These can be taken care of by proper design modifications and are outside the scope of this paper.

In summary, it may be concluded that in present FCC operation with present catalysts, it is safe to assume that there are only three steady states with the upper one being linearly stable. This might change should radically new catalysts introduced. The reader may ask why, given the present model, such a speculative discussion (albeit backed by computations) is needed. The reason is methodological. First, the implications of model uncertainties should always be investigated. Second, for most new processes there are no reliable models available. The first reasonable model for the complete FCC system came about twenty-five years after the first FCC went into operation. In such cases the analysis is limited to the type presented here and it is important for designers to familiarize themselves with such an approach.

4.4. Input Multiplicities and Multiple Steady States

Nonlinear systems like an FCC can have, in addition to multiple steady states, input multiplicities. The system has many state variables. Keeping a limited number constant by control does not mean that the system has a unique steady state for these control settings. If n measured variables are kept constant by an nxn dynamic control matrix there could be different combinations of the n manipulated inputs that give the same results in the n controlled variables. This is also true for $n=1$. Koppel (1982) called these input multiplicities and Balakotaiah and Luss (1985) investigated their properties. The important control questions are, therefore, what are the minimum requirements leading to guaranteed uniqueness, and what kind of multiplicities can one expect from different control structures?

The difference between multiple steady states and input multiplicities needs to be explained in more detail. If control loops are introduced into a nonlinear system, the nature of the system changes. If the nature of the control law is properly defined (This includes the structure of the controller and any nonlinearities due to saturation.), one can analyze the system under control for bifurcations in the same way one would an open loop system. In fact many reactors, such as in crystallization or polymerization, have internal feedback loops that lead to bifurcations [Liss and Shinnar (1973)]. Conceptually, there is no difference between a control loop and an internal feedback mechanism.

This is very complex. In practice a simplified approach is used that is sufficient for controller design. Consider for example a square control matrix [F_{cat}, F_{air}, T_{rgn}, T_{ris}]. It is possible to investigate the open loop bifurcation structure for the region of F_{cat} and F_{air} of interest. The results will be complex surfaces of T_{rgn} and T_{ris}. It is then possible to look at a given combination of T_{rgn} and T_{ris} and find if there are any combinations of $[F_{cat}, F_{air}]$ that lead to this set point. There could be none which makes this set point non-permissible. If there is more than one value of the vector of manipulated variables for a given vector of set points, this is an input multiplicity. Balakotaiah and Luss (1985) discuss in detail this type of bifurcation analysis.

For controller design, it is useful and important to know not only the number of such possible input conditions, but also the properties of the open loop system at each point (stability, gains, etc.). The values of F_{cat} and F_{air} can be computed for a given set point $[T_{rgn}, T_{ris}]$ without knowing anything about the design of the controller. This approach teaches nothing about the closed loop stability of the different steady states, since this is a function of controller design. The result is the open loop stability at each point.

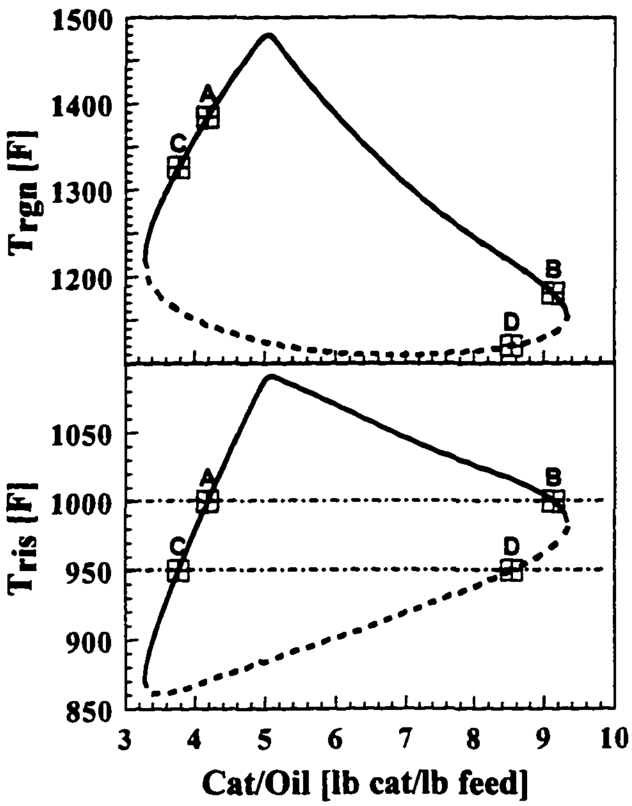
The analysis will not give all the steady states for this controller setting, which strongly depends on controller design. Even a simple stable steady state can be converted into a limit cycle by a feedback controller. For an FCC no controller can eliminate the cold steady state, which exists for all settings and control structures. Furthermore, there is a potential for further stable steady states at the saturation values of either F_{cat} or F_{air} or of both together, with different values of $[T_{rgn}, T_{ris}]$. When the valve is fully open, the operation may be stable despite the deviation from the set point.

The use of input multiplicities simplifies the problem by allowing the impact of controller design on each of the possible operating points to be studied. This is not equivalent to a bifurcation analysis of the system, and all the laws and rules about the nature of steady states or their stability have here no meaning. Balakotaiah and Luss

(1985) also point out that constraints on the manipulated variables have to be taken into account in determining practical input multiplicities. It will be shown later that in FCC's the matrix [F_{cat}, F_{air}, T_{reg}, T_{ris}] has always a double input multiplicity of which one value is often unachievable due to saturation. Other matrices may have quadruple multiplicities. This is important as it is much easier to take care of a double input multiplicity with integral controllers [Morari (1983)] than with quadruple multiplicities. When there are two set points, the gain has to change sign such that only one set point is stable with an integral controller. In quadruple input multiplicities, two set points may have gains with equal signs.

The most simple case is input multiplicities with a single control loop. Edwards and Kim (1988) discuss a case that occurred in an actual plant operation. A unit operating with control of both T_{rgn} and T_{ris} was put on manual control of the air rate. T_{ris} was still controlled by adjusting catalyst circulation via a slide valve. The unit was operating initially in a partial combustion mode and drifted to full CO combustion. Arandes and de-Lasa (1992) analyzed the same case with their model and observed similar behavior.

The reason the unit drifted to full CO combustion can be explained by looking at Fig. 7, part of which is given again as Fig. 21. Fig. 21 shows only the two hot steady states for a fixed value of Air/Oil when Cat/Oil is varied. For a range of Cat/Oil, there are three steady states, the upper one being stable. For this specific case, there are two Cat/Oil values for which T_{ris} = 1000°F, one in complete CO combustion, point A, and one in partial combustion, point B. Both cases are open loop linearly stable as all the eigenvalues are negative and real. The gain of T_{ris} with increasing Cat/Oil is positive in



	A	B
Tris [F]	1000	1000
Trgn [F]	1384	1181
Cat/Oil	4.18	9.13
CO2/CO,cyc	1509	0.98
Eigenvalues	-2.94	-0.92
[1/min]	-0.41	-0.83
	-0.33	-0.28
	-0.05	-0.03
	-0.02	-0.02

	C	D
Tris [F]	950	950
Trgn [F]	1326	1120
Cat/Oil	3.77	8.54
CO2/CO,cyc	508	1
Eigenvalues	-2.75	-0.84
[1/min]	-0.37	-0.79
	-0.30	-0.26
	-0.04	*0.03*
	-0.02	-0.02

Figure 21: Cross-cuts of T_{rgn} and T_{ris} at constant Air/Oil = 0.6. Properties for two points at the same T_{ris} are given.

full CO combustion (point A) and negative in partial combustion (point B), as explained in Section 3. As both steady states are stable, a stable feedback controller can be designed measuring T_{ris} and manipulating catalyst flow (expressed as Cat/Oil in the figures). This can be done for each of the two steady states by choosing the right sign for the gain. But in this case, only one of the two steady states will be linearly stable under control as a feedback controller with integral action is always unstable for positive feedback. In many cases such an input multiplicity is therefore much less of a problem than one would initially conclude [Morari (1983)].

This explains the occurrence mentioned by Edwards and Kim (1988). If T_{rgn} is kept constant by control, the open loop gain for the control loop [T_{ris}, F_{cat}] is positive, both in partial and complete CO combustion. In partial combustion, if both control loops operate, the system is stable for a positive gain in both loops. If the air rate is put on manual control, T_{rgn} is no longer constant and the open loop gain of the reactor control loop reverses sign making the system unstable as an integral controller cannot operate with positive feedback. The unit will either drift to the only hot stable state, which maintained a positive gain, or wind down to a cold state. Therefore, to put T_{rgn} control on manual in partial CO combustion, the sign of the control of T_{ris} via F_{cat} has to be reversed.

In the Edwards and Kim case, the unit drifted into full CO combustion but, with a different disturbance, it could have gone in the other direction by increasing F_{cat} until either the unit loses the upper steady state and winds down or F_{cat} reaches saturation. If there is a stable steady state at F_{cat}(max) than the unit could reach it and stay there despite the fact that T_{ris} is less than the set point, but the operator probably would intervene. It is well known that an open loop stable system with two control loops can be stable when both operate and becomes unstable when one loop is opened [Grosdidier, Morari and Holt (1985)]. This is the case here as for the coupling of [T_{rgn} -F_{air}] and [T_{ris}-F_{cat}] the RGA elements are negative.

It should be noted that the steady state A (complete CO combustion) in Fig. 21 is a valid stable operating point but not a sensible one. An operator would reduce the air rate to bring a unit in state B to complete CO combustion. Point A has too much excess air. In Fig. 21 and similar figures all possible steady states are shown. Only a fraction of the steady states in such plots is economically attractive.

Edwards and Kim argue correctly that these are multiple steady states in the

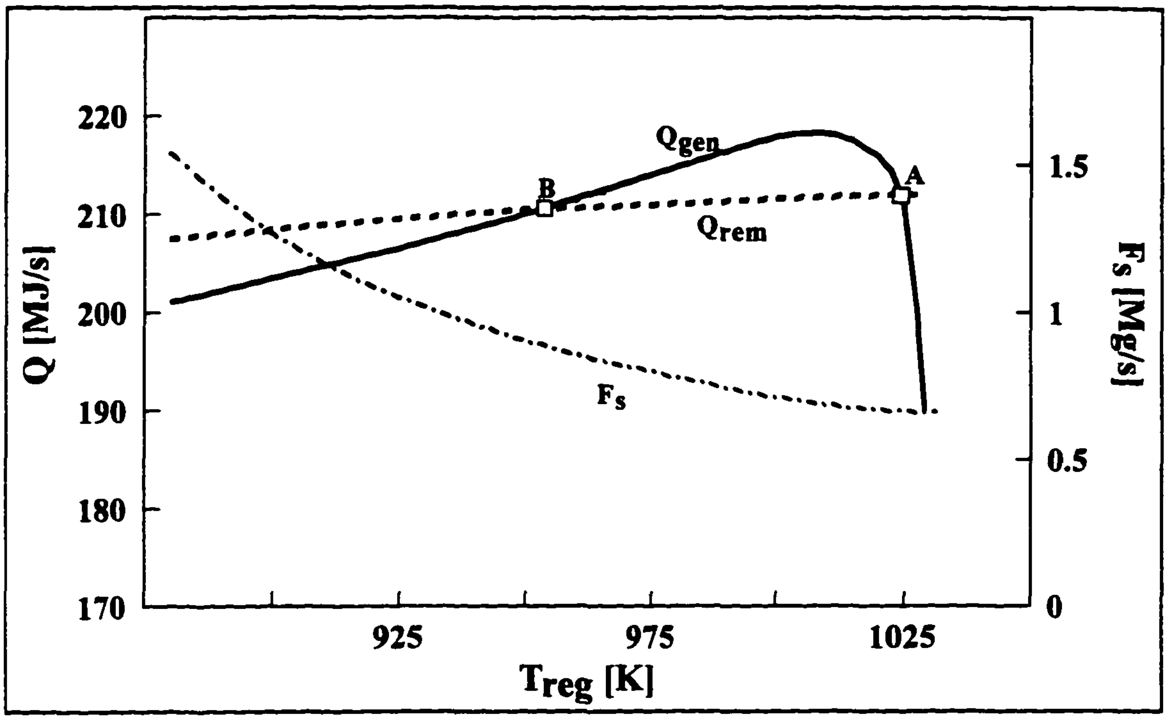


Figure 22: Heat generation and heat removal from Arandes and de Lasa (1992). Air flow rate and riser top temperature are kept constant. Catalyst circulation rate varies.

classical sense. The catalyst circulation is an internal variable, adjusted by a control circuit, and can be looked at as part of the system. Edwards and Kim (1988) and Arandes and de-Lasa (1992) plotted the heat balance varying both T_{rgn} and F_{cat}, manipulating F_{cat} such that T_{ris} stays constant, as can be seen in Fig. 22 reproduced from Arandes and de-Lasa (1992). This is equivalent to an analysis of input multiplicities and gives two of the proper upper stable steady states A and B analogous to those in Fig. 21, but misses the lower cold steady state as well as the intermediate unstable ones. Edwards and Kim also draw conclusions from Fig. 22 about the stability of the system and conclude that operating in partial combustion is unstable. They base their conclusion on the slope argument, comparing the slopes of the heat removal and heat generation lines at the intersection. Fig. 22 gives no information about stability or total number of bifurcations. In the analysis in this paper, all conclusions on stability are based on computing eigenvalues for the open loop at each point. Such an analysis shows that both steady states A and B in Fig. 21 and 22 are linearly stable for the example chosen. But this result can not be generalized. For other cases, Fig. 21 shows that both could be unstable or only one could be stable. Thus for example in Fig. 21 a set point of 950°F for T_{ns} at point C (complete CO combustion) would be stable, whereas the second hot steady state, point D (partial combustion) would be linearly unstable. Again such conclusions have to be based on a rigorous analysis of the open loop. With a proper controller on T_{ris} some unstable steady states can be stabilized, but this is outside the scope of this work.

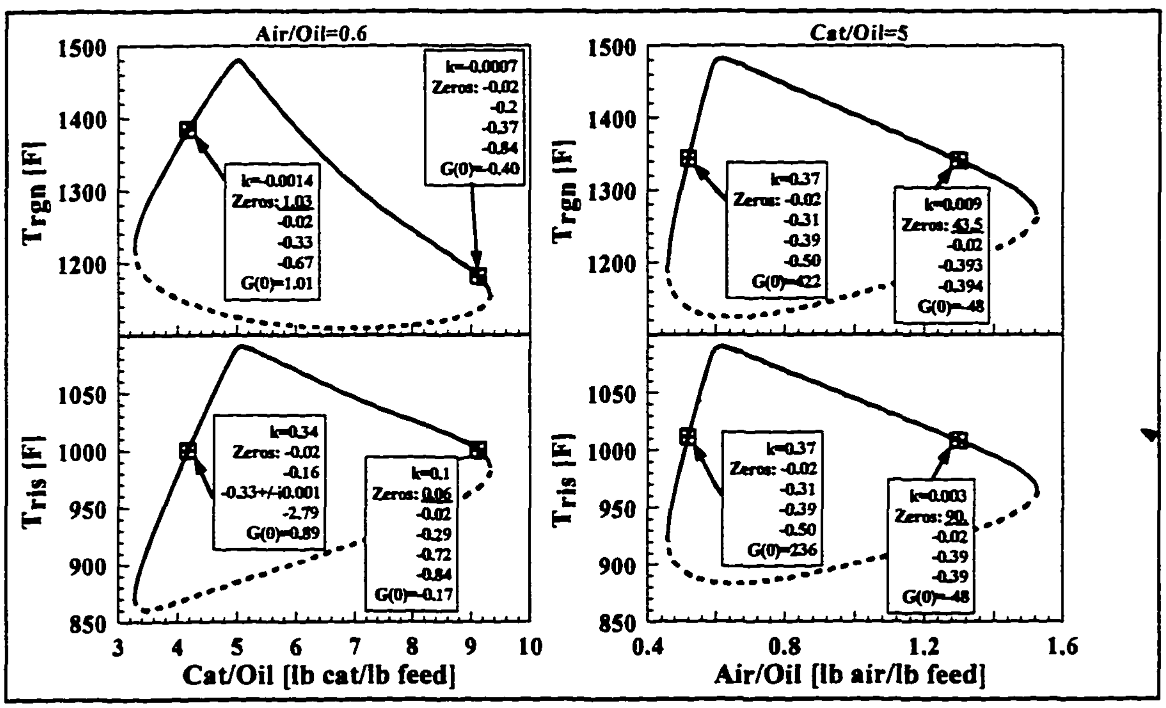


Figure 23: Cross-cuts at constant Air/Oil and Cat/Oil. The gain, zeros, and steady state gain for some points are shown. (Gains are in temperature/variable changed, zeros are in 1/min).

It is also worthwhile to recognize that a similar multiplicity will result if Cat/Oil is kept at a fixed value and T_{rgn} is controlled by manipulating air rate. This can be seen from Fig. 7 part of which is shown in Fig. 23. Again, the region for which the steady states are linearly stable as computed from the eigenvalues is indicated. In addition, the zeros for some stable steady states are shown. For open loop stability, only the sign of the poles (eigenvalues) is important. For control and understanding the dynamic response, however, knowledge of the zeros is important, as RHP (positive) zeros imply inverse response. Interestingly, a change in Cat/Oil shows an inverse response (RHP zero) for T_{ris} in partial combustion and an inverse response for T_{rgn} in complete combustion. When changing Air/Oil at fixed Cat/Oil, there is a RHP zero for both T_{rgn} and T_{ris} in complete combustion. However this zero does not show in the dynamic responses given in Fig. 12 and 13 because its time scale is much shorter then the other zeros and poles.

4.4.1. Input Multiplicities with Different Control Schemes

In Fig. 21, the input multiplicity arose when Air/Oil was kept constant and T_{ris} was controlled by manipulating Cat/Oil. This is not a special case. Any control based on manipulating F_{air} and F_{cat} will have input multiplicities. This is an unavoidable property of an FCC and has to be understood for proper control. But showing it is more complex than is depicted in Fig. 21 as there are four dimensions. One way to represent this is by plotting lines of constant T_{rgn} and T_{ris} in the plane [Cat/Oil, Air/Oil] for a specific case of input conditions. This is shown in Fig. 24. Both lines form ellipsoids, as would be expected from Fig. 21. Therefore, either two or four intersections would be expected.

Fig. 24 shows four intersections. However, each isotherm is a projection of a curve in a three dimensional space. Only two of these intersections are real multiplicities. This can be checked as follows. At each [Air/Oil, Cat/Oil] point in the plot there are three steady states. An intersection is real only if both temperatures refer to the same steady state. In the example of Fig. 24 only two intersections fulfill this condition, one

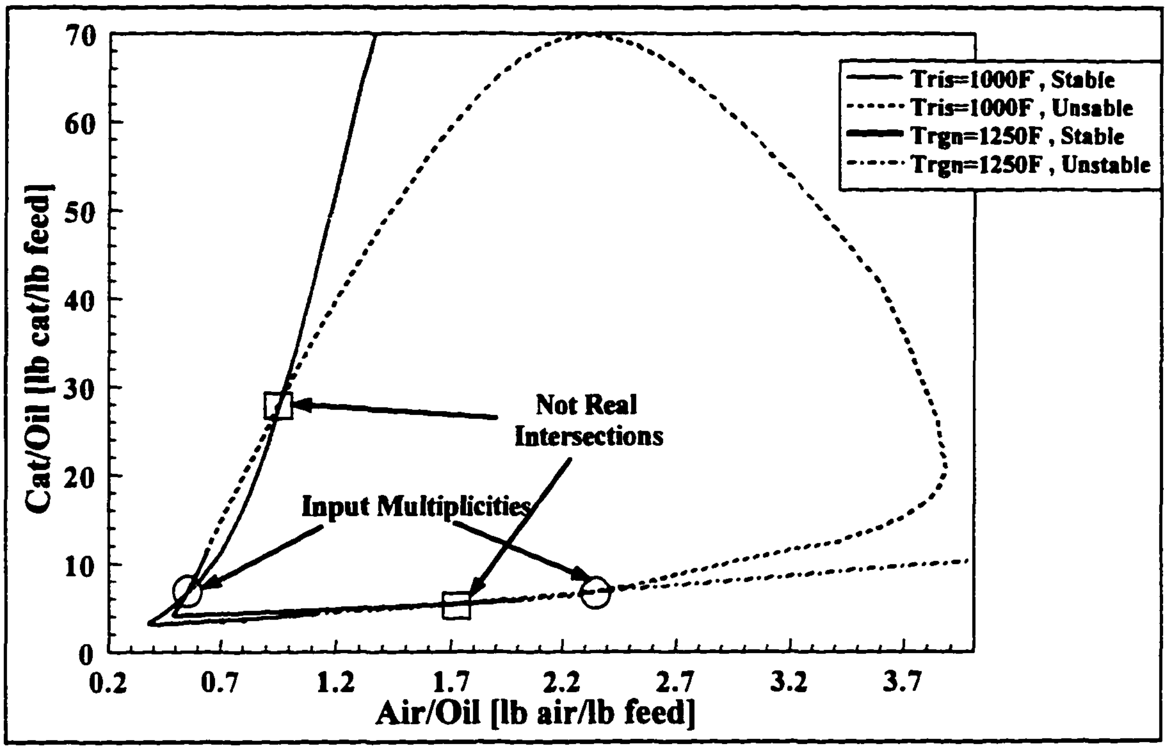


Figure 24: Projection of constant T_{rgn} and T_{ris} curves on the Air/Oil - Cat/Oil plane.

relating to a stable the other to an unstable steady state. For other combinations of T_{rgn} and T_{ris} both could be unstable or both stable. For present operating conditions and catalysts, it is highly unlikely there could be more than two real intersections. The reason is that if T_{rgn} and T_{ris} are kept within reasonable operating range, Cat/Oil vary only over a narrow range. Sufficient heat is needed to vaporize the oil. This is fixed for a given T_{ris}. The heat of conversion is small compared to the heat of vaporization. Furthermore, with present catalysts conversion is tied to coke make (see Section 3), which again cannot vary very much. As the ellipsoid has no inflection points around the surface, this constraint makes it unlikely to have more than two intersections. Realizing this, the map for a given T_{ris} giving these two steady states for each T_{rgn} can be plotted, as shown in Fig. 25. Plotting it this way makes sense as T_{ris} is usually controlled with a very tight spec.

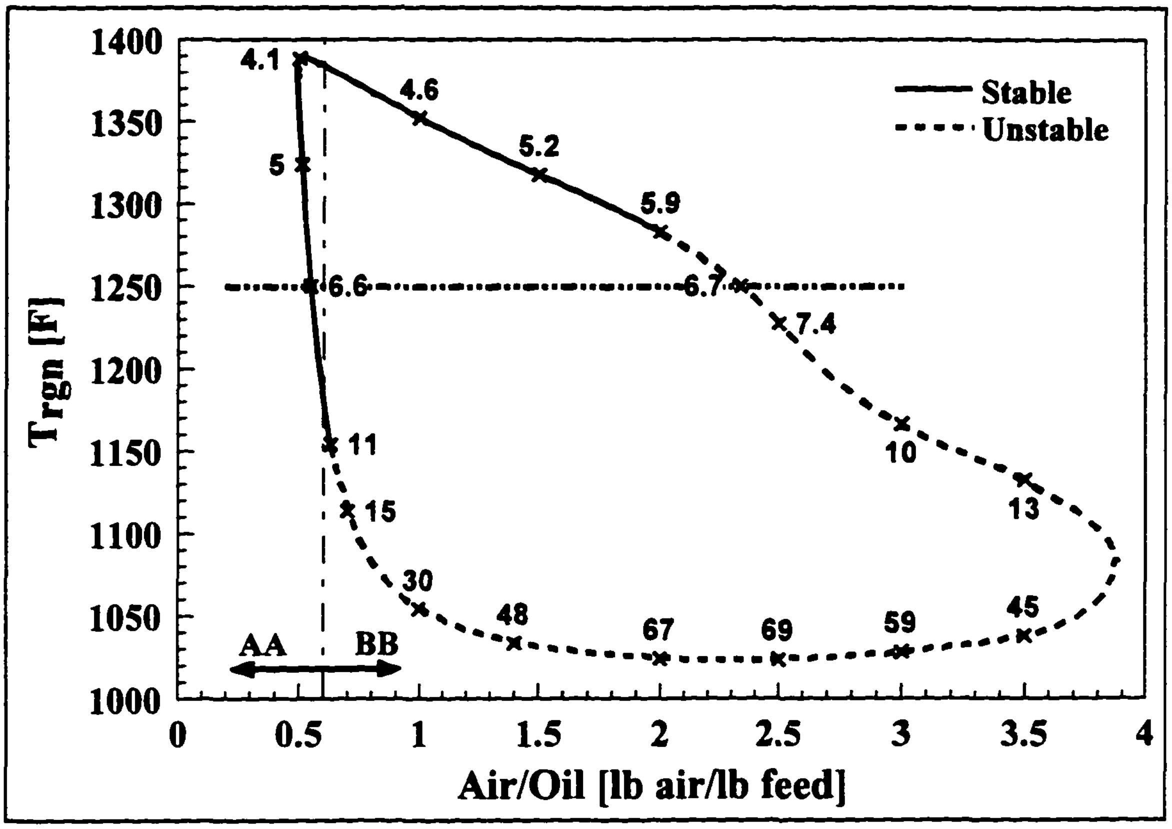


Figure 25: Cross-cut of T_{rgn} at constant T_{ris} = 1000°F. T_{ris} is kept constant by adjusting Cat/Oil as indicated by the numbers on the plot.

In Fig. 25 a plot of T_{rgn} versus air rate (expressed as Air/Oil) for the condition where T_{ris} is controlled by F_{cat} at a constant value of 1000°F is given. The plot would look identical if in the actual control T_{rgn} is controlled by F_{cat} and T_{ris} by F_{air} (if all of $[T_{rgn}, T_{ris}, F_{air}, F_{cat}]$ are fixed the steady state is always unique). But in practical control, all four values can not be fixed. Only the set points for the measured variables can be chosen.

At the minimum and maximum of Fig. 25, the unit is not controllable with these two loops as the gains are zero. The region around the minimum of T_{rgn} in Fig. 25 is not in an interesting range. Not only is this operating point linearly unstable, but it is also at too low regenerator temperature and therefore a very high value of Cat/Oil. The range close to the maximum value of T_{rgn} refers to complete CO combustion and is of commercial interest. Units operating in full combustion operate always close to the maximum. Therefore, in complete combustion it is best to use a control that keeps excess oxygen and T_{ris} constant.

There is one difference between Fig. 25 and Fig. 21 (or 22). In Fig. 21 both operating points (A and B) are of commercial interest and within the constraints of the unit. In Fig. 25, this is only true close to the maximum of T_{rgn} (full CO combustion) where the slope is practically zero. There excess oxygen is controlled. In partial combustion only the part to the left of the minimum is of commercial interest (see Fig. 25, region AA). On the right side excess air is high to no economic advantage. Units do not operate with that much excess air. Therefore the unit would hit a constraint if it tried to drift to region B-B. A detailed consideration of control (such as proper tuning) will be discussed in Section 5. At present, the emphasis is on understanding input multiplicities which, just like multiple steady states, are an inherent feature of present FCC operation.

No control scheme in present FCC's (with no air preheat) can avoid multiple steady states. However, properly designed control schemes can avoid input multiplicities. What is important to understand is that no 2x2 control matrix using F_{air} and F_{cat} as manipulated variables has unique steady states. This is not only true for $[T_{rgn}, T_{ris}]$ but also for $[T_{ris}, O_{2,sg}]$, $[T_{ris}, \Delta T]$, $[T_{ris}, T_{sg}]$ etc., if F_{air} and F_{cat} are the manipulated variables. An example for these three combinations is given in Fig. 26. The case $[T_{ris}, O_{2,sg}]$ is similar to the case $[T_{ris}, T_{rgn}]$ because the multiplicities are always double, and in the interesting range ($0.5\% < O_{2,sg} < 2\%$) the second operating point has a much higher air rate. The plot $[T_{ris}, T_{sg}]$ is however, much more complex because it has quadruple

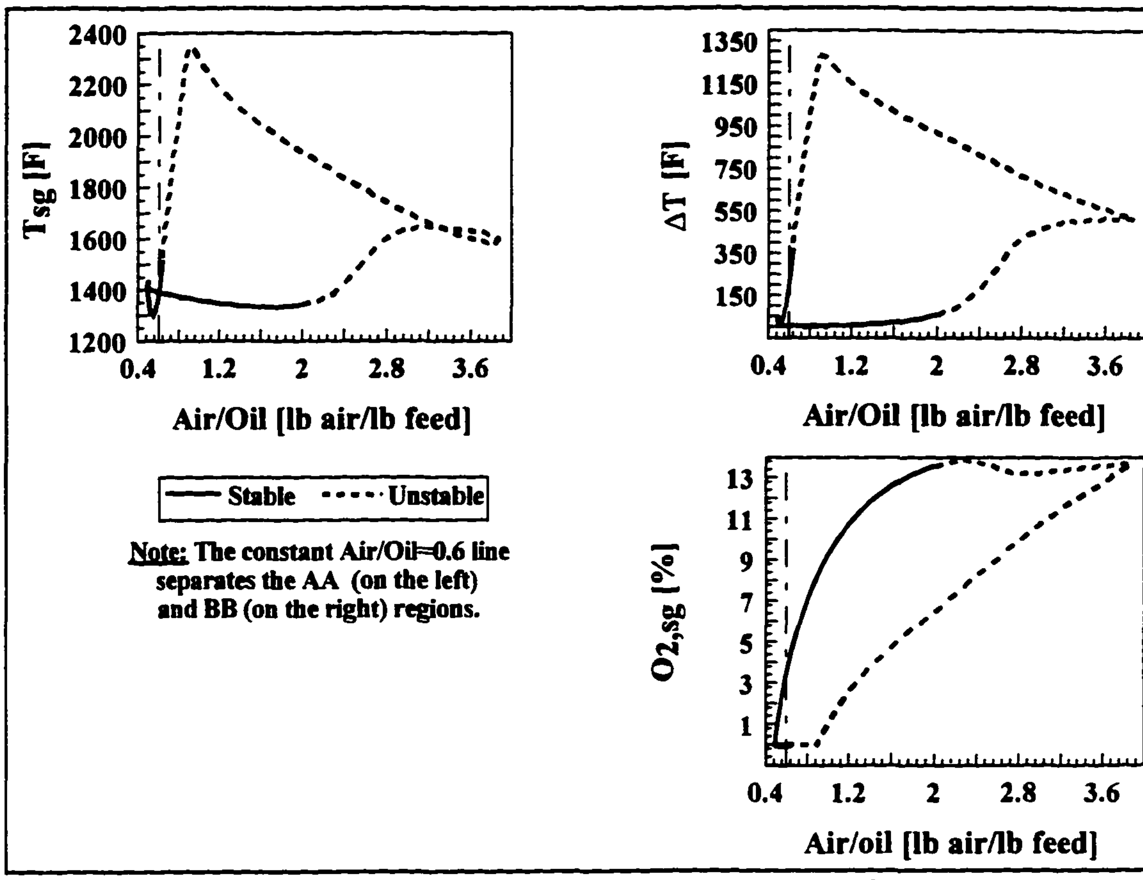


Figure 26: Cross-cuts of T_{sg}, ΔT, and $O_{2,sg}$ at constant T_{ris} = 1000°F. T_{ris} is kept constant by adjusting Cat/Oil.

multiplicities in a practical region of T_{sg} (AA). Furthermore, the differences in air rate are quite small and the slopes are identical. This presents practical control problems. The minimum also occurs in the region of practical interest. The case [T_{ris}, ΔT] has also quadruple multiplicities but not in the range where ΔT control is practical (AA), but each specific case must be checked. For the cases given in Fig. 25 and 26, the matrix [T_{ris}, T_{sg}, F_{air}, F_{cat}] is the only one for which input multiplicities are a real problem for the designer. The reason that the scheme [T_{ris} T_{sg}] has quadruple steady states is that the surface of possible upper steady states in the space [F_{cat}, F_{air}, T_{ris}, T_{sg}] is not uniformly concave as in the case of [T_{ris}, T_{rgn}]. This can be seen from Fig. 27 which is equivalent to Fig. 21. The same is true for [F_{cat}, F_{air}, T_{ris}, ΔT]. While Fig. 25 and 26 give no direct information on controller design and tuning, they are essential for choosing control loops and understanding their pitfalls.

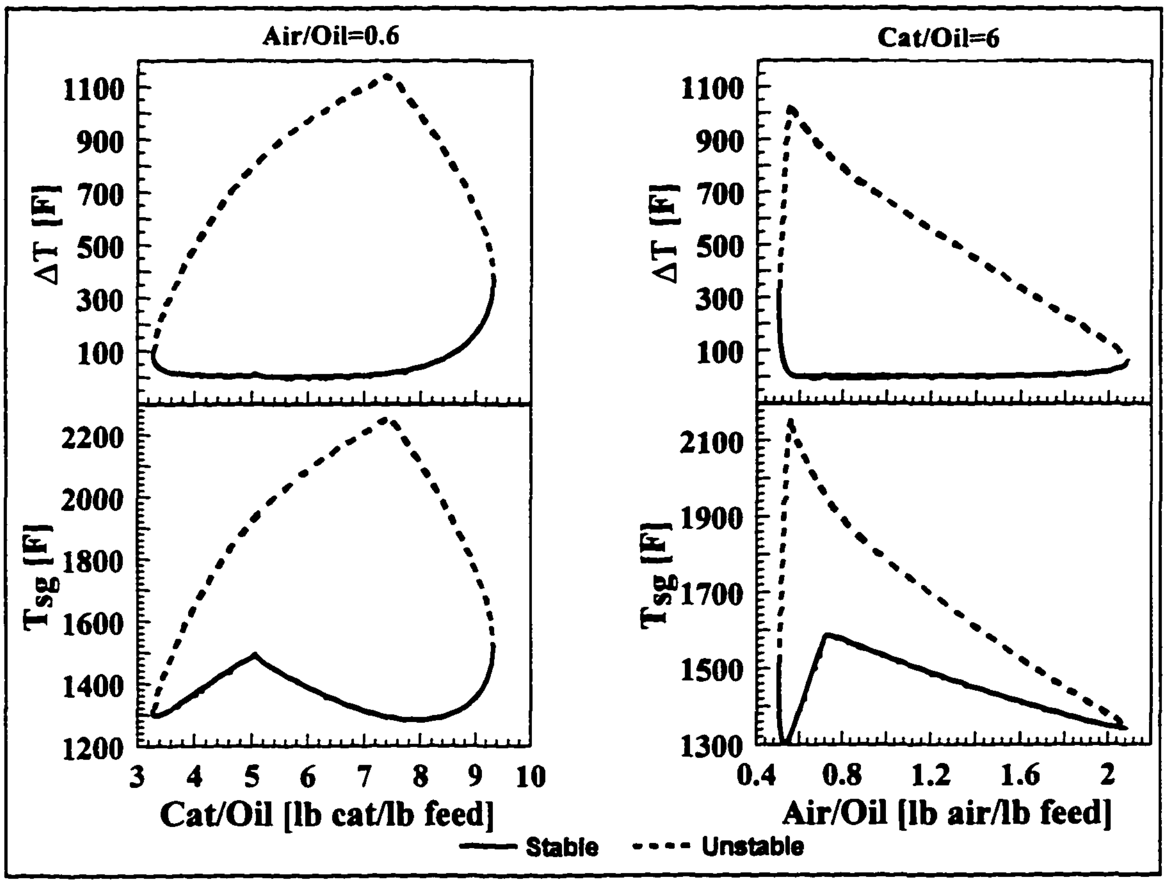


Figure 27: Cross-cuts of ΔT and T_{sg} at constant Air/Oil and Cat/Oil.

In Figs. 25 and 26 there is a problem with ΔT and therefore with T_{sg} at regenerator temperatures below 1120°F. As explained in Section 3, at low T_{rgn} ignition in the cyclones ceases, and ΔT is zero. This model is unreliable below this T_{rgn}. For the purpose of these two plots the ignition was left to happen at any T_{rgn}, but no FCC operates at such low temperatures.

There is one last problem to address. If F_{cat}, F_{air} are controlled at a unique steady state for $[T_{ris}, T_{rgn}, F_{cat}, F_{air}]$ is the steady state of the total system unique? The analysis in Section 4.3 of this paper shows that this is so. The crucial assumption is that for given values of Cat/Oil, T_{ris} and C_{rgc} the reactor (riser) itself has a unique steady state, which is true for this model and also for present FCC catalysts.

4.5. Impact of Operating Conditions on the Existence and Stability of the Steady States

In the previous sections it was shown that:

1. With presently used catalysts and designs, an FCC has either one or three steady states. If there is only one, it is the cold steady state and is of no interest.
2. For any given operating conditions with two hot steady states, the upper one is linearly stable and the lower hot state is unstable.

From the stand point of the operator who has imperfect knowledge of the catalyst state and feed properties, the question is different. Is a given operating point open loop stable or not? What changes in conditions will make a stable steady state unstable or vice versa, and for what range of operating conditions has the FCC a useful hot steady state (stable or unstable)? So, instead of looking at the stability and features of a single operating point, an operating map is needed.

This is a multi-dimensional problem. There are large numbers of inputs: feed rate, air rate, catalyst circulation rate, feed temperature, feed quality, catalyst type and activity, and combustion promoter level, to name some. The latter two are adjusted by the addition and withdrawal of catalyst, and it will be assumed that they are fixed at specific levels. Further, different types of catalysts will not be considered; the primary goal is simply to illustrate the basic principles and trends. As a result, for a given set of inputs the steady state is characterized by the temperatures.

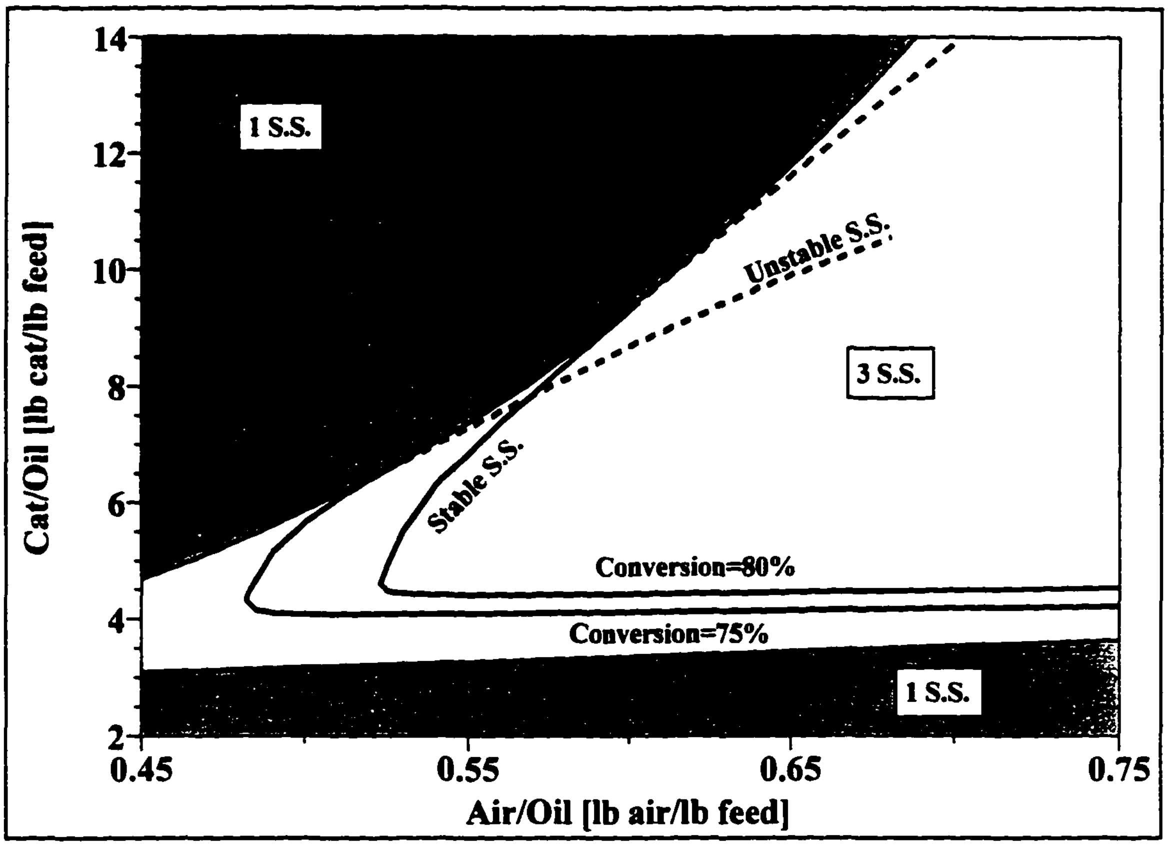


Figure 28: Steady state operating map (manipulated variable plane): base case with constant conversion lines.

Regrettably, it is hard to plot more than two dimensions. First, the boundaries for F_{air} and F_{cat} where three steady states exist when all other inputs are fixed are considered. For the base case, these are shown in Fig. 28, where again they are normalized by the feed rate, and Air/Oil and Cat/Oil are used instead of F_{cat} and F_{air}. Also shown are lines of constant conversion which indicate the useful operating range. Part of the constant conversion lines involve unstable steady states, as indicated on the plot. In this case there are no hot steady states for low values of Air/Oil regardless of the value of Cat/Oil (this limit is not shown). For a given permissible Air/Oil, there are a maximum and minimum permissible values of Cat/Oil. Similarly there are corresponding limits for Air/Oil for a given value of Cat/Oil. However, for this case the maximum permissible value of Air/Oil for reasonable values of Cat/Oil is outside the practical range of interest for operation.

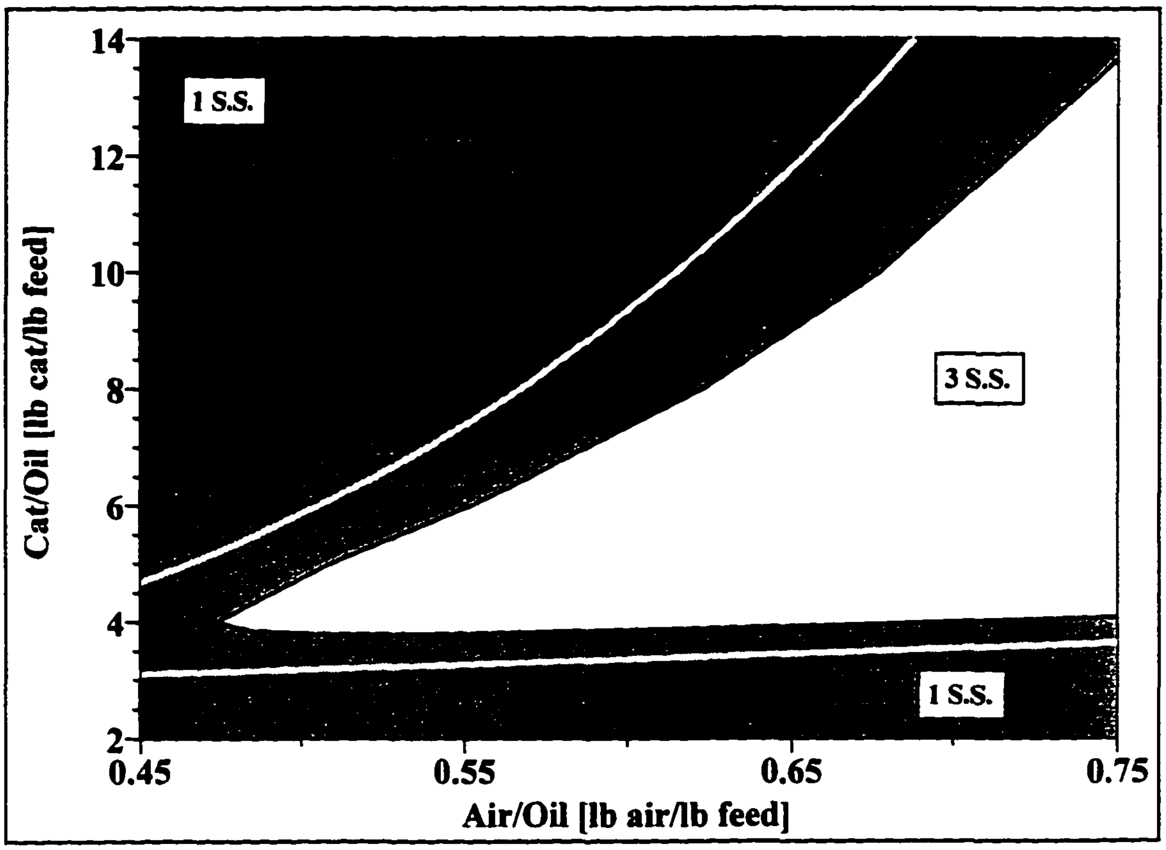


Figure 29: Steady state operating map (manipulated variables plane): feed preheat at 600°F. White lines are the base case of Fig. 28.

Second, the effect of changes in inlet conditions change the boundaries. Fig. 29 shows the impact of feed temperature. Lower feed temperature or, in other words, decreasing or eliminating feed preheat, decreases the range of permissible operating conditions. Decreasing catalyst activity or decreasing the coking tendency of the feed stock have a impact similar to decreasing feed temperature in that they narrow the range where an upper steady state exists.

In the same way adding a feed with a Conradson Carbon Residue (CCR) content widens the operable region at constant conditions while simultaneously reducing conversion. There is one exception however. If the CCR content becomes high (above

5%) the minimum Air/Oil increases reversing the above trend. There must be sufficient air to at least combust all the Conradson Carbon. This is not true for a high coking feed as in this case the higher coking rate results in a higher coke level on regenerated catalyst. This reduces catalyst activity and conversion but total coke make will change only slightly. The unit will remain stable. The formation of coke from Conradson Carbon is (at least in this model), independent of catalyst activity and for a high CCR, the minimum Air/Oil to combust it is large. If the CCR is not completely combusted then it would continue to build up leading to an instability of a different type.

Fig. 28 and 29 indicate that under present operating conditions with present catalysts it is highly unlikely that changes in feed stock or catalyst activity will cause all hot steady states to disappear. What might happen is that in open loop, the upset due to such a perturbation will cause the unit to wind down or for the steady state to exceed the temperature constraints of the unit. Normally, therefore, at least reactor temperature is controlled. For the operator it is therefore much more illuminating to re-plot the data in Fig. 28 in a way that represents the open loop stability and feasibility of different choices for the set point of temperature controllers. This is done in Fig. 30, which represents the same case as in Fig. 28.

The coordinates in Fig. 30 are T_{rgn} and T_{ris}. When evaluating FCC operation, it is customary to think in terms of T_{ris} and T_{rgn} despite the fact that sometimes other combinations of variables are controlled. Thus in complete CO combustion, excess oxygen in the flue gas is normally controlled. In partial combustion the temperature rise across the cyclones (which is a measure of excess oxygen) is sometimes controlled. One could re-plot Fig. 30 for any such combination, but plotting T_{rgn} and T_{ris} has been found

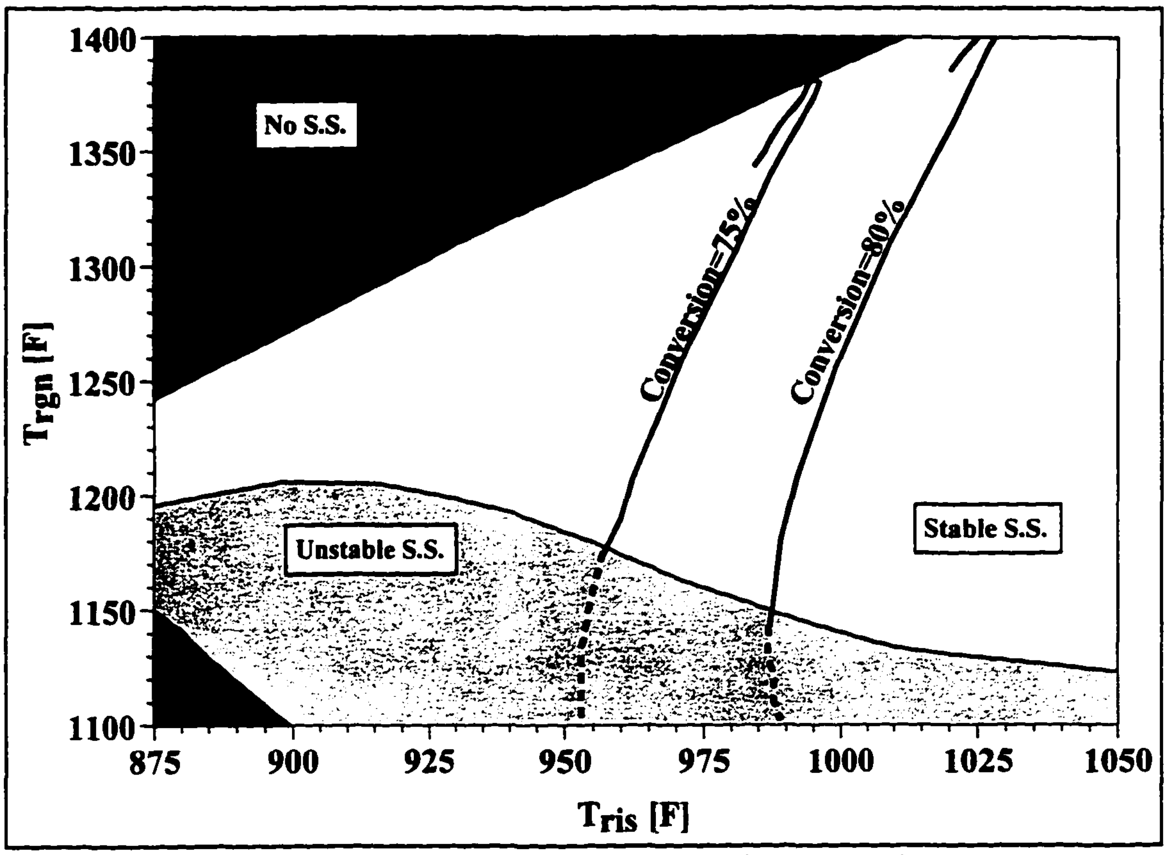


Figure 30: Steady state operating map (outputs plane): base case with constant conversion lines.

to be more illuminating.

The plot has three areas. For each T_{ris} there is a high limit to T_{rgn} , above which there is no steady state, a middle limit below which only unstable steady states exist, and a lower limit below which no steady state exists. For the case presented, this lower limit is below any practical choice of set points.

There is one caveat. While fixing Air/Oil, Cat/Oil, T_{rgn}, and T_{ris} leads to a unique steady state, fixing T_{rgn} and T_{ris} alone does not do so. There is a region of input multiplicities. The same applies to any other combination of two outputs. This was discussed in Section 4.4. It was shown in Section 4.4 that for partial combustion the second steady state with the same [T_{rgn}, T_{ris}] occurs at much higher air rates, and is not a practical operating point. Therefore operation is always to the left of the minimum in Fig. 25. In partial combustion the value of T_{rgn} below which the steady state in Fig. 25 becomes unstable is significantly lower on the left side than on the right side (high air rates). This is always so. But as the right side is impractical, the stability limit plotted in Fig. 30 is based on the left stability limit of Fig. 25.

The most interesting feature of Fig. 30 is the lower limit of linearly stable operation. It is highly unlikely that a perturbation in feed stock qualities or catalyst properties will make a given combination of T_{rgn} and T_{ris} currently used in partial combustion non-feasible. (In complete combustion, as said, T_{ris} and $O_{2,sg}$ are normally controlled. T_{rgn} there will strongly depend on feed stock qualities and cannot be a-priori chosen.) However a stable steady state in partial combustion could become unstable if feed properties are perturbed. It is also possible that this setting could lead to values of F_{air} and F_{cat} that, for the given feed rate, are outside the capabilities of the unit. Flow rate constraints, while important, are outside the scope of this work.

The main value of plots like Fig. 30 is to show what changes can de-stabilize (in open loop) a stable FCC operation. As the operation is normally in closed loop control, such changes do not have to lead to actual instability. But as controllers are often not tuned or designed to stabilize the system, understanding the likelihood of an open loop stable operating point to become linearly unstable is important. This is shown in Fig. 31 where only the dividing line between linear stable and unstable operation is plotted for different perturbations and changes in operating conditions.

Decreasing the feed preheat (or reactor feed temperature) moves the boundary to higher values of T_{rgn} and T_{ris}, thus increasing the probability that a realistic control setting

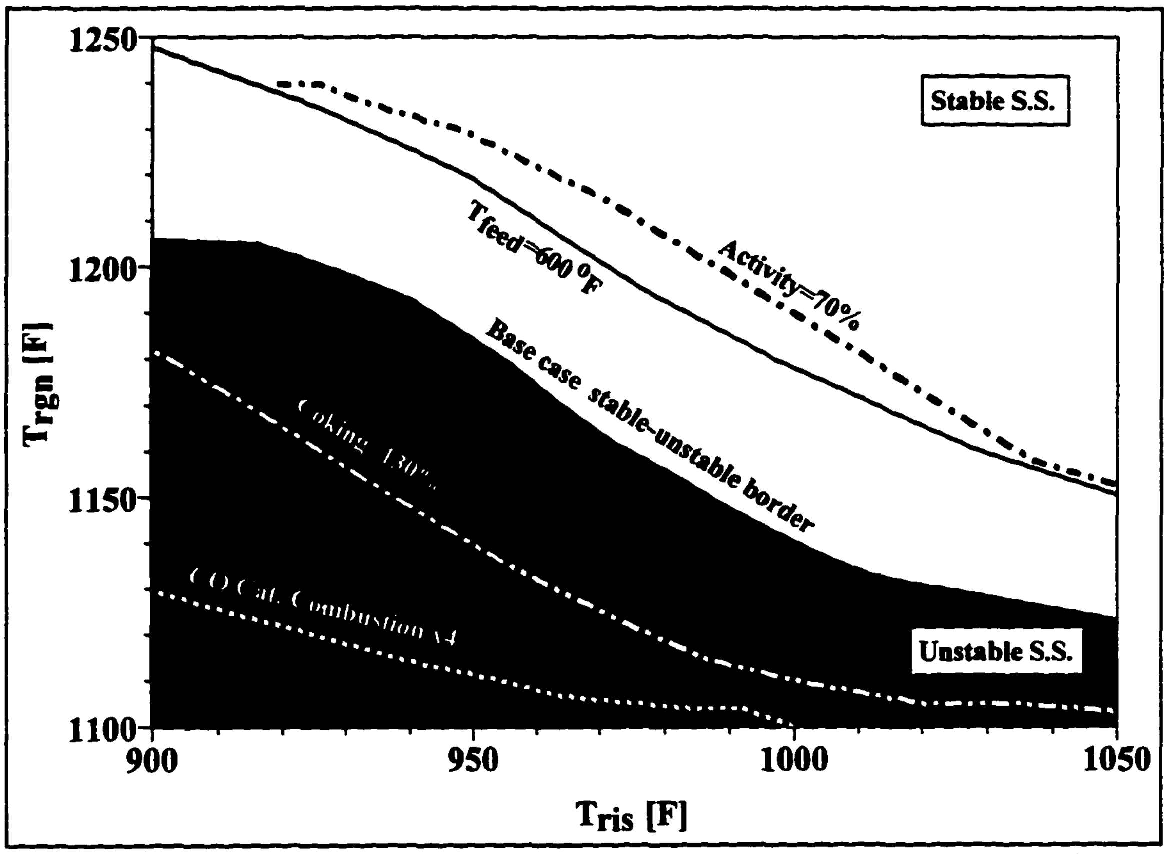


Figure 31: Steady state operating map (outputs plane): effect on unstable-stable border of: (1) A = -30%; (2) T_{feed} = 600°F; (3) z = +30%; (4) CO catalytic combustion rate = x4.

is at an open loop unstable steady state. The same is true for lower catalyst activity and lower coking feeds (more selective catalysts). All changes that were studied that narrow the operable space in Fig. 28 move the stability boundary to higher values of regenerator and reactor temperature. Fig. 31 also shows that CO combustion promoters lower the temperature limits below which the operation becomes unstable. As operation in full CO combustion always involves high temperature (above 1250°F) and use of promoters, this means that regeneration in full CO combustion will always be linearly stable, at least for practical operating conditions. Looking at present operating conditions, which involve normally higher regenerator temperature (>1230°F), higher reactor temperatures (often around 1000°F), preheat, and heavier coking feeds, it is quite unlikely for a unit to be linearly unstable. However, if conditions prevalent twenty or thirty years ago (lower preheat, lower T_{rgn} and T_{ris}) are considered, the probability for a unit to be at an linearly unstable steady state was quite reasonable.

Understanding the nonlinear topology of the system is important for the FCC because both intentional and non-intentional perturbations in feed stock and catalyst properties can be quite large. A 30% perturbation in coking tendency or catalyst activity has a reasonable likelihood of occurrence. One important aspect of FCC operation in the past was the concern of the operator of losing the unit due to wind down to a cold steady state. Kurihara (1967) discusses this. It usually occurred when a higher coking feed stock was introduced. In the way FCC's were operated (and still are) as the waste basket of the refinery, this sometimes occurred without clear warning to the operator. Regenerator temperatures dropped and coke buildup on the catalyst occurred. Fig. 29 and 31 shed some interesting light on that phenomenon. At constant Cat/Oil and Air/Oil increasing the coking tendency of the feed will always stabilize an open loop stable steady state. The same is true at constant T_{rgn} and T_{ris}. This is even true for moderate amounts of Conradson Carbon. Conversion will decrease but the unit will remain stable. As will be shown later, this is not always true for other control schemes. Twenty years ago, the conventional control scheme was based on controlling riser top temperature and the temperature differential across the regenerator cyclones. This can lead to wind down. The analysis presented in Fig. 29 and 31 is helpful in differentiating between inherent instabilities and instabilities caused by the control scheme. Adding a coking feed does not lead to inherent open loop de-stabilization.

5. The FCC Control Problem

In this section the partial control approach is applied to the control of the FCC. The first step of the evaluation is to define the reduced model M of Eq. 2 for the process. This was done for the FCC in Section 3 and the appendix (Arbel et al., 1995a). Table 3 gives a list of input and output variables for the FCC. It is immediately clear that the number of manipulated variables available is much smaller than that of outputs to control.

Table 3: List of variables for FCC

Manipulated Variables	Fast	Slow
	Air flow rate Catalyst flow rate Feed temperature Catalyst cooler	Catalyst properties and additives (combustion promoter, etc.) Feed rate Feed quality
Specifications	Product	Constraints
	Octane Conversion Alkylation feed Gasoline yield Light cycle oil yield and properties Isobutene, propylene to propane ratio, C_2^-	CO, NO_x, SO_2 in flue gas Wet gas rate Air rate, Cat. circulation rate Flue gas temperature Riser temperature Regenerator temperature
Model State variables	Riser temperature Regenerator temperature Regenerator wall temperature Coke on regenerated catalyst Coke on spent catalyst	

The list in Table 3 is in no way exhaustive. Specifications and design have evolved with time, increasing the number of both control and manipulated variables.

Feed stocks have continuously changed due to the need to treat heavier crudes and due to hydrotreating. So have catalysts. As the constraints on gasoline spec have changed so have the demands on the product slate. Environmental constraints on CO, NOx and SO_2 emissions were unknown until the late 1960s. These pushed forward the use of catalyst additives to control emissions as well as the move to operate at full CO combustion in the regenerator.

Using the notation it is possible to define the different vectors for the FCC based on Table 3:

U_d={F_{air}, F_{cat}, T_{feed}, Catalyst Cooler Duty}

U_s={Catalyst properties (A,z,additives), F_{feed}, Feed quality and composition}

Y_d={T_{ris}, T_{rgn}, T_{sg}, ΔT, $O_{2,sg}$}

Y_s={Octane, Conv. Y_g, Y_{wg}, C_{rgc}, etc.}

Y_p={all product specs, all operating constraints}

The design has the largest impact on control via the number of manipulated variables available for control. Many old FCC units still have limited ability to vary catalyst circulation rate rapidly and over a wide range (Avidan and Shinnar, 1991; Huq, Morari and Sorensen, 1994). Feed preheaters were also less common twenty-five years ago, but most units are equipped with them now. Catalyst coolers are now slowly introduced.

When the optimization of a unit is considered, it is very important not to limit the control to the primary control structure but to include feed composition and catalyst management. But, before any optimization is applied, the primary control structure is crucial for filtering disturbances and maintaining stability. The focus of this paper is on the choice of this structure. In fact, it is limited to an even more specific example. That is the case common to many plants where the manipulated variables for the primary control structure are air rate and catalyst circulation rate. This eliminates from the evaluation a very critical part of designing a primary control structure namely, the choosing of the manipulated variables. This is best if done during the design of the unit itself, because adding manipulated variables later is usually very expensive. This part is skipped in order to limit this paper to a reasonable length. The methodology of evaluation discussed applies, however, to any manipulated variable entering the structure such as feed temperature. Hopefully, a future paper will discuss the more general problem of the impact of the type and number of manipulated variables available.

Table 4: Possible combinations of manipulated and controlled variables for primary control structures

Regime	**Controlled Variables**	**Manipulated Variables**
Partial CO Combustion	T_{ris}, T_{rgn}	F_{air}, F_{cat}
	$T_{ris}, \Delta T$	
	T_{ris}, T_{sg}	
	$T_{rgn}, \Delta T$	
Complete CO Combustion	$T_{ris}, O_{2,sg}$	

In industry and in literature, different combinations of measured variables have been proposed for this structure when operating in partial CO combustion and are given in Table 4. For steady state control, it is not important how the variables are connected. This is however important in dynamic control, and the conventional pairings are given in Fig. 32. Table 4 and Fig. 32 also contain one combination [T_{ris}, $O_{2,sg}$, F_{cat}, F_{air}] which is used only in complete combustion and is the only one the authors are aware of being used in that regime that is based on using F_{cat} and F_{air} as the manipulated variables.

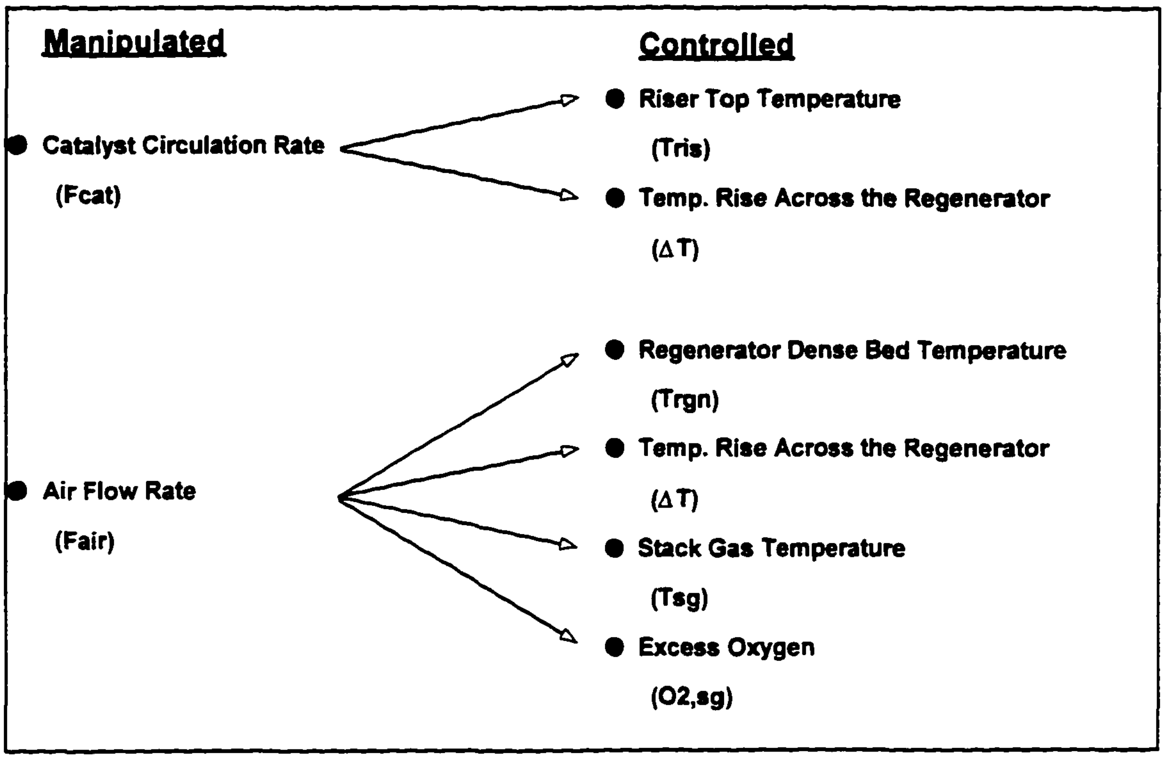


Figure 32: Five possible pairing for primary dynamic control structures

Historically for this type of reactor, the set [ΔT, T_{ris}, F_{air}, F_{cat}] was used first with various definitions of ΔT ((T_{sg} - $T_{rgn(bed)}$), (T_{sg} - $T_{cyclone\ entrance}$), etc.). In partial combustion, ΔT across the cyclones is related to the amount of uncombusted oxygen entering the cyclones. The ΔT used here (T_{sg} - T_{rgn}) is slightly more complex because it contains some afterburn in the dilute phase (see Section 3) but behaves very similarly and the results presented apply to both cases. However, cyclone temperatures are very often not uniform and these results do not apply when hot cyclone temperature is used instead of an average cyclone temperature.

The impetus for the use of ΔT and T_{sg} was two fold. First was the need to protect the cyclones from excess temperatures. When there are problems with air distribution,

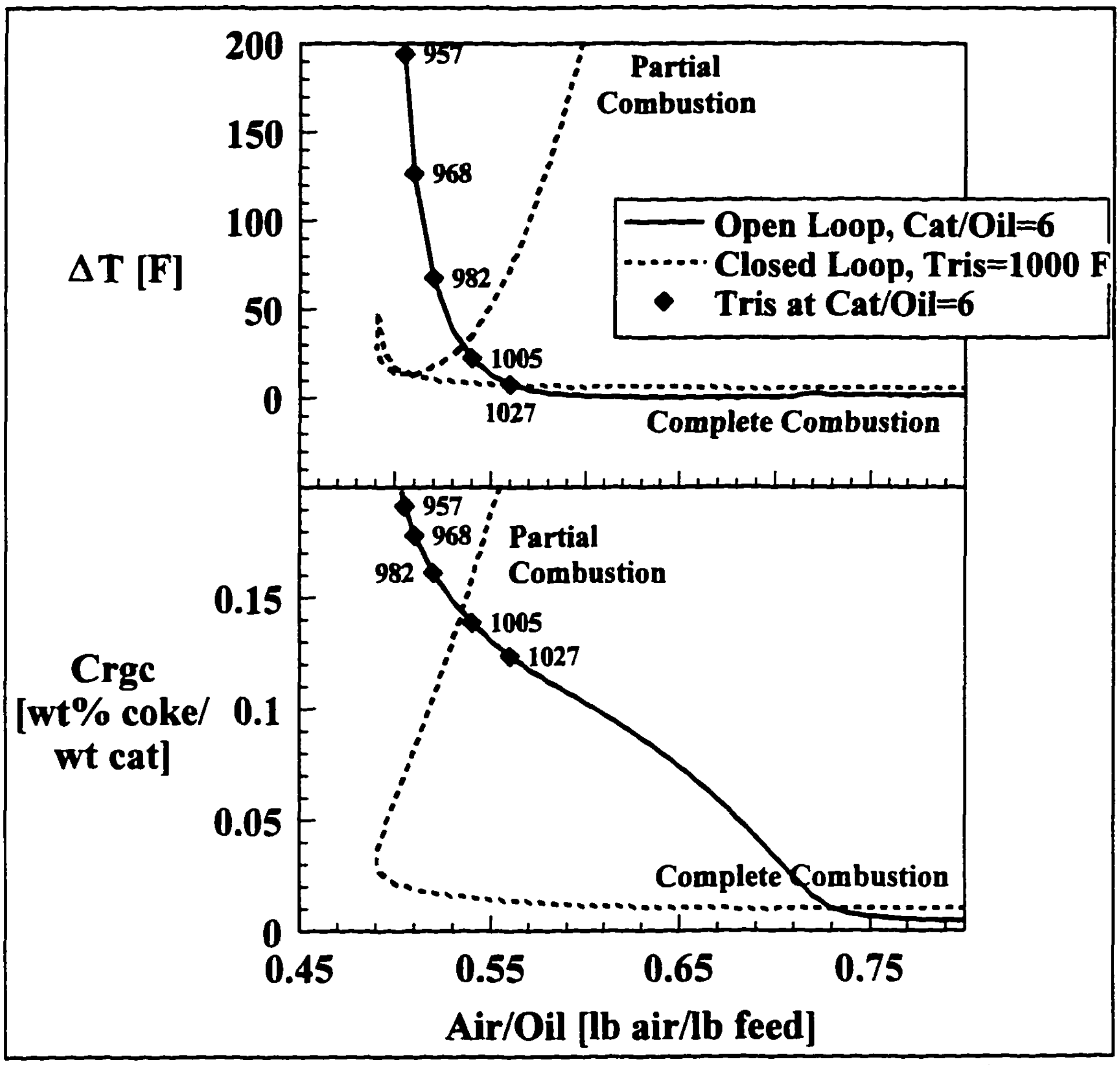


Figure 33: ΔT and C_{rgc} vs. Air/Oil -- Open loop (constant Cat/Oil = 6) and closed loop (T_{ris} = 1000°F)

some cyclones can get hotter than others, and in old designs this could lead to holes in the cyclones. This was be solved by using better materials or by injecting steam into the hot cyclones. The second reason for the use of ΔT and T_{sg} was the belief that a high ΔT is beneficial as it indicates excess oxygen and therefore a cleaner regenerated catalyst. As can be seen from Fig. 33, for operation in partial combustion, increasing air rate (shown as Air/Oil) while F_{cat} (shown as Cat/Oil) is fixed decreases ΔT and reduces C_{rgc}. In closed loop, when T_{ris} is kept constant by manipulating F_{cat}, this trend reverses and now increasing air rate leads to an increase in both ΔT and C_{rgc}, the latter due to a decrease in T_{rgn}. The use of T_{sg} instead of ΔT was suggested for similar reasons but its behavior is more complex because, as was shown in Section 4, it has multiplicities in the region of interest and therefore can, in closed loop, change the sign of the gain depending on the operating point.

The scheme $[T_{rgn}, \Delta T, F_{air}, F_{cat}]$ was suggested by Kurihara (1967) to stabilize the system and was claimed to have advantages for dynamic control. It found no widespread use as control of reactor temperature is very important (Shinnar, 1976). Stability used to be a significant concern, as units frequently crashed in the sense that they lost the upper steady state and started to wind down. This normally occurred when a heavier feed was introduced. It was shown in Section 4 that heavier feeds increase the range of temperatures at which the system is open loop stable. Therefore, open loop instability could not have been the cause of these crashes. It will be later shown that these were nonlinear effects of the choice of the primary control structure.

5.1. Evaluation of the Control Structures for Partial Control

The next step in the evaluation is to examine the control structures combinations of Table 4 in the light of the criteria set in Section 2. This will be done in reference to four sets of operating conditions summarized in Table 5. As stated in the last section the concern is with design compromises, as some structures may have advantages for some criteria and be inferior for others. Because there are strong interactions between the criteria, it is difficult to discuss them completely separately. They will be dealt with as they will arise in the discussion.

Table 5: Four operating points and conditions used in the examples

		Case I	Case II	Case III	Case IV
Stability		Stable	Stable	Unstable	Stable
CO combustion		Partial	Partial	Partial	Complete
Relative Cat. CO Comb. Rate		1	1	1	5
Relative Activity		1	1	1	1
Relative Coking Rate		1	1	1	1
Tfeed	oF	670	670	400	670
Air/Oil	lb air/lb feed	0.5447	0.5639	0.795	0.5141
Cat/Oil	lb cat/lb feed	6.423	7.836	9.265	4.121
Tris	oF	1000	980	980	1000
Tmix	oF	1036	1010	1009	1044
Trgn	oF	1255	1180	1230	1389
ΔT	oF	42	245	243	9
Tsg	oF	1297	1425	1473	1398
O2sg	mol%	0	0	0	1
Conversion	wt% feed	80.2	78.8	83.2	75.8
Gasoline	wt% feed	47.5	49.0	45.4	48.4
Wet Gas	wt% feed	28.6	25.5	32.0	24.1
Crgc	wt% cat	0.171	0.293	0.187	0.019
Csc	wt% cat	0.924	0.938	0.918	0.915
Coke/Feed	lb c/lb oil	4.83	1.19	6.77	3.69

Feed composition: 0.17Ph, 0.2Nh, 0.24Ah, 0.125Rh, 0.14Pl, 0.085Nl, 0.02Al, 0.02Rl

5.1.1. Input Multiplicities and the Constraint of Finite Gain

First, let us review some results of Section 4 (Arbel et al. 1995b) that dealt with multiple steady states and instabilities. Two points are crucial for understanding the control needs:

- Output multiplicities -- For a large range of inputs FCC has three steady states one of which is cold. Sufficiently strong perturbations can crash the unit to that state.
- Input multiplicities -- No combination of two output variables defines a unique steady state. One of the inputs (F_{cat} or F_{air}) has also to be specified for uniqueness.

The input multiplicities for the case [T_{rgn}, T_{ris}] are given in Fig. 25. Note that every operating point in partial combustion has a counterpoint with the same set points of [T_{ris}, T_{rgn}] but at a much higher air rate, while Cat/Oil remains almost constant. As such high air rates are not feasible in a real unit subject to operating constraints, the operating regime in partial combustion has no reachable multiplicities.

But even if a structure would have input multiplicities in the practical range of partial combustion, that does not mean that control would be difficult or that there would be stability problems. Morari (1985) has shown that for a single loop input multiplicities are normally not a problem as in most cases the sign of the gain is different for the two steady states. In a single loop an integral controller cannot stay stable if the sign of the gain changes. However, in a nonlinear system, it is perfectly possible, though much more rare, to have input multiplicities with both states having the same gain sign. For a 2x2 matrix this is even more complex as the steady state can stay stable in special cases when both signs are reversed.

One can not analyze stability problem solely relaying on steady state maps such as that in Fig. 25. There are, however, several other important constraints that one can deduce from such figures. Note in Fig. 25 that there are both a maximum and a minimum. There T_{rgn} is not controllable with F_{air} keeping T_{ris} constant. The lower minimum is outside present interest, but could be of interest if the operation of an FCC with a temperature sensitive catalyst would be of interest. For present operation the upper maximum is important as it is the operating region encountered when operating in full CO combustion. Also note that there are set points combinations which do not correspond to a feasible steady state, a problem discussed in Section 4.

In Section 4, where Fig. 25 is first discussed, the complete range of permissible air rates was given as the topology of the steady states was discussed. Fig. 34 shows only the practical range of air rates for four structures. Also shown is the effect of increasing the coking rate, as changes in the coking tendency of the feed are the most common perturbation in the FCC. Fig. 4 clearly indicates that in full CO combustion the structure [T_{rgn}, T_{ris}] is not useful, even in the area where the gain is finite but small. An increase in coking rate would increase T_{rgn} far above the control capability of the structure. This also refers to the structures [T_{ris}, T_{sg}] and [T_{ris}, ΔT]. The only solution is to use in full combustion the structure [T_{ris},$O_{2,sg}$] where the main purpose is to keep the unit in the full combustion regime. This will be discussed further later.

The structure [T_{ris}, ΔT] has similar properties to [T_{rgn}, T_{ris}]. It has reachable input multiplicities close to the transition to full CO combustion, but there are none in the practical range of partial combustion. [T_{ris}, T_{sg}] has a much more complex structure. It has a minimum right in the normal operating range of partial combustion. This minimum

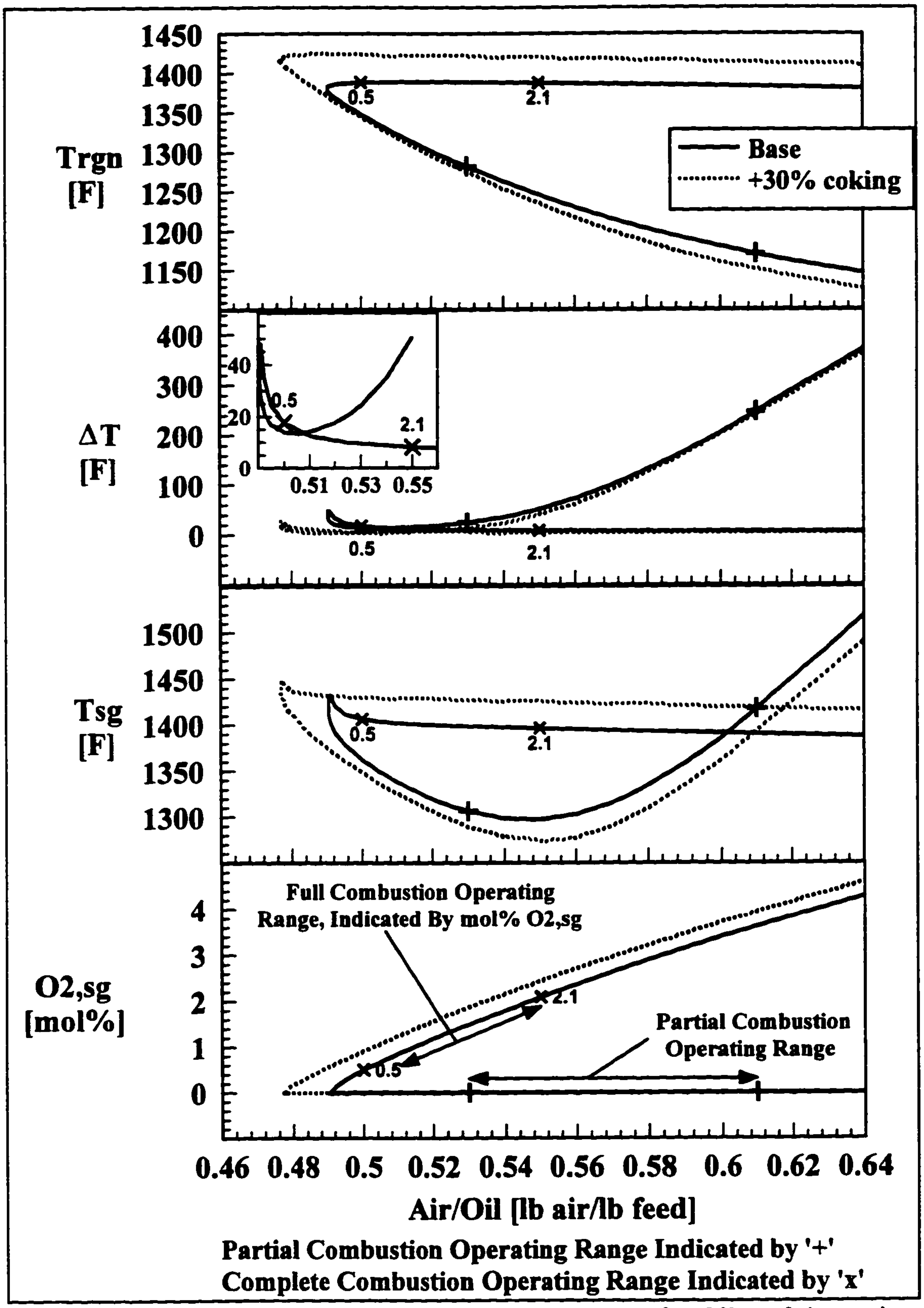


Figure 34: Practical range of cross-cut at constant T_{ris} = 1000°F. Effect of increasing coking rate on multiplicities and closed loop gains.

divides the operating range into two areas that can not be reached from each other by a change in set point without re-tuning the controllers because the gains change sign. Furthermore, a decrease in coking rate may move a set point outside the permissible range. The complex topology of the [T_{ris}, T_{sg}] control structure is due to the definition of T_{sg} as $T_{sg}=T_{rgn}+\Delta T$. This is a combination of two variables that move in opposite direction in a complex way. In general the authors would recommend to be very careful in using such structures.

Fig. 35 gives a typical plant for the Kurihara scheme. For presently common regenerator temperatures (T_{rgn}>1250°F) it also has complex quadruple input multiplicities in the range of present operating conditions. For lower regenerator temperatures (1180 - 1200°F) there is not a problem.

To summarize this section, analysis of the input multiplicities shows that none of the structures using [T_{rgn}, F_{air}] is useful in complete CO combustion. It also shows that the structure [T_{ris}, T_{sg}, F_{air}, F_{cat}] is not useful for partial control if one wants to change operating conditions over a wider range.

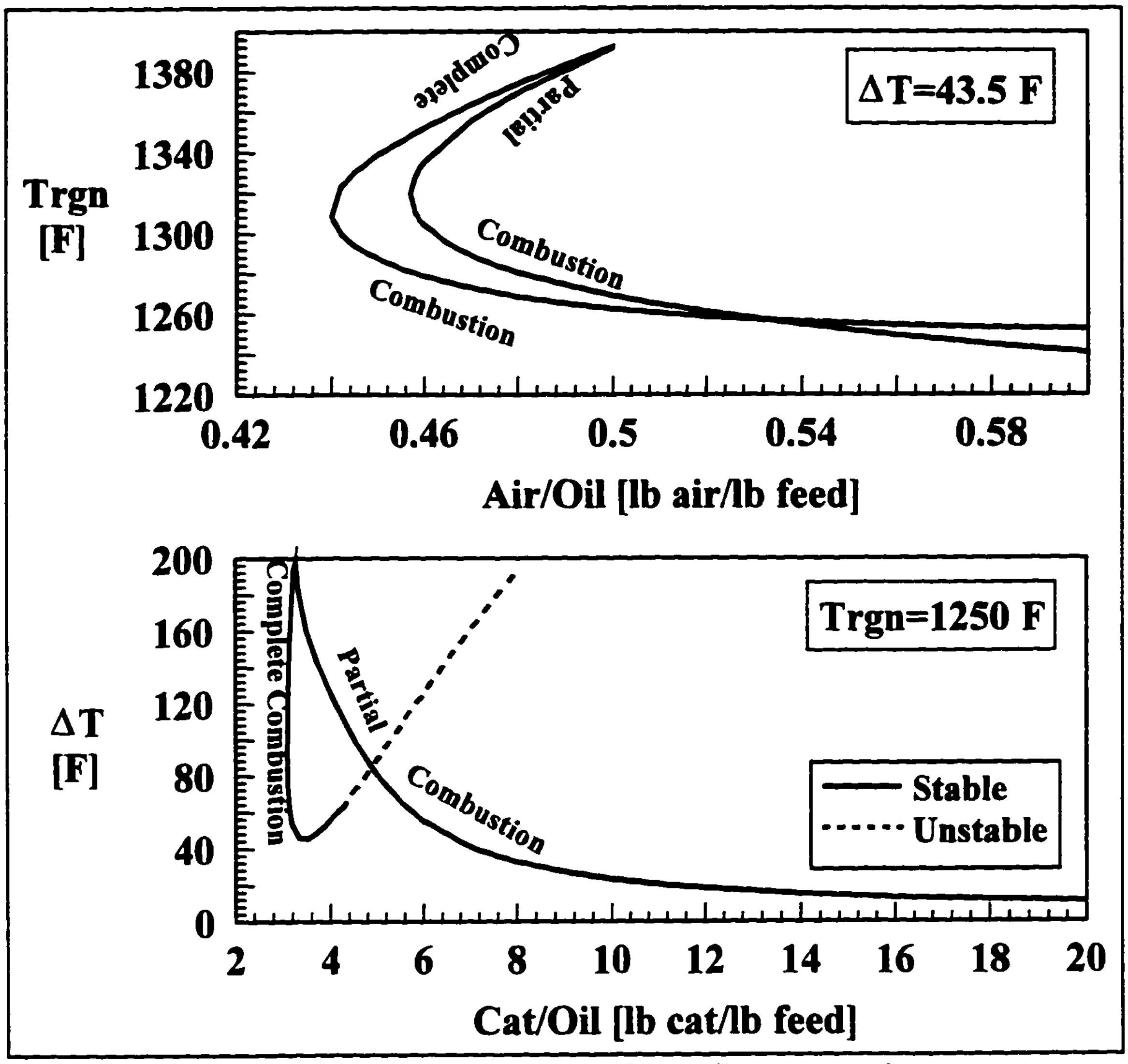


Figure 35: Multiplicities and closed loop gains for Kurihara's control structure: $[(T_{rgn}\text{-}F_{air}),(\Delta T\text{-}F_{cat})]$

5.1.2. Modellability and Dominance

There are two distinct aspects to the problem of modellability in partial control.

- The relation of the vector of product specification and constraints Y_p to the set points chosen.
- Our ability to predict the impact of inputs, process conditions and perturbations on the variables chosen for the set point.

In other words, what information is available for deriving a process model such as given in the appendix (Arbel et al., 1995a). Even a far more simplified form of such a model may sometimes be enough for control. In some cases such models can be obtained for existing units by a set of on-line step experiments. Relation between Y_p and set points can also be obtained by statistical methods. However, for designing or choosing a control structure for a new process a kinetic modelling approach is essential. One should keep in mind that the manipulated variables and their constraints are set during the design phase. But even when deriving what is essentially a statistical relation, it is always preferable and more efficient to base it on a sound kinetic model (Shinnar, 1978).

Since any model is based on data, let us consider what data are available for an FCC and how these are obtained. Both for the reactor and the regenerator, data are obtained either in a small bench scale unit or in pilot plants with input temperature control. Completely integrated adiabatic pilot plants are expensive to operate for data collection. For obtaining proper kinetic data for a model, non-integrated pilot plants that allow us to independently control the state of the catalyst entering the reactor (coke level, activity, etc.), the temperature profile and the space velocity have a significant advantage.

Let us look again at Eq. 6 (or A-2) which is a typical lumped kinetic model for the reactor:

$$\frac{dy_i}{dh} = \tau_r A \rho_c \frac{(1-\epsilon)}{\epsilon} \phi \frac{1}{1+k_h y_{Rh}} \sum_{k=1}^{9} a_{ik} y_k K_{k0} e^{-E_k/RT} \; ; \; y_i(h=0) = y_{i,feed} \qquad (9)$$

Eq. 9 has an interesting property. Catalyst activity (A), coke on regenerated catalyst (ϕ) and space velocity (τ_r) appear together in a term in front of the summation sign. This gives a scale factor determining the advancement along a specific composition trajectory. The direction of trajectory in composition space, which determines final product composition depends on the terms inside the summation sign. Note that the relative magnitudes of the reaction rates at any point are solely a function of inlet composition and temperature. As the activation energies are different for the different rate expressions in Eq. 9, the final composition is a strong function of the temperature history along the riser. The difference between the mixed bottom and the top temperature in a modern riser can be quite large (30-100°F) which has a very significant impact on product distribution. Therefore representing a riser by a stirred tank reactor is insufficient if one wants to properly predict and correlate product composition.

This is an interesting aspect of modellability. It is often not recognized that for control there is a minimum model complexity that strongly depends on Y_p. The model given in the appendix was sufficient twenty years ago but is insufficiently detailed today as Y_p has changed. Today the interest is not only in conversion and gas make, but also in gasoline composition and octane as well as on the composition of the wet gas. These are strong functions of the temperature profile. For our purposes this is not important. All one would have to do to modify the model appropriately is to derive a more detailed reaction rate matrix with appropriate number of lumps having the same form of Eq. 9. This was not done here as no non-proprietary data are available and it is not needed for presenting out ideas.

Looking further at Eq. 9 note that at constant feed conditions and catalyst activity the reactor output is determined by four factors:

1. The term $\tau_r \rho_c (1-\varepsilon)/\varepsilon$ which at constant density and feed rate is solely a function of F_{cat}, which, when used to control T_{ris}, is mainly a function of T_{rgn} and T_{ris}.
2. T_{ris}, which is a set point adjustable by control.
3. T_{mix}, which is a function of T_{feed}, F_{cat} and T_{rgn}.
4. C_{rgc}.

We note that at constant feed composition, feed rate, T_{feed} and catalyst activity the reactor outlet composition is determined by C_{rgc}, T_{rgn} and T_{ris}, where T_{rgn} and T_{ris} have a stronger impact compared to C_{rgc}. While the riser outputs can never be exactly predicted, changing T_{ris} and T_{rgn} independently allows to significantly move the individual elements of Y_p and this feature is called dominance.

Thus higher values of either T_{ris} or T_{mix} will give at constant conversion a higher ratio between wet gas and gasoline, increase the octane of the gasoline and produce more C_4 olefines. While not all of these are in this model it is an important consideration for modern FCC operation. What is in this model is the impact of T_{ris} and T_{mix} on the relative coke make at constant conversion. Therefore, the only way to use laboratory experiments to construct a model that predicts the impact of process conditions on Y_p is based on knowing and controlling T_{rgn} and T_{ris}, which is another aspect of dominance.

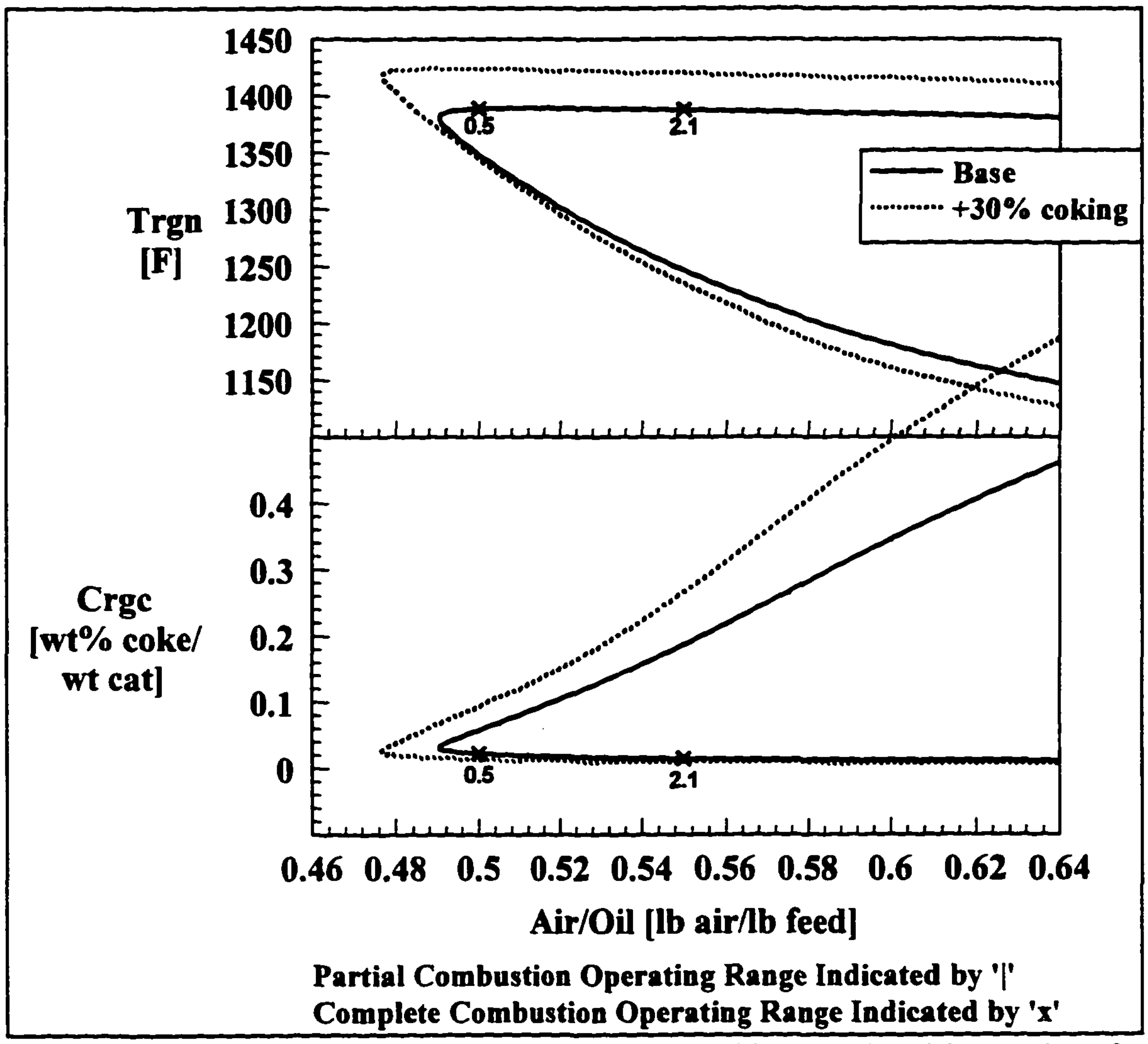


Figure 36: Effect of increasing coking rate on multiplicities and closed loop gains of T_{rgn} and C_{rgc} at constant T_{ris} = 1000°F

Let us now look at the Regenerator. In the regime of partial combustion the heat balance is dominated by the CO_2/CO ratio, which is determined by the relative magnitude of the reaction rates of Carbon and CO combustion (see Section 3). The ratio is, therefore, a function of C_{rgc} and the ratio of the rate constants. The activation energies for Carbon and CO combustion are different. The CO_2/CO ratio at the catalyst surface (the Arthur ratio) is also a function of T_{rgn}. Therefore the CO_2/CO ratio depends on T_{rgn}

and C_{rgc} and these two have a dominant impact on regenerator performance. At constant T_{ris}, both T_{rgn} and C_{rgc} are (in the regime of partial combustion) a unique function of the air rate (Fig. 36) and therefore of each other. C_{rgc} decreases with increasing T_{rgn}. As T_{rgn} is easier to measure and also more dominant in the riser it is normally chosen as a set point. When two dominant variables that impact on Y_p are coupled with each other there is always an advantage in terms of the controllability of Y_p to decouple them. In the case of T_{rgn} and C_{rgc} it requires use of additional manipulated inputs such as a catalyst cooler. This will be discussed in a future paper.

Modeling the regenerator presents a further difficulty. The fluid-dynamics are more complex compared to the riser and have a stronger impact on the performance, especially in large units. The impact of imperfect catalyst mixing is much bigger for ΔT (and therefore T_{sg}) as compared to T_{rgn} and C_{rgc}. In fact, bypass of air and imperfect mixing can reverse the sign of the steady state gain of ΔT. This is shown in Fig. 37 based on a simplified model of a bypass. While in partial combustion ΔT in this model is uniquely related to T_{rgn}, the ability to predict it is much more limited compared to T_{rgn}.

The modellability in complete combustion is much simpler than in partial combustion. Instead of predicting the CO_2/CO ratio in partial combustion, it is enough to know that the hold up of air and solids is enough for complete combustion. The measurement of excess oxygen in a well designed regenerator will guarantee complete CO combustion and almost complete coke combustion (Fig. 36). As the coke on regenerated catalyst is now almost constant and the heat balance directly predictable from the coke make the two variables (C_{rgc} and the CO_2/CO ratio) that caused the complex interactions with the riser are eliminated. This makes it much easier to predict unit performance from

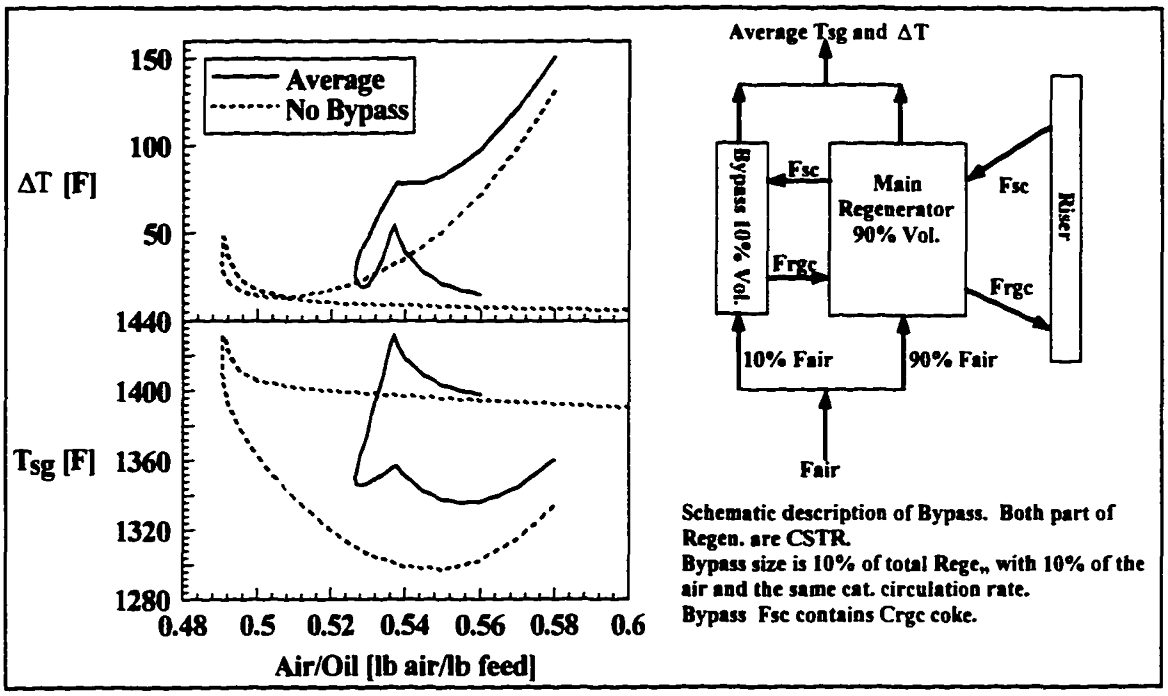


Figure 37: Impact of bypass in the regenerator on multiplicities and closed loop gains of ΔT and T_{sg} at constant T_{ris} = 1000°F.

laboratory experiments. However, the ability to control T_{rgn} by air rate is also lost as was noted in the previous section (Fig. 34). As T_{rgn} is still a dominant variable for the reactor, this limits the capabilities to control the reactor performance. This will be discussed later.

Another aspect of modellability is the permissibility of a given set of set points. This has several aspects that were discussed in Section 4 (Arbel et al., 1995b):

- There may not exist a stable steady state for the set points chosen by the operator in which case the unit will either crash to the cold steady state or move to a new steady state with different T_{rgn} and T_{ris} than those chosen, with the manipulated variables staying at their constraint values.
- The steady state is permissible but the manipulated variables exceed their constraints, a situation that can be corrected by reduction of the feed rate.
- The unit may become unstable, either due to change in feed or catalyst properties

or by choosing set points that have this property.

- Another constraint is violated which requires change of set points or other action.

The range of permissible set points was discussed in Fig. 30 that gives an example for the limits for permissible set points for [T_{rgn} , T_{ris}] for a specific case. While these limits are not very accurate due to model uncertainties and a lack of knowledge of exact catalyst and feed properties, such plots give guidelines for the operator. Similar plot plots for [ΔT, T_{ris}] or [T_{sg} , T_{ris}] were not given as the model uncertainties are far too large to construct such plots from available data.

There is yet one more aspect of modellability that merits more detailed discussion than given here. That is the accuracy of the model and its sensitivity to changes in feed and catalyst properties. An example was given in Fig. 37 where the strong sensitivity of ΔT to bypass was shown. Another example is given in Fig. 38 where Fig. 34 is given again comparing two different combustion rates in the regenerator. The impact of a 30% increase of the relative coking rate was already shown in Fig. 34. Note that in partial combustion the effect in both cases is minimal for T_{rgn}, much more profound for ΔT, T_{sg} and C_{rgc}. The change shown in Fig. 38 leads to a reversal of the gain (and RGA) for the case of T_{sg}. Changes in the catalytic rate of CO combustion due to metal impurities have a similar effect. This sensitivity would completely mess up any empirical correlations between Y_p and [T_{ris}, ΔT]. Fig. 36 and 38 also show that C_{rgc} is quite model sensitive as well, which is another reason for not using it.

Fig. 34, 36, 37 and 38 present different types of model uncertainties that one encounters in controller design. One is that the parameters of the model change when operating conditions change leading to a change in the gains and time constants. This

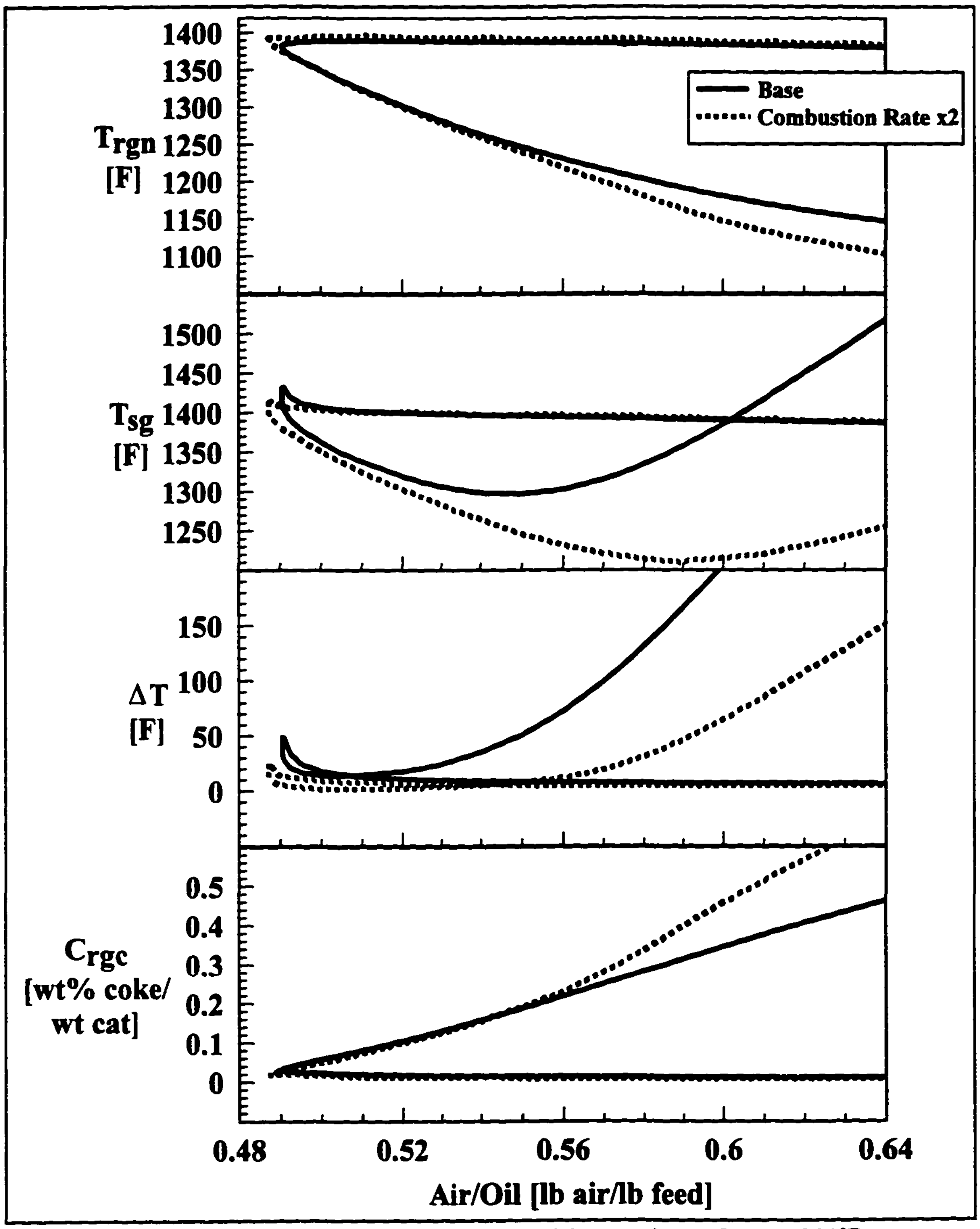


Figure 38: Sensitivity of multiplicities and closed loop gains at T_{ris} = 1000°F to combustion rates in the regenerator.

present a problem in partial control when it is desirable to change set points and process constants without having to re-tune every time one changes set points. This should be taken into account in the tuning. Fig. 37 and the case of T_{sg} in Fig. 34 present a more difficult problem as the transfer function itself has large uncertainties that can reverse gains.

5.1.3. Evaluation of Control Structures for Partial Combustion

In the last two sections several aspects of the control structures such as the impact of the nonlinear nature of the FCC and modellability and dominance were evaluated. For the regime of partial combustion a clear advantage for the pair [T_{ris}, T_{rgn}] was noted in the above discussion. In fact for the concept of partial control, namely the use of set points to change Y_p, [T_{ris}, T_{rgn}] is the obvious choice. However, this does not mean that [T_{ris}, T_{rgn}, F_{cat}, F_{air}] is an adequate or sufficient control structure. In fact, as was pointed out in Section 2 and Table 2, this structure is an anchor in the overall control strategy which also uses other more slowly manipulated variables. These, T_{feed}, catalyst activity, feed composition, etc., are used only for the steady state control. Therefore, some other aspects of the structure have to be evaluated:

1. How efficient is it in changing Y_p?
2. How well does it deal with common disturbances and can it maintain the unit stable?
3. Is the response fast enough or should it be speeded up by some inferential control using other measured variables?
4. How well can the controller perform dynamically over a wide range of set points and conditions without changing the tuning?
5. How well does it integrate with the overall control strategy using the slow

manipulated variables in Table 1 which are or should be used in any optimization or control strategy.

5.1.3.1. Effect on Y_p

The effect of changing the set points of T_{rgn} and T_{ris} is given in Table 6 for two base cases (Case I and II of Table 5). Note the strong impact on Y_p which can be seen both from the changes in set points and from comparing the two base cases. As explained before, the impact is even bigger as T_{rgn} and T_{ris} also effect octane and wet gas production. However it has to be admitted here that any of the other control schemes have a range of set points for which the ability to control Y_p by changing the set points is no less. This is due to the fact that in the range of partial combustion for which a given pair of set points leads to a unique steady state, fixing [T_{ris}, ΔT] is equal to fixing [T_{ris}, T_{rgn}]. There are two limitations to this statement. The range of operating conditions for which this applies is more limited due to the input multiplicities for [T_{ris}, T_{sg}] and [T_{rgn}, ΔT] compared to [T_{rgn}, T_{ris}] (Fig. 34). But even more important, a lot is lost in terms of modellability as all the model predictions for Y_p require knowledge of T_{rgn} and T_{ris}. This would be less of a problem if ΔT or T_{sg} would give a good inferential estimate of T_{rgn}. That this is not so is a special problem of the FCC. As was shown in Section 5.1.2 ΔT and therefore T_{sg} are very model sensitive and much more sensitive than T_{rgn} to changes in operating condition. This limits the use of ΔT as an inferential estimate for T_{rgn}. As these changes in ΔT and T_{sg} have little impact on Y_p there is no benefit in developing a better method to estimate them. One can just as well measure T_{rgn} directly. While ΔT can be controlled at fixed operating conditions, it is not easily integrated into

Table 6: Control action needed for change in set points and resulting change in output for the structure [T_{ris}, T_{rgn}, F_{air}, F_{cat}]. Changes are in units from s.s.

		Set point Change:	Tris + 10	Trgn + 10
Case I of Table 5:				
Air/Oil	0.5447	lb air/lb feed	0.0183	-0.0057
Cat/Oil	6.423	lb cat/lb feed	0.435	-0.253
Tris	1000	oF	10	0
Tmix	1036	oF	1	-2
Trgn	1255	oF	0	10
ΔT	41.8	oF	-5.2	-8
Tsg	1297	oF	-5.3	1.8
O2,sg	0	mol%	0	0
Conversion	80.3	wt%	1.49	-0.3
Gasoline	47.5	wt%	-1.22	0.13
Wet Gas	28.6	wt%	-0.36	0.84
Crgc	0.1711	wt%	0.018	-0.0167
Csc	0.9238	wt%	0.0018	-0.0014
Coke/feed	4.83	lb c/lb oil	0.22	-0.10
Case II of Table 5:				
Air/Oil	0.5639	lb air/lb feed	0.0179	-0.0091
Cat/Oil	7.839	lb cat/lb feed	0.625	-0.371
Tris	980	oF	10	0
Tmix	1010	oF	9	1
Trgn	1180	oF	0	10
ΔT	245	oF	-20	-42
Tsg	1425	oF	-20	-32
O2,sg	0	mol%	0	0
Conversion	78.8	wt%	1.5	-0.1
Gasoline	49	wt%	-1.0	0.0
Wet Gas	25.5	wt%	2.3	0.2
Crgc	0.2927	wt%	0.0264	-0.03
Csc	0.9376	wt%	0.0045	-0.0037
Coke/feed	5.06	lb c/lb oil	-0.17	0.21

an overall control strategy to take care of changing operating conditions, feed stocks and process goals.

5.1.3.2. Disturbance Rejection

Table 7 gives the response of the control at one set point to an increase of 30% in the coking rate for all four structures. This illustrates another point. In the past one aspect of main concern in FCC control in partial combustion was the fear of a crash. While the detailed data are proprietary, there is one common trend in these occurrences. They always occurred when a feed with higher coking tendency was introduced either intentionally or accidentally. This seems strange at first as the analysis shows that more coke would always stabilize the system both in open loop and with fixed [T_{rgn}, T_{ris}].

Table 7: Comparison of different control structures for partial combustion -- Control action needed and changes in outputs after 30% increase in coking rate. Change in units from steady state.

	Case II	Control:	Open Loop	Tris Trgn	Tris Tsg	Tris ΔT	ΔT Trgn	Tris Crgc
Air/Oil	0.5639	lb air/lb feed	0	-0.0161	0.0082	0.0022	-0.1041	-0.0316
Cat/Oil	7.836	lb cat/lb feed	0	-0.177	1.143	0.812	-2.656	-0.99
Tris	980	oF	17	0	0	0	-55	0
Tmix	692	oF	15	0	-17	-13	0	13
Trgn	1180	oF	21	0	-29	-23	0	23
ΔT	245	oF	-156	-85	30	0	0	-147
Tsg	1425	oF	-135	-85	0	-23	0	-123
Conversion	79	wt% feed	-2.2	-4.8	-4.6	-4.6	-15.5	-5.2
Gasoline	49	wt% feed	-0.1	0.6	0.7	0.7	-0.8	0.5
Wet Gas	25	wt% feed	-2.4	-5.5	-5.4	-5.4	-13.9	-5.6
Crgc	0.2927	wt% cat	0.0522	0.0672	0.1671	0.1431	-0.0681	0
Csc	0.9376	wt% cat	0.0827	0.085	0.1066	0.1008	0.0637	0.0741
Coke/Feed	5.05	lb c/lb oil	-0.99	0.02	0.19	0.16	-1.03	-0.13

Table 7 gives a potential answer to this problem of crashing. At fixed T_{rgn} and T_{ris} the change in F_{air} (Air/Oil) and F_{cat} (Cat/Oil) are minimal. One is unlikely to run into a constraint in F_{air} or F_{cat}. Conversion and wet gas production are reduced. One could compensate for the reduction by changing the set points. In the case of constant ΔT or T_{sg}, T_{rgn} decreases drastically. As an operator usually watched T_{rgn}, he would intervene manually for a strong decrease in T_{rgn}. If he would increase air rate he would further decrease T_{rgn} and crash the unit. It was also observed that C_{rgc} increased, an effect called "snowballing" (Edwards and Kim, 1988). Table 7 shows that this happens in ΔT control when coking rate increases. This is obvious from Fig. 38 as an increase in F_{air} at constant T_{ris} always increases C_{rgc}. The case [T_{rgn}, ΔT] which was introduced to stabilize the system, causes a real upset. The changes in F_{air} and F_{cat} are large and could lead to saturation of F_{air}. Further, T_{ris} drops drastically, becoming uncomfortably close to the lower limit of operability for the riser. This case of [T_{rgn}, ΔT] control is a dramatic example of why one cannot use purely linear considerations in the choice of a control structure. Also given is Table 7 are the results for the structure [T_{ris}, C_{rgc}]. Note that T_{rgn} increases significantly which reduces conversion. As T_{rgn} is more dominant and more modellable than C_{rgc}, T_{rgn} is preferable for disturbance rejection.

Table 8 gives similar results for a second case limited to [T_{ris}, T_{rgn}]. The control is very satisfactory. Some disturbances can lead to problems while controlling the unit via this structure. If there is a strong decrease in catalyst activity or coking rate, the system can become unstable or the settings of T_{rgn} and T_{ris} can become non-permissible. An experienced operator will recognize this but proper diagnostic methods would here be desirable but are outside the scope of this work. There is no two dimensional control structure that can eliminate this problem. Increasing T_{rgn} in most cases stabilizes the system.

Table 8: Control action needed and changes in outputs after a 30% increase in coking rate -- Impact of using slow manipulated variables.

			+30% in Coking and:		
	Case I		(only coking)	+30% Activity	-70F Tfeed
Air/Oil	0.5447	lb air/lb feed	-0.0055	0.003	0.0603
Cat/Oil	6.423	lb cat/lb feed	-0.14	-0.111	0.647
Tris	1000	oF	0	0	0
Tmix	1036	oF	-3	-2	-4
Trgn	1255	oF	0	0	0
ΔT	42	oF	-18	-28	-10
Tsg	1297	oF	-18	-28	-10
Conversion	80	wt% feed	-4.3	-3.4	-2.3
Gasoline	48	wt% feed	1.2	1.2	1.0
Wet Gas	29	wt% feed	-5.7	-4.8	-3.8
Crgc	0.1711	wt% cat	0.0475	0.1071	0.051
Csc	0.9238	wt% cat	0.0796	0.1677	0.0782
Coke/Feed	4.83	lb c/lb oil	0.10	0.30	0.68

Today this is less of a problem as feed stocks are heavier and catalysts more active. However, in some cases feed stocks have been hydrotreated and have therefore lower coking rates. One best takes care of such stability problems by using one of the slow manipulated variables, e.g., increasing catalyst activity or adding a promoter to increase the heat supplied by the coke.

In Table 8 also given are some examples to illustrate the impact of some slow variables. The impact of feed composition was already discussed. Also shown are the impact of increasing catalyst activity as well as reducing feed temperature after an increase in coking rate. The fact that T_{rgn} and T_{ris} can be controlled independently over a significant range allows together with control of feed temperature and catalyst activity to change not only Y_p over a wide range but to take care of a wide range of feed stocks albeit with some compromises as to conversion and yield.

Comparing Cases I and II and their perturbations also shows another interesting aspect of the control. In Case I (higher T_{rgn} and T_{ris} the conversion increased while gasoline yield decreased resulting in higher wet gas production. This can be beneficial if there is enough capacity in the wet gas compressor, gas plant, and alkylation plant. In other cases the opposite may be true. Increasing catalyst activity further accentuates this effect. If a large fraction of the gasoline is cracked, octane increases since the aromatics in the gasoline crack faster than paraffins. Tables 6, 7 and 8 give a good illustration of what is meant by partial control and dominance.

5.1.3.3. Stability and Time Response

The evaluation of the different control structures in the previous sections give a clear advantage to the structure $[T_{ris},T_{rgn},F_{air},F_{cat}]$ over the other three structures. It has a clear advantage in modellability and dominance and it is the only structure for which one can clearly relate the settings to desired process outputs. It also has advantages with regard to multiplicities.

For these reasons let us focus on the structure $[T_{ris},T_{rgn},F_{air},F_{cat}]$. If the system cannot be adequately stabilized or perturbations that are normally encountered cannot be taken care of, this structure should be rejected despite its other advantages. As said before, linear methods can be used as long as the different steady states over the expected operating regime in partial combustion are considered. The problems encountered in evaluating stabilization and time response are really not specific to partial control and would occur in any 2x2 control problem. However, the results of the evaluation are important for evaluating the sufficiency of the structure. Including here all four structures would make the paper unnecessarily long especially as it will be shown that the preferred structure is fully adequate.

Let us start with the linerized transfer function given in Table 9 and the controller gain limits given in Table 10. Also given in Table 10 are the RGA for different configurations. Note that the RGA's in Table 10 have a negative diagonal for all the control structures considered for partial combustion.

The conventional approach to design (Bristol, 1966; Shinskey, 1967; Shinskey, 1984; McAvoy, 1983, Hovd and Skogestad, 1994) would give preference to the pairing [(F_{air}-T_{ris}), (F_{cat}-T_{rgn})] along the positive diagonal. However, conventional practice in FCC operation has always been [(F_{air}-T_{rgn}), (F_{cat}-T_{ris})] along the negative RGA diagonal. This introduces a stability problem. If the loop [T_{rgn}, F_{air}] fails or is put on manual after it was tuned with positive gain the system becomes unstable. This has been demonstrated in a commercial FCC by Edwards and Kim (1988) and was discussed in Section 4. One can however live with that as such failures are rare and, fortunately, the system is quite slow to respond, giving the operator ample time to intervene. In general, in adiabatic reactors a competent designer will introduce a slow mode providing sufficient thermal inertia to give the operator time to respond.

Table 9: Transfer functions for the control structure $[T_{ris}, T_{rgn}, F_{air}, F_{cat}]$ at two points in partial combustion.

	Case I of Table 5 : Stable Point, T_{rgn}=1255°F, T_{ris}=1000°F, T_{feed}=670°F	Case III of Table 5 : Unstable Point, T_{rgn}=1230°F, T_{ris}=980°F, T_{feed}=400°F
$\frac{T_{rgn}(s)}{F_{air}(s)}$ $\left[\frac{°F}{lb\ air/lb\ feed}\right]$	$\frac{1.5(0.0003+s)(0.005+s)(0.0084+s)(0.010+s)}{(0.00029+s)(0.00085+s)(0.0047+s)(0.0093+s)(0.01+s)}$	$\frac{1.3(0.0003+s)(0.0069+s)(0.012+s)(0.014+s)}{(-0.00025+s)(0.00032+s)(0.0074+s)(0.0137+s)(0.015+s)}$
$\frac{T_{rgn}(s)}{F_{cat}(s)}$ $\left[\frac{°F}{lb\ cat/lb\ feed}\right]$	$\frac{-0.13(0.0003+s)(0.0095+s)(0.0047 \pm i\ 0.0013+s)}{(0.00029+s)(0.00085+s)(0.0047+s)(0.0093+s)(0.01+s)}$	$\frac{-0.12(0.0003+s)(0.0036+s)(0.0081+s)(0.0139+s)}{(-0.00025+s)(0.00032+s)(0.0074+s)(0.0137+s)(0.015+s)}$
$\frac{T_{ris}(s)}{F_{air}(s)}$ $\left[\frac{°F}{lb\ air/lb\ feed}\right]$	$\frac{0.86(0.0003+s)(0.0055+s)(0.0084+s)(0.0099+s)}{(0.00029+s)(0.00085+s)(0.0047+s)(0.0093+s)(0.01+s)}$	$\frac{0.86(0.0003+s)(0.0079+s)(0.012+s)(0.014+s)}{(-0.00025+s)(0.00032+s)(0.0074+s)(0.0137+s)(0.015+s)}$
$\frac{T_{ris}(s)}{F_{cat}(s)}$ $\left[\frac{°F}{lb\ cat/lb\ feed}\right]$	$\frac{22(-0.00086+s)(0.00031+s)(0.0053+s)(0.0084+s)(0.0096+s)}{(0.00029+s)(0.00085+s)(0.0047+s)(0.0093+s)(0,01+s)}$	$\frac{18(-0.0019+s)(0.00031+s)(0.0075+s)(0.012+s)(0.014+s)}{(-0.00025+s)(0.00032+s)(0.0074+s)(0.0137+s)(0.015+s)}$

Table 10 : Some linear properties of different control structures.

Steady State Gains and RGA for Case I of Table 5:

Case I	**Steady State Gains**		**RGA**	
	Air/Oil [°F/(lb air/lb feed)]	Cat/Oil [°F/(lb cat/lb feed)]	Air/Oil	Cat/Oil
T_{rgn}	1716	-74	-1	2
T_{ris}	1071	-22	2	-1
T_{sg}	271	-25	-0.3	1.3
T_{ris}	1071	-22	1.3	-0.3
ΔT	-1445	48	-1.6	2.6
T_{ris}	1071	-22	2.6	-1.6
T_{rgn}	1716	-74	-3.6	4.6
ΔT	-1445	48	4.6	-3.6

Proportional Controller Limits, K=KcGp(0) (conventional negative feedback control):

	Case I of Table 5: Stable	**Case III of Table 5: Unstable**
T_{rgn}/F_{cat}	K>-1.0	K>1.0
T_{rgn}/F_{air}	K>-1.0	K>1.0
T_{ris}/F_{cat}	0.98>K>-1.0	5.9>K>1.0
T_{ris}/F_{air}	K>-1.0	K>1.0

With 2% control action in Air/Oil and 10% in Cat/Oil the maximum temperature gain is:

	Air/Oil [°F]	Cat/Oil [°F]
T_{rgn}	34	7.4
T_{ris}	21	2.2

We are not aware in industry of anybody using the positive diagonal pairing. While there are many historic reasons for the present scheme the positive diagonal pairing still merits evaluation. It has two disadvantages:

1. The first is modellability. It is counter-intuitive. The central concern of the control is the reactor. The laboratory data that underlie the model and control strategies are based on decoupled, stand alone laboratory units in which the reactor is fed a catalyst of fixed temperature and coke level. Intuitively it would be preferred to control the regenerator temperature by using the air flow as manipulated variable. This partially decouples the reactor from the regenerator. The only coupling left is through the coke level on regenerated catalyst. In modern operating conditions ($T_{rgn} > 1230°F$) this level is kept low minimizing the interactions. On the other hand, controlling the reactor with air flow involves complex interactions between regenerator and reactor that are hard to model accurately. Obviously for steady state control this makes no difference, but again it is difficult to think one way and tune loops another way.

2. The scheme along the positive diagonal does not allow fast control of T_{ris}. As the holdup of the reactor is low compared to the regenerator's the decoupling using the conventional control scheme allows conceptually fast and easy control of the reactor. The regenerator provides the slow time scale needed to help an adiabatic system with multiple steady states and allows the operator to make adjustments without crashing. Modern reactors are designed with short residence time and respond very fast. This was not always so as some other designs had similar time scale for both reactor and regenerator.

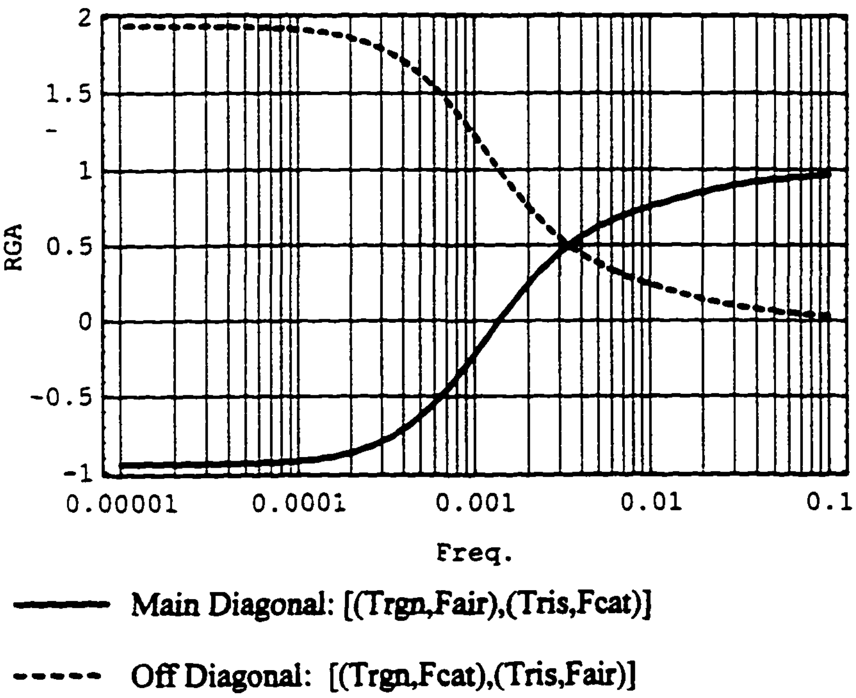


Figure 39: Dynamic RGA for the control structure [T_{ris}, T_{rgn}, F_{cat}, F_{air}]. All conditions are of Case I of Table 5.

As the emphasis in control is on the reactor in terms of product properties and conversion, a tight control (1 to 2°F) on it is desired. The regenerator dominates stability considerations. As it responds slowly fast control is not needed nor desired. There is no problem if T_{rgn} changes over a 20°F range. The fact that it is hard to achieve this goal with the positive diagonal scheme is directly apparent from the dynamic RGA shown in Fig. 39. Note that the RGA for the positive diagonal changes to zero at higher frequencies where the negative diagonal goes to unity. The RGA does not give information about the time response of each separate loop, but the information in Fig. 39 coupled with the knowledge about the response of the decoupled reactor to F_{cat} is sufficient to indicate that the chance of fast control of T_{ris} is better along the negative diagonal.

In terms of stability the positive diagonal pair has no problem. The transfer functions have an overall behavior that is approximately first order, and the constraints on control action are dominated by avoidance of saturation and oscillatory response. Knowledge of the RGA is, however, insufficient to show if the controller can stabilize an unstable steady state. Such cases do not arise intentionally but due to changes in coking rate or catalyst activity the gains change. Given in Fig. 40 are stability limits for two steady states with different gains, one stable and one linearly unstable. The stable steady state was chosen for normalization. It is asked that the control structure can stabilize well an unexpected change to an unstable state. The controller along the positive diagonal is a perfectly reasonably controller. The fact that the negative diagonal pairing has in this case decisive advantages due to the reasons listed above is very specific to the FCC. The only conclusion that can be generalize is that care should be taken in setting up rules. Design is often based on compromises between conflicting criteria.

In Fig. 40 the controller constants K were normalized by the steady state gains to reflect conventional negative feedback controllers. To judge what K's are permissible one has to look at saturation constraints. Most present FCC's (but not all) operate very close to the air constraints and allow no more than 2% for dynamic control. In the past catalyst flow rate was often the constraint. At present a 10% leeway in catalyst circulation rate

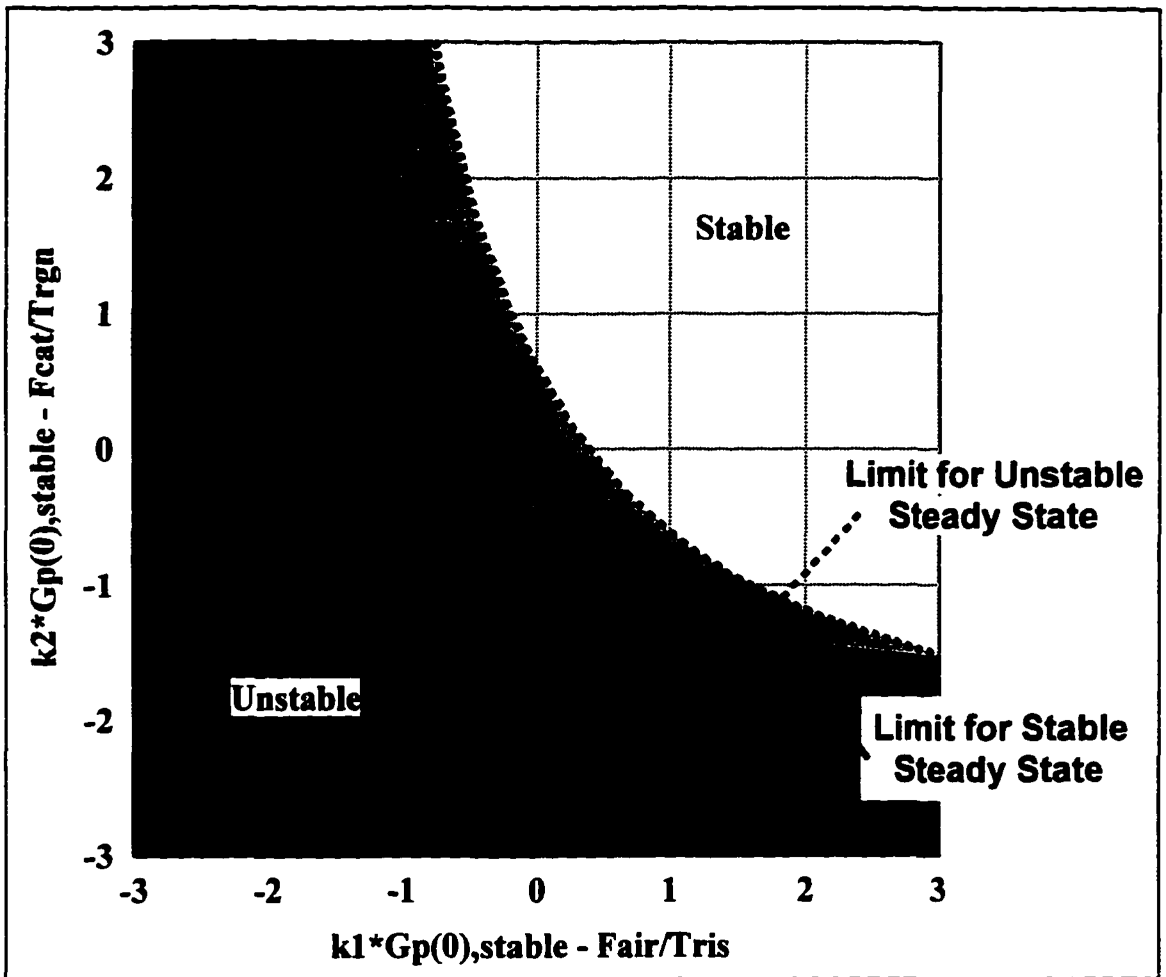


Figure 40: Stability limits change due to transition from stability to instability for the control structure $[(T_{ris}\text{-}F_{air}),(T_{rgn}\text{-}F_{cat})]$ with a P controller.

for dynamic control is reasonable. The maximum available steady state compensation in temperature is given in Table 10. The ratio between the maximum compensation and the derived range of temperature to be controlled dynamically gives an estimate as to what K's are permissible. This is true for both conventional paring and paring along the positive diagonal.

For integral control based on the positive diagonal pairing, changing either one or both of the signs of the controller's gains will make the system unstable. For the control along the negative diagonal, the sign of the controller gains for negative feedback control

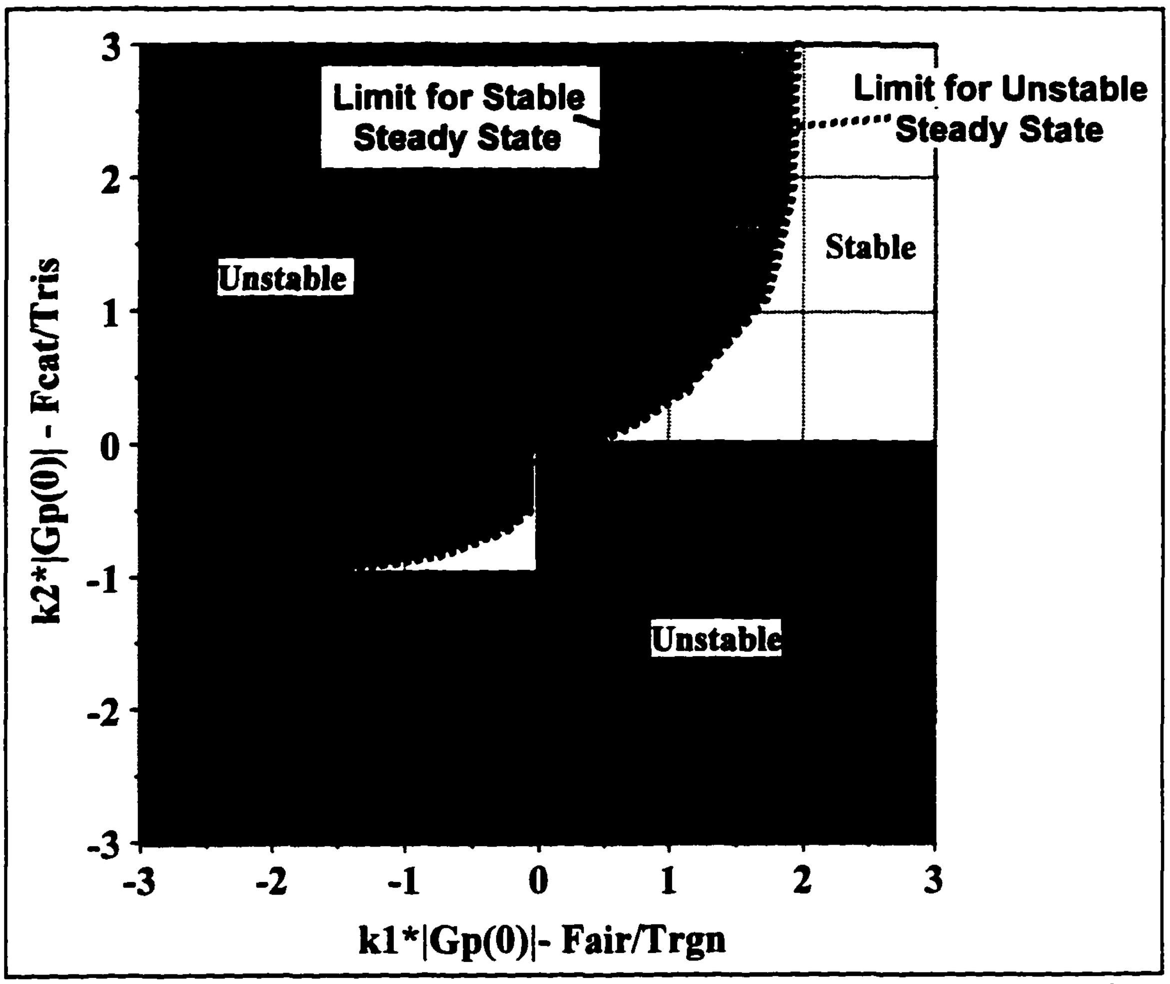


Figure 41: Stability limits for the control structure [(T_{ris}-F_{cat}),(T_{rgn}-F_{air})] at Case I of Table 5 with P and PI controllers (PI=Ki(1+0.001/s)).

will be determined on which loop is closed first. Shown in Fig. 41 are the stability limits for the conventional paring based on the negative RGA. The system remains stable with integral control if both signs are reversed. Negative feedback control has here no clear meaning. However, there are two different sign combinations with totally different properties. Unfortunately, use of the RGA offers no help as it is identical for both choices. For the structure [(F_{air}-T_{rgn}), (F_{cat}-T_{ris})] closing the air loop first is obvious intuitive choice, based on the same reasoning used for choosing the negative diagonal. The emphasis is on the reactor so it is better to decouple it. At constant T_{rgn} it becomes much easier to control and intuitively it is assumed that increasing catalyst flow will increase reactor temperature. It is counter-intuitive that with an uncontrolled regenerator the steady state gain for the loop [T_{ris}, F_{cat}] has a negative sign. It is the result of a complex interaction. It makes therefore sense to close the air loop first. There are other reasons. There is a lot of experience available with FCC's where the only dynamic control is on the regenerator air control. This was common in older units such as Model IV. Units with fast acting slide valves came later. It made sense to use the experience from other units. Since then it was learned that it is hard to control an FCC with fixed air, manipulating only F_{cat} even if it is desirable to do so to maximize production rate.

There is another intuitive reason why choosing positive gain for both circuits is preferable. The dynamic RGA indicates that control along the negative diagonal allows complete decoupling at high frequencies. To utilize these for fast reactor control it is obvious that the sign of the fast initial response of T_{ris} with F_{cat} (which is positive) must be used. It turns out that reversing the sign (closing the reactor first) also has other problems as can be seen from Fig. 41. There is an obvious problem with the signs normalization in such figures. However, for this specific example there is no problem with the symmetries of the gains because at this steady state the RGA is -1.0. That means that closing one loop first reverses the sign but retains the same magnitude for the gains. Therefore, the normalization of the sign for the negative diagonal pairing is based on the loop normally closed first. Note from Fig. 41 that the stability limits for negative K (reactor closed first) are much narrower. There is a simple explanation for this. The open loop transfer function for T_{ris}/F_{cat} has a right hand side zero, a 180° phase lag at zero frequency. This strongly limits control action. Fig. 42 gives the frequency responses of T_{ris}/F_{cat} both open loop and with the loop T_{rgn}/F_{air} closed with reasonable integral action.

Note that the phase lag disappears. This is also apparent from Fig. 41. If K_{rgn} is sufficiently large then there are no limits on K_{ris}. Also shown is the open loop response of T_{rgn}/F_{air}. It is essentially a first order lag that comes from the thermal inertia of the regenerator.

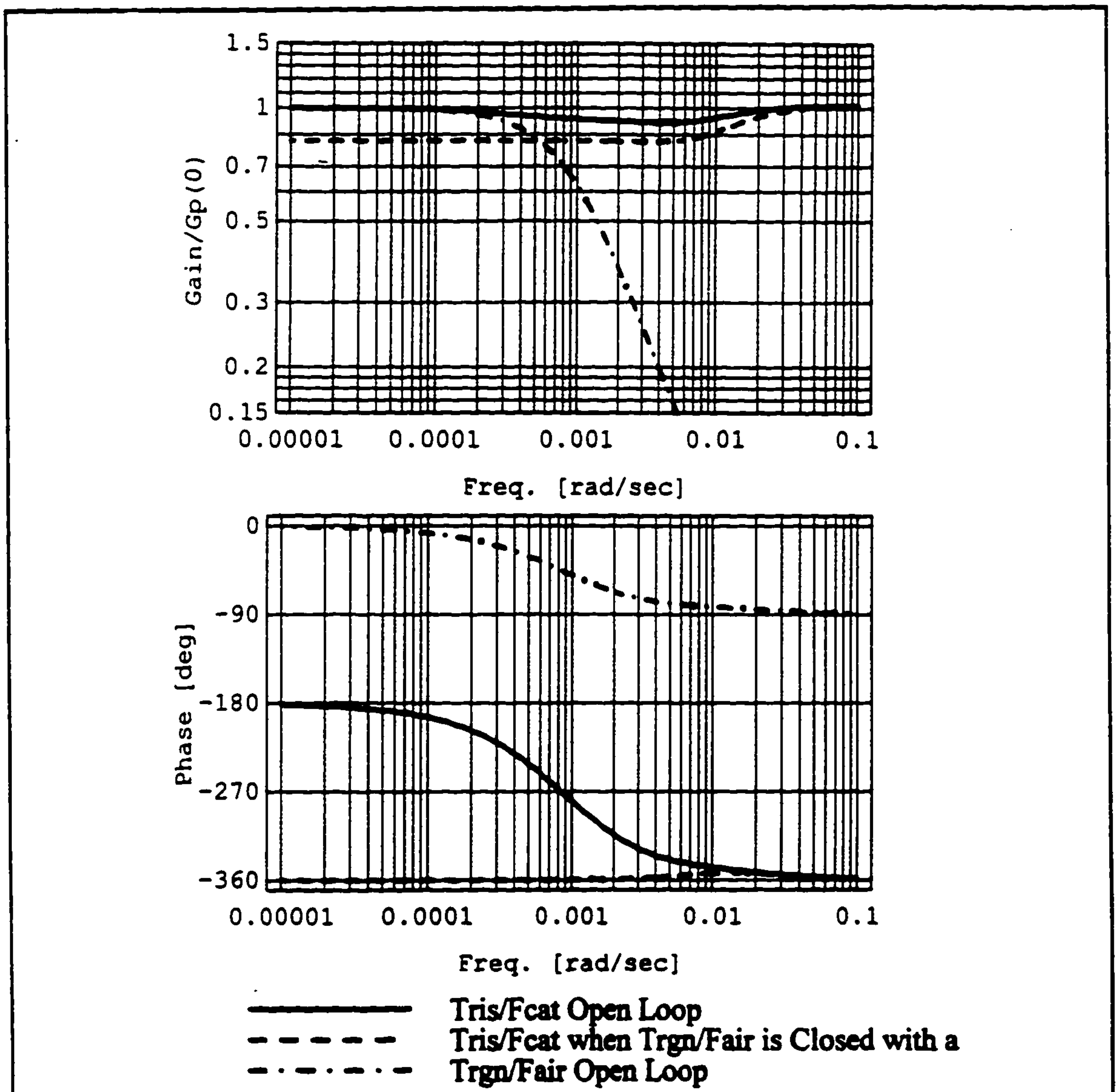


Figure 42: Open loop frequency response of: (1) T_{ris}/F_{cat}; (2) T_{ris}/F_{cat} when T_{rgn} is controlled by F_{cat} with a PI; (3) T_{rgn}/F_{air}

This shows another limitation of the loop with negative gains (reactor loop closed first). Changes in set point or in catalyst or feed properties will change the open loop gain. K_{max} for this loop is the inverse of the normalized gain and therefore a controller based on closing [T_{ris}, F_{cat}] first could become unstable due to changes in the gain. While such a controller could in principle stabilize an unstable steady state, such a stabilization is totally non-robust due to sensitivity to the process gain. On the other hand the conventional controller tuning where [T_{rgn}, F_{air}] is closed first is very robust as long as the set points remain permissible. The stability limits are wide and the main constraints on control action are oscillatory response and saturation. Properly tuned, these controllers can also stabilize an unstable steady state, but it is preferable to avoid this situation.

Low model sensitivity means that less model information is required for control of this loop. Minimum model information required is a very important consideration in the design of control loops for partial control of complex systems. Regrettably there is no literature for designing control algorithms along the negative diagonal. These are highly dependent on model properties and specific goals which can not easily be generalized.

5.1.3.4. Time Response

It is important that the loop finally chosen should have a reasonable time response. It is hard to define a general criterion. The response has to be fast enough to stabilize the system against anticipated perturbations. Perturbations in catalyst activity are slow and do not require fast response. Changes in feed properties can be fast but in proper operation one tries to protect the unit from such disturbances by using a large feed tank as a filter. No control can adequately operate in an adiabatic reactor if disturbances are not filtered. In some sense the slow time response of the unit due to thermal inertia provided by the catalyst holdup in the regenerator offers here a protection. The cost of this protection is that the regenerator responds slowly.

The reactor has very small holdup and in this scheme it allows very fast response. When the sign is reversed the reactor control loop has a right hand zero and is therefore limited due to the inverse response at low frequencies. In the positive diagonal control response of the reactor loop is slow as it is not decoupled from the regenerator holdup.

So let us concentrate on the preferred design [T_{ris}, T_{rgn}, F_{cat}, F_{air}] with positive gains. Note from Fig. 41 and 42 that the main limits on controller gain are due to saturation of the manipulated variable and oscillating response. The limits on the gains are given in Table 10. Fig. 42 shows that the [T_{rgn}, F_{air}] loop is approximately a first order filter with a time scale of about thirty minutes which limits fast control whereas the reactor loop is approximately a simple gain allowing fast and accurate control. The fact that the regenerator is slow is desirable as it allows the operator to recover from mistakes and deal with changes in operating conditions. Thus the control structure when properly tuned should have adequate time response. A typical set of set point changes in T_{ris} and T_{rgn} is given in Fig. 43 and 44 with two different sets of controller constants. Note that the response of the reactor is fast if properly tuned. It is also shown that reducing the gains of the reactor control gives a smoother and a faster control of T_{rgn}. This is a very interesting phenomena that, while theoretically known, has received little attention. If large changes in set points are required, it is preferable to loosen up the loop that normally is tightly controlled, to make it easier to move the unit. An interesting parallel problem exists, for example, in economics. Is it possible to stimulate the economy while maintaining tight control on inflation?

Fig. 43 and 44 also give the output of the manipulated variable as a percentage of its value at steady state. Note that neither variable becomes saturated for a set point change of 5°F. The time response of the regenerator while slow is adequate. Faster control would lead to saturation in most cases and as said before even then it can only be done if the reactor control is reduced otherwise the response of T_{rgn} is oscillatory. More complex algorithms based on decoupling could improve the operation but are outside the scope here. For choosing the structure it is sufficient to show that it can provide adequate control and time response.

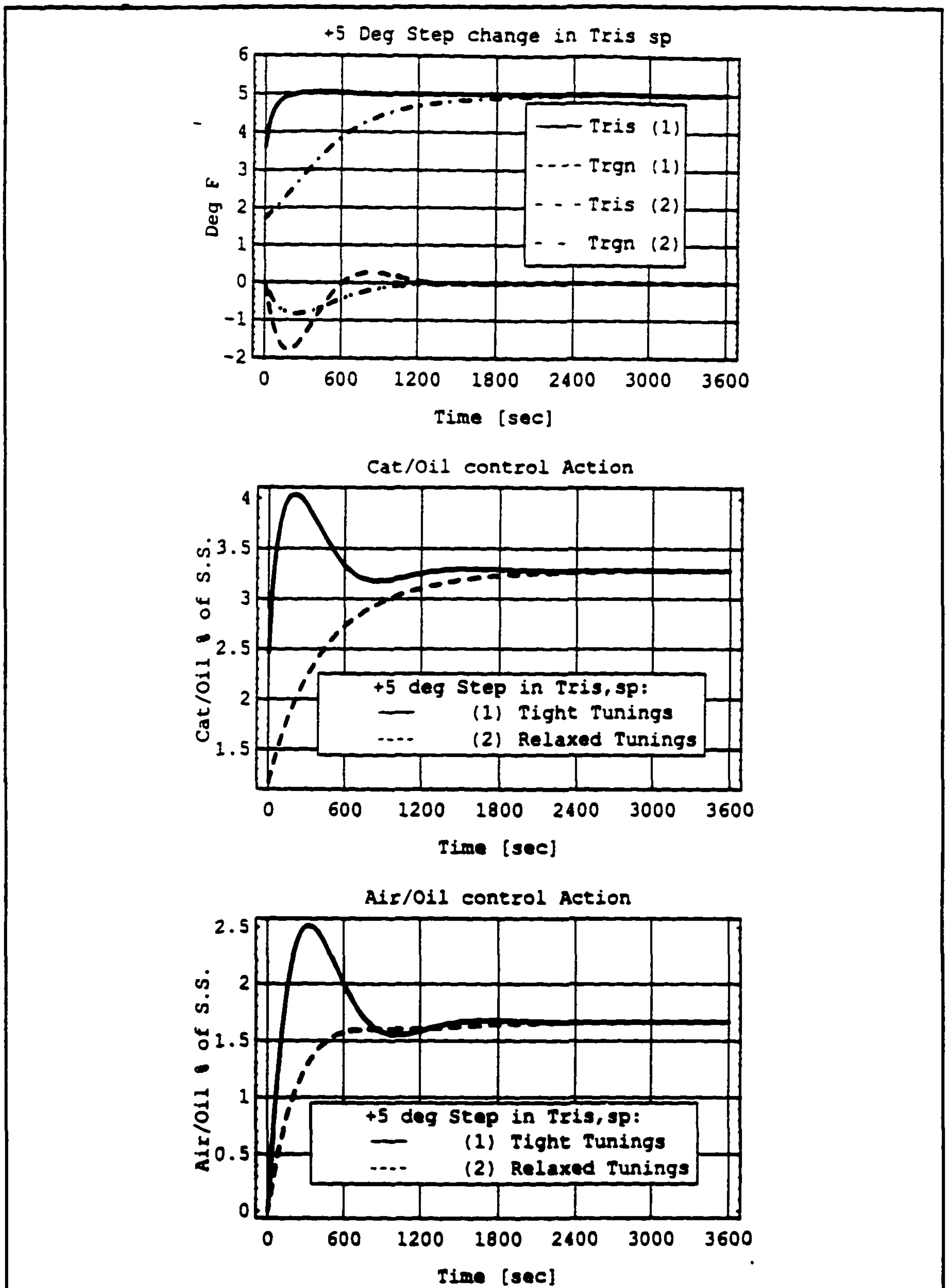


Figure 43: Time response to a step in the set point of T_{ris} -- Effect of control on T_{ris}: (1) - Tight control; (2) Relaxed control; (at Case I of Table 5)

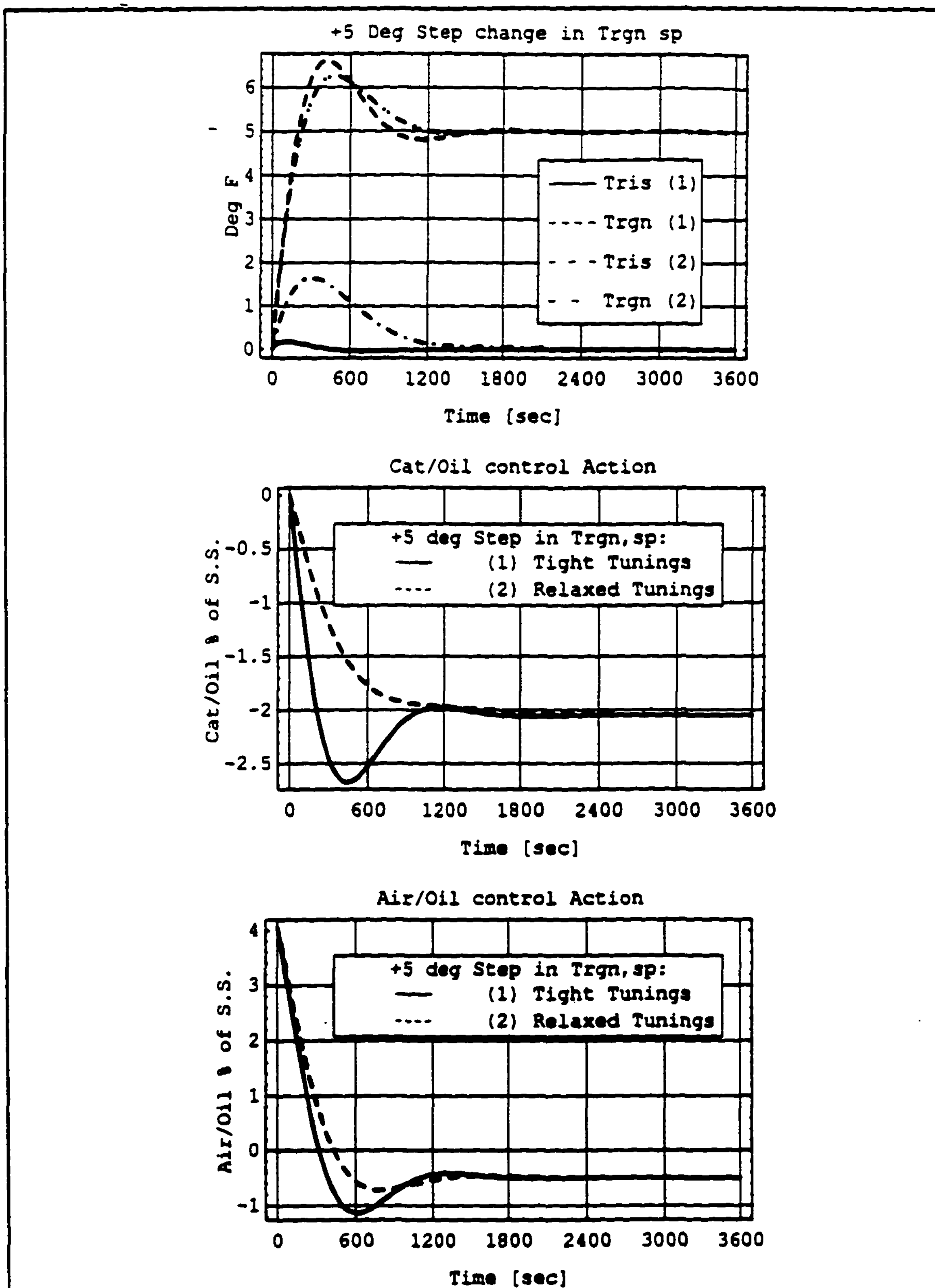


Figure 44: Time response to a step in the set point of T_{rgn} -- Effect of control on T_{ris}: (1) - Tight control; (2) Relaxed control; (at Case I of Table 5)

5.2.4. Evaluation of Control Structures for Complete Combustion

For the operation in complete CO combustion it is a common industrial practice to use combustion promoters. Even though operation is possible without them, promoters reduce CO emissions and ΔT, and thus make the operation smoother. Fig. 45 shows the effect of promoter on T_{rgn} and $O_{2,sg}$ at constant T_{ris}. Also shown in Fig. 45 are the effects of higher activity and higher coking rate. The addition of promoter affects the values of T_{rgn} only in partial combustion. In complete combustion T_{rgn} is not effect by the promoter and still has a flat maximum with F_{air}. As discussed before, it is therefore not controllable with F_{air} in this region. Using F_{cat} instead runs a large risk of losing the full combustion because it will increase the demand for oxygen. The only possibility left using F_{air} and F_{cat} is the structure [$O_{2,sg}$, T_{ris}, F_{air},F_{cat}] and the evaluation will concentrate on it. For cases where this combination proves insufficient, addition of one or more loops will be considered.

The conventional pairing for this structure is [($O_{2,sg}$ - F_{air}),(T_{ris} - F_{cat})]. This is also the pairing suggested by the RGA (see Table 11). The effect of the conventional paring is to keep the unit in complete combustion by guaranteeing a minimum of excess oxygen. This is usually maintained below 2% for economic efficiency. It is possible to operate at 0.5% excess oxygen. This requires more attention as the operation is closer to the stability limit, as can be seen from Fig. 45. What helps is that in complete combustion T_{ris} increases almost linearly with increasing F_{cat} without any input multiplicities because T_{rgn} is almost constant at this region.

Excess O_2 has to be controlled. Otherwise an increase in coking rate might move the system out of full combustion. Aside from keeping the system stable in full

Table 11: Transfer functions, open loop gains and rga for the control structure $[T_{ris}, O_{2,sg}, F_{air}, F_{cat}]$ in complete combustion.

Transfer Functions:

	Stable Point in Complete Combustion, Case IV of Table 5: T_{rgn}=1388°F, T_{ris}=1000°F, T_{feed}=670°F, Xpt=5
$\frac{O_{2,sg}}{F_{air}}$ $\left[\frac{mol\%}{lb\ air/lb\ feed}\right]$	$\frac{4.4(0.00029+s)(0.00078+s)(0.0054+s)(0.0067+s)(0.099+s)}{(0.00029+s)(0.00078+s)(0.0053+s)(0.0068+s)(0.014+s)}$
$\frac{O_{2,sg}}{F_{cat}}$ $\left[\frac{mol\%}{lb\ cat/lb\ feed}\right]$	$\frac{-0.052(0.00029+s)(0.00083+s)(0.0054+s)(0.0069+s)(0.98+s)}{(0.00029+s)(0.00078+s)(0.0053+s)(0.0068+s)(0.014+s)}$
$\frac{T_{ris}}{F_{air}}$ $\left[\frac{mol\%}{lb\ air/lb\ feed}\right]$	$\frac{0.68(0.0054+s)(0.0056+s)(0.00029 \pm i\ 0.000023+s)}{(0.00029+s)(0.00078+s)(0.0053+s)(0.0068+s)(0.014+s)}$
$\frac{T_{ris}}{F_{cat}}$ $\left[\frac{mol\%}{lb\ cat/lb\ feed}\right]$	$\frac{46(0.0003+s)(0.0026+s)(0.0052+s)(0.0054+s)(0.011+s)}{(0.00029+s)(0.00078+s)(0.0053+s)(0.0068+s)(0.014+s)}$

Open Loop Gains and RGA:

Case IV	Steady State Gains		RGA	
	Air/Oil	Cat/Oil	Air/Oil	Cat/Oil
$O_{2,sg}$	33 [mol%/(lb air/lb feed)]	4.5 [mol%/(lb cat/lb feed)]	0.979	0.021
T_{ris}	1.1 [°F/(lb air/lb feed)]	106 [°F/(lb cat/lb feed)]	0.021	0.979

combustion, it has no impact on the other variables. This is illustrated in Table 12 where the effect of changing the set points of $O_{2,sg}$ and T_{ris} for Case IV of Table 5 is shown.

It does, however, impact on NOx and CO emissions (traces) which are not predicted by the model. Changing the set point of T_{ris} has a strong impact on conversion and wet gas and also strongly affects product properties such as octane and gas composition.

Table 12 also shows the effect of increasing the coking rate both in open loop and when controlling T_{ris} and $O_{2,sg}$. When not under control the unit loses complete combustion as discussed above. The main problem when under control is the increase in T_{rgn}. Encountered here is a problem discussed already in Section 5.1.1. In complete combustion the regenerator is much easier to model. If the holdup is sufficient and the system is promoted, the regenerator model is simply a dynamic heat balance. But while modellability is gained, the capability for partial control is reduced as it is no longer possible to control T_{rgn} by manipulating F_{air}. While T_{rgn} is no longer dominant for the regenerator, it is still a very dominant variable with regard to the riser performance. At a given T_{ris} it controls F_{cat} and though it T_{mix}.

Table 12: Control action needed and changes in outputs for the control structure [T_{ris}, $O_{2,sg}$, F_{air}, F_{cat}] in complete combustion

			Set Point Change		+30% in Coking Rate			
	Base		+10 oF Tris	+0.1% O2,sg	Open Loop	Tris O2,sg	Tris O2,sg -30% Activity	Tris O2,sg -70 oF Tfeed
Tfeed	670	oF	0	0	0	0	0	-70
Air/Oil	0.5141	lb air/lb feed	0.0128	0.003	0	-0.0116	-0.0123	0.0503
Cat/Oil	4.121	lb cat/lb feed	0.095	0	0	-0.425	0.026	0.041
Tris	1000	oF	10	0	31	0	-0.1	0
Tmix	1044	oF	11	0	31	-4	-7	-4
Trgn	1389	oF	11	0	51	35	-12	35
O2,sg	1.00	mol%	0.00	0.10	-1.00	0	0	0
Conv	75.8	wt%	1.5	0.0	-0.8	-6.1	-6.6	-3.5
Gasol	48.4	wt%	-0.5	0.0	-1.2	-0.9	-1.0	-0.2
Wg	24.1	wt%	2.0	0.0	0.2	-5.2	-5.5	-3.6
Crgc	0.0193	wt% cat	-0.0014	-0.0007	0.0183	-0.0054	0.8707	0.97
Csc	0.9152	wt% cat	0.0004	0	0.0779	0.0752	-0.8942	-0.8998
coke/feed	3.69	lb c/lb oil	0.09	0.00	0.25	-0.08	-7.30	-7.75

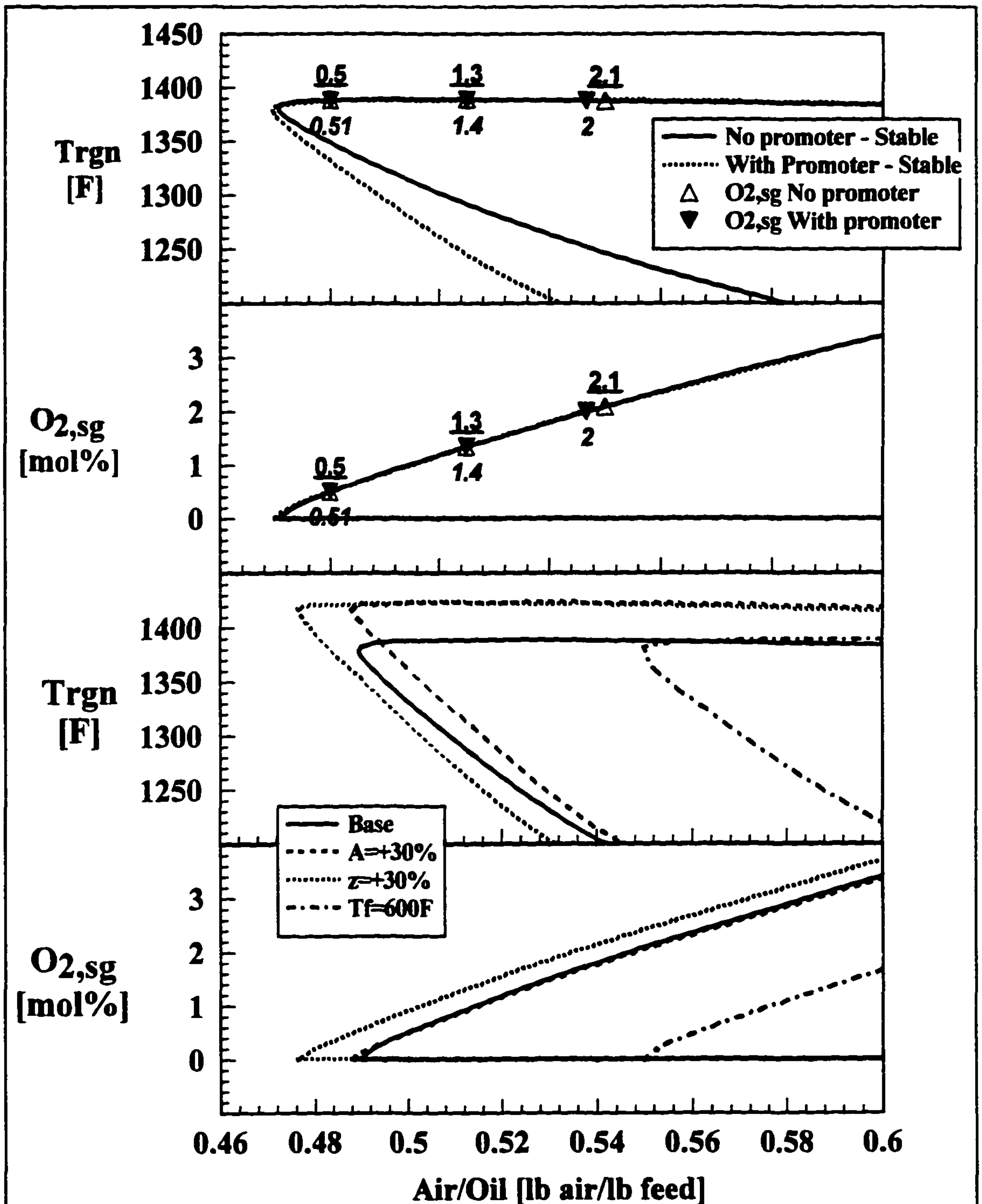


Figure 45: Multiplicities and steady state gains for T_{rgn} and $O_{2,sg}$ at constant T_{ris} = 1000°F -- Impact of: (1) CO combustion promoter; (2) A; (3) z; (4) T_{feed}

There are several ways to reduce T_{rgn}. One is to reduce T_{ris} which is often undesirable because of its impact on the product outputs. However, if T_{rgn} increases too much, it may be necessary to reduce T_{ris} temporarily as this is the only fast variable that can reduce T_{rgn}. Another way to reduce T_{rgn} is to use an additional manipulated variable. These can be feed temperature or a slow manipulated variable such as catalyst activity or feed composition as discussed in Sections 2 and 5. These slow variables should be part of any control strategy in complete as well as in partial combustion. Table 12 shows the impact of using catalyst activity and feed temperature to restore conversion or reduce T_{rgn} similar to Table 8. Reducing feed temperature increases F_{cat} and therefore conversion and wet gas make. It also significantly increases the air rate. It does not decrease T_{rgn} as also noted in Fig. 45. Decreasing catalyst activity reduces T_{rgn} with a smaller impact on conversion, which is lower than in the base case. The main variables for control of T_{rgn} are therefore the slow ones such as feed composition and catalyst activity.

While operating in complete combustion has some advantages such as cheaper investment (no CO boiler), it reduces the overall control capabilities and the ability to process heavier feed stocks. One option to overcome this is a catalyst cooler. A catalyst cooler allows one to control T_{rgn} independently which removes all the control disadvantages of operating in complete combustion.

Not only is modellability easier for complete combustion, but so are dynamic stabilization and control. The stability problem in complete combustion is different than in partial combustion. Due to the presence of combustion promoters, all practical complete combustion controller settings with 0.5-2% excess oxygen represent stable steady states. Low activity or low coking rate could drop T_{rgn} below 1320°F where complete combustion is hard to achieve without promotion. With sufficient promoter it is possible to go even lower, say to 1270°F, while staying in complete combustion. So the stability problem is one of preventing the loss of complete combustion because of insufficient air or too low T_{rgn}. Fig. 45 shows that lower coking rate and lower T_{feed} requires more air, and since most units operate close to maximum air capacity caution is here advised. A constraint on air rate can always be removed by reducing feed rate.

The need to keep the unit in full CO combustion has a interesting implication with respect to stability. In partial combustion failure of the loop [T_{rgn}, F_{air}] means that the unit becomes unstable due to the negative RGA. Here the RGA is positive but this is of no help. If the excess air is low and the air loop fails, the unit can, after a perturbation in coking rate, either crash or go to partial combustion. Thus operating close to the air constraint is not advisable.

The transfer functions for the structure [$O_{2,sg}$, T_{ris}, F_{air},F_{cat}] are given in Table 11 together with the open loop gains and constraints for proportional controllers. Fig. 46 shows the open loop frequency response of T_{ris}/F_{cat} and $O_{2,sg}/F_{air}$. As can be seen all the transfer functions are proper, similar to that of T_{ris}/F_{cat} in partial combustion, but without the RHP zero that adds a phase problem. From Tables 11 and 12 as well as from Fig. 47 which shows the stability limits for this structure it is clear that the stability is limited only by oscillatory response and controllers saturation. Further, the dynamic RGA shown in Fig. 48 shows that the two loops are decoupled at all frequencies unlike the case of partial combustion. All these findings shows that there is problem with tight and fast control of both T_{ris} and $O_{2,sg}$. However, it should be remembered that the response of T_{rgn}, for example to increase in coking rate, will still be slow because of the thermal inertia of the regenerator.

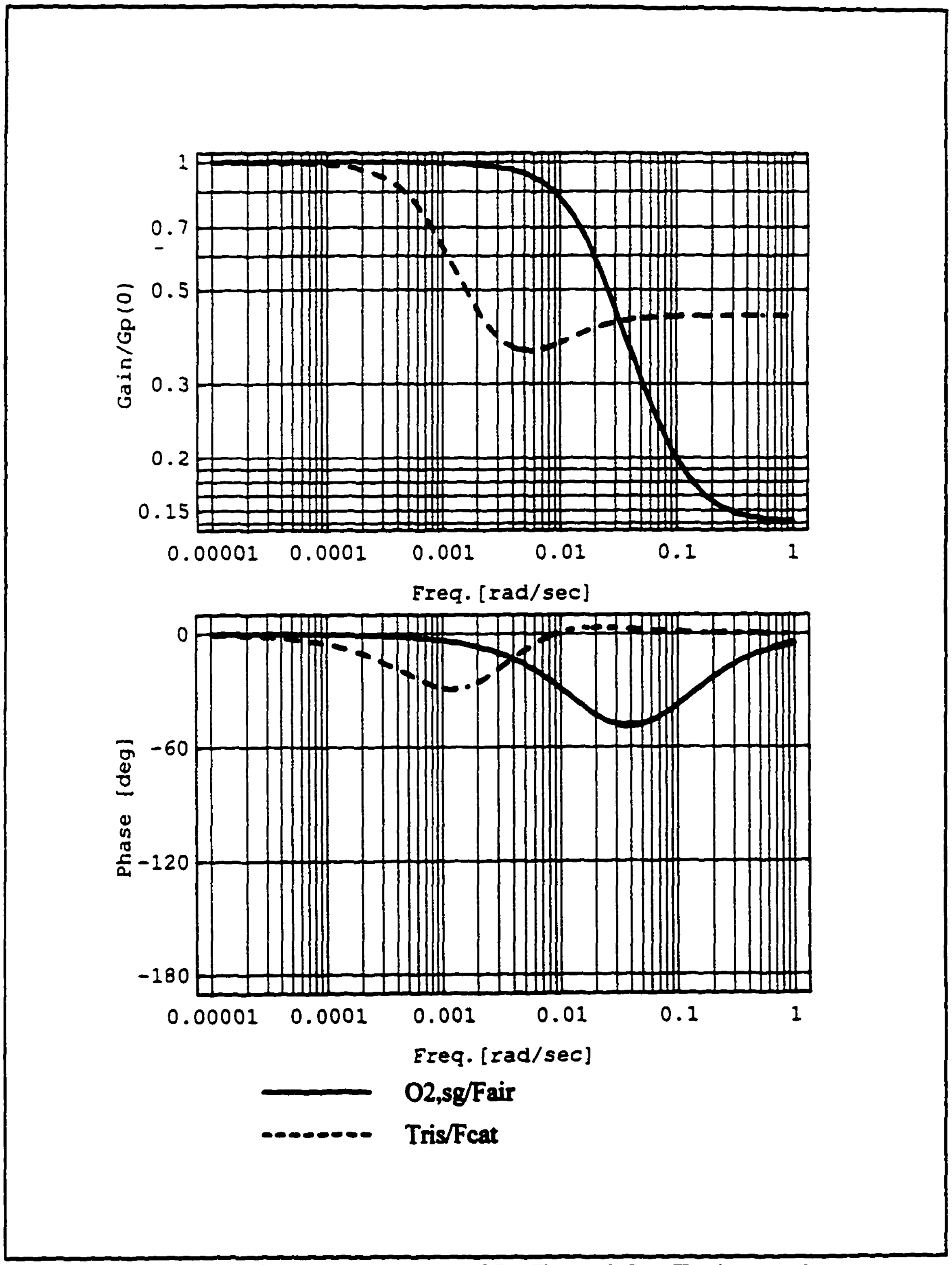


Figure 46: Open loop frequency response of T_{ris}/F_{cat} and $O_{2,sg}/F_{air}$ in complete combustion (Case IV of Table 5)

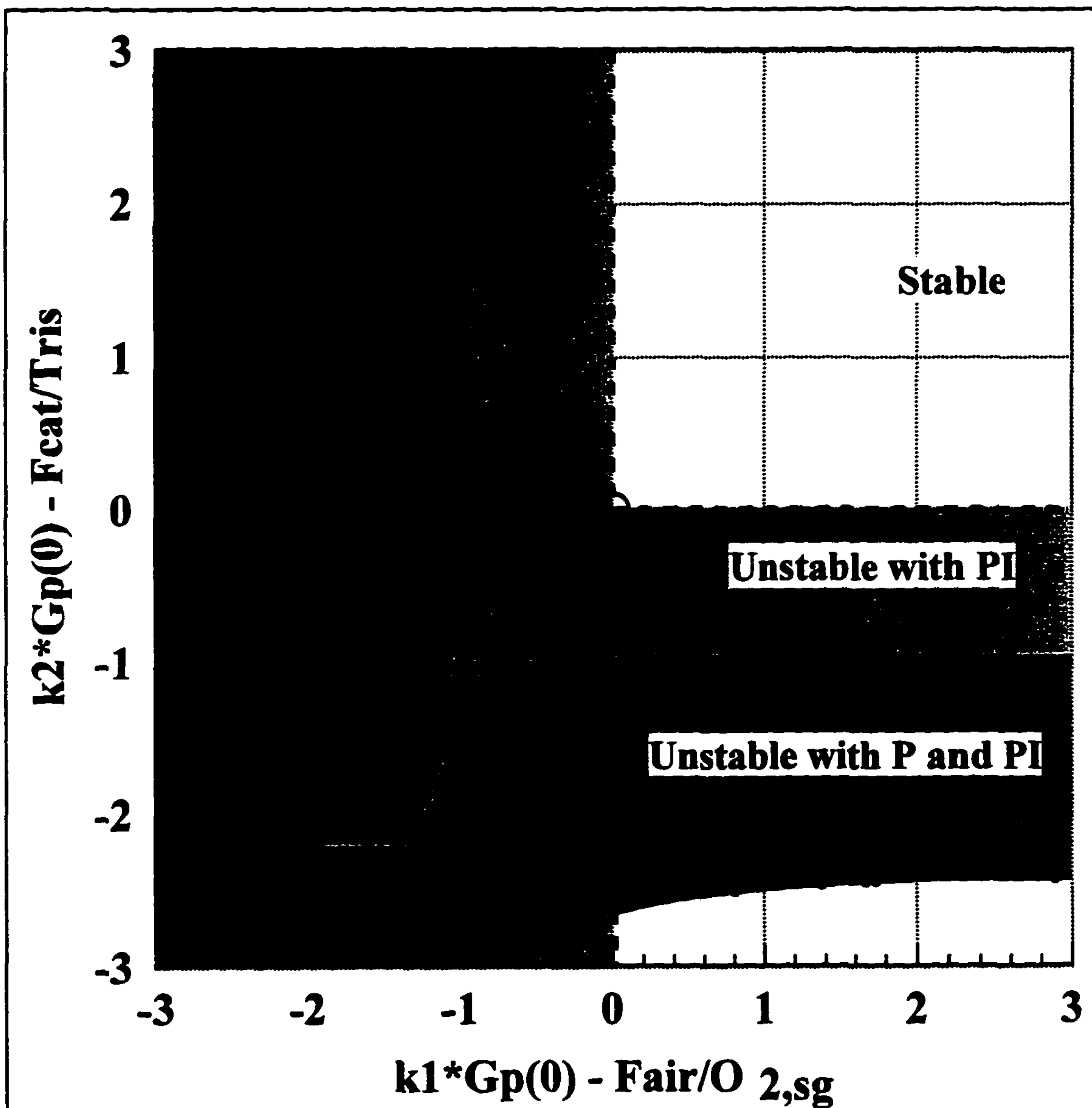


Figure 47: Stability limits for the control structure $[(T_{ris}\text{-}F_{cat}),(O_{2,sg}\text{-}F_{air})]$ at Case IV of Table 5 with P and PI controllers. (PI=Ki(1+0.001/s))

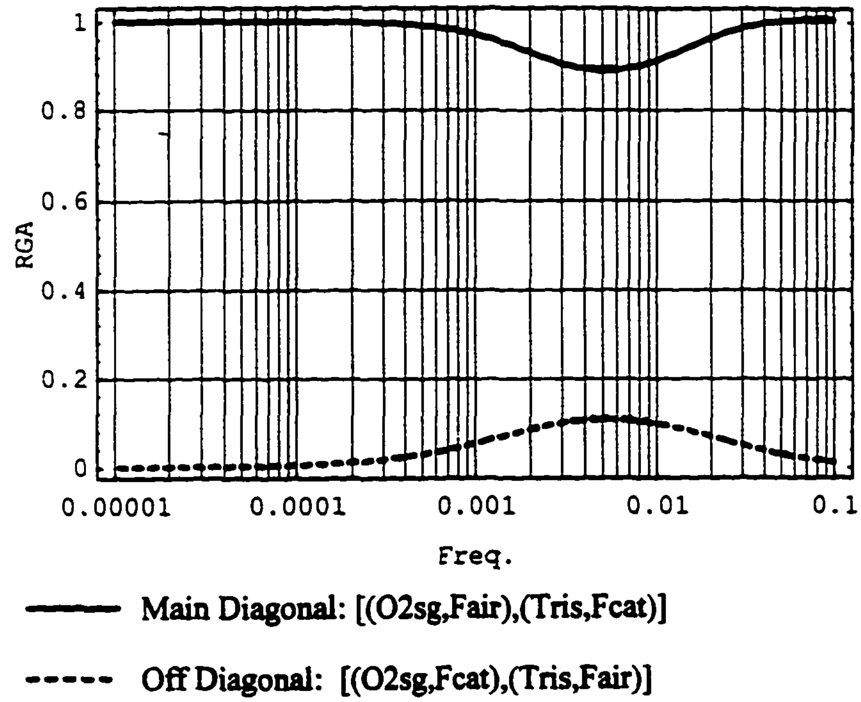


Figure 48: Dynamic RGA for the control structure [$O_{2,sg}$, T_{ris}, F_{air}, F_{cat}] (at Case IV of Table 5)

5.3 Sufficiency

One of the conditions in the list of criteria is that the chosen control structure should be sufficient to meet the goals of the process. It is an important criterion in partial control that has received only little attention. What is meant by it is the ability of the overall control strategy to meet the following criteria. The process has to be able to keep Y_p within the specified goals and to deal with the perturbations expected to occur in an economical acceptable way. Y_p is imposed from the outside. But in determining it one has to take into account the capabilities of the process. The ability of any control structure to achieve these goals is determined in the plant design as it depends not only on the design of the process but also on the number and magnitude of the manipulated variables provided in the original design. It is therefore important to understand what sufficiency means and how the ability to provide a sufficient control structure depends on the design. During the early years (1938) of FCC operation any product coming out of it was acceptable and there are still plants for which this is so. Today the specifications on gasoline have become much tighter due to environmental considerations. This and the need to crack heavier feed stocks give an economical advantage to units with a greater capability for control.

Those units that are limited in their ability to manipulate catalyst circulation rate have an economic penalty as one cannot control T_{ris} and T_{rgn} independently. Much of the development in FCC technology such as feed preheat, catalyst cooler, and the slide valve for controlling F_{cat} has been motivated by the need to provide better controllability. The section on control of complete combustion is a good illustration. While in complete combustion a 2x2 control matrixes used similar to partial combustion, the ability to control Y_p and to deal with heavier feeds is much more limited, a disadvantage that has to be weighted against the cost advantages of the unit. While for some needs this control structure is sufficient for achieving the demand goals, changing needs may justify the conversion of the unit to partial combustion, just as some other units as the Model IV, are being converted to allow control of F_{cat} with a slide valve. There are other industries in which specifications imposed from the outside are much tighter than in refining. There the need to insure sufficiency of the control structure to meet Y_p is a much more urgent concern but the example given here still illustrates the concept.

6. Summary and Conclusions

The purpose of this work is to define and outline the problem of partial control. Partial control is the control problem of plants or processes in which there are many more outputs to control than manipulated variables to control them with. In addition to this most of these processes are also nonlinear which makes the control problem even more complex. Since in partial control it is impossible to keep all the outputs at their set points, it is at least desired to keep the process within a given limits of constraints and specifications. This can be done, and actually had been the practice in industry, by using only much smaller set of decentralized control loops utilizing some or all of the available manipulated variables. This set is called primary control structure

The first objective of the primary control structure is to keep the process at steady state in face of perturbations in the uncontrolled inputs. The second objective is to move the system from one operating point to another following set points changes. These set points are set by a higher level of the overall control strategy, such as optimizing supervisory control.

The questions in the design of the primary control structure are how to choose the outputs to be used with the manipulated variables, how to couple them and, how many loops are enough to achieve the control goals. In order to answer these questions a set of criteria is suggested (Table 1) and an outline for the investigation of the control problem and different solutions is given. As an example to this process the control of a Fluidized Bed Catalytic Cracker (FCC) is investigated using the proposed procedure.

The first step in any investigation of a process is to obtain an accurate as possible model. The main problem here is what is a good model and how simple can it be yet capture the essentials of the process. It is shown that for the FCC a model based mainly on heat and mass balances of the two reactors compromising the unit is adequate. The model suggested behaves very much like real units. It is shown that the basic results based on the model are not sensitive to substantial changes in the model parameters.

Using the model it is now possible to identify the different inputs and outputs for the FCC. The nature and number of manipulated variables are determined in the design stage of the process. Although possible, it is quite costly to add manipulated variables when the process is already up and running. On the other hand, it is relatively easy to add output measurements. As this question of design is a heavy problem to it self, it is not discussed here. Rather, the practice in industry and the literature discussions are used as guidelines.

Like in many other processes, it is possible to identify three groups of inputs. The first is that of the known but unmeasurable disturbances. In the FCC these are feed composition and coking rate. The second group is that of the slow varying inputs. These can be controlled but change only slowly relative to the characteristic time of the system. In the FCC this group include the catalyst properties such as activity and other additives. The last group holds the manipulated variables which can be changed fast relative to the time scale of the process. These are used for the dynamic control of the system. In the FCC they are mainly four namely, air flow rate to the regenerator, catalyst circulation rate between the regenerator and riser, feed preheat temperature and, catalyst circulation rate to the catalyst cooler, if one is available. This work concentrates on the first two.

The list of outputs to be controlled in the FCC is quite long. It include the product description (conversion, gasoline and wet-gas yields, octane, etc.) and process variables related to constraints (regenerator temperature, riser top temperature, stack gas temperature, temperature rise from the regenerator bed to the stack gas, etc.). Again, this work concentrates on those variables which are in industrial use or were discussed in the literature. Therefore, the relevant outputs are regenerator temperature (T_{rgn}), riser top temperature (T_{ris}), stack gas temperature (T_{sg}), the difference T_{sg}-T_{rgn} (ΔT) and, Oxygen in the stack gas ($O_{2,sg}$).

The next step is the investigation of the nonlinear features of the process. The open loop analysis of the FCC (with no control on it) shows that for any given set of fixed inputs there are either one or three steady state. This was checked for a variety of conditions within acceptable operating range. For other cases, when some parameters are changed drastically, there could be five steady states. However, these parameters do not reflect realistic conditions and are based on hypothetical assumptions that do not hold in reality. There is no reported evidence for the existence of more than three open loop steady states in current FCC operation.

In the case there is only one steady state it is a trivial cold steady state where no reactions take place. This is also the case with the first of any three steady states. Although such a steady state is of no operational interest it is important to be able to predict for which operating conditions only such state exists and under which circumstances a change of conditions will lead the unit to such a state. Of the two other states of any three, the middle one was found to be always linearly unstable, as is the case with any system with three steady states. The upper steady state was found to be always linearly stable. No evidence for limit cycles in any of the conditions investigated was found. From the results of this analysis it is clear that it is desired to know under which input conditions there are hot steady states, which of these steady states are inside the operating range of interest and, what is the chance that such state is linearly unstable or that due to a change in operating conditions a stable state changes to unstable.

Further study of the steady states discovered that in addition to output multiplicities there are input multiplicities. That is, not only that for a given set of inputs there are more than one set of outputs, but that a given set of outputs can results from more than one set of inputs. It is shown that for all cases there are at least two input multiplicities while for some there are four. For most of the range where multiplicities occur they do not impose an operational problem because one of the multiplicities is at input settings that are outside the unit capabilities. However, there are some conditions at which multiplicities are a problem. This is true mainly for the outputs T_{sg} and ΔT for which there are regions of four multiplicities.

The possible control structures composed of the inputs and the outputs coupling are next evaluated under the light of the five major criteria of Table 1. Not in any order of importance, these criteria are modellability, dominance, sufficiency, nonlinearities and time scale of response. Modellability is the sensitivity of a variable to its modeling or, how well can it be modeled based on laboratory information and how good is its prediction. Dominance is the ability of a given structure to effect all of the not-controlled outputs with the minimum control action. Sufficiency is the ability of any chosen control structure to achieve the control objectives. If not sufficient, the control structure has to be modified by adding more loops or choosing another structure all together. The nonlinear considerations include stability limits for both open and closed loop, outputs and inputs multiplicities and gain changes. The time scale of the response is an important consideration even if not an overruling one. It is not a unique property of the control structure and here the properties of the process itself must be taken into account.

The control structures studied in this work are: $[T_{ris}, T_{rgn}]$, $[T_{ris}, T_{sg}]$, $[T_{ris}, \Delta T]$, $[T_{rgn}, \Delta T]$ and $[T_{ris}, O_{2,sg}]$ where air flow rate (F_{air}) and catalyst circulation rate (F_{cat}) are used as the manipulated variables in all of the structures. The FCC has two main operating regimes. One is partial combustion where the coke is combusted off the catalyst into a mixture of CO and CO_2. The second is complete combustion where the coke is combusted fully to CO_2. It is shown that only the last of the five structures, in which the Oxygen outlet is controlled, is good for complete combustion. From the other four, the best for partial combustion is shown to be $[T_{ris}, T_{rgn}]$.

$[T_{rgn}, \Delta T]$: This structure, known as the Kurihara scheme, was suggested in order to stabilize the system in face of perturbations in coking rate. It does this, as is shown, but at the price of a strong change in T_{ris}. The fact that T_{ris} is not controlled is too costly in the effect on the product outputs. T_{ris} is a dominant variable and must be controlled in order to keep the product outputs within acceptable limits of specs. Another problem with this scheme are the input multiplicities. These have a complex behavior in the range of interest for operation. Furthermore, ΔT is a variable that is very hard to model. As shown, it is very sensitive to model uncertainties such as bypass in the regenerator's bed.

$[T_{ris}, T_{sg}]$: Here T_{ris} is controlled and therefore the change in the product outputs due to perturbations are smaller compared to the Kurihara scheme. However, the main problems with this structure are its complex input multiplicities and the problem of modelling T_{sg} and predicting its gain. As shown, there is no way of accurately predicting the behavior of T_{sg}. Its gain changes sign for different set points within the operating range. The point at which the sign changes can not be predicted because of the sensitivity to model uncertainties such as flow patterns and other parameters.

[T_{ris}, ΔT]: The problem here are similar to [T_{ris}, T_{sg}]. Although there are no input multiplicities in the practical operating range of partial combustion (these appear at much lower air rates) there is a strong problem of modellability.

[T_{rgn}, T_{ris}]: This scheme came out as the preferred for the control of the FCC in partial combustion mode. Both T_{ris} and T_{rgn} are relatively easy to model based on laboratory results. In the range of partial combustion operation there are only two input multiplicities, one of which is always at air rates that are outside the reach of the unit, and therefore of no interest. Controlling these two outputs keep the unit away from its constraints and well within acceptable changes in the product outputs after a disturbance is introduced.

In the steady state analysis of the control structure there is no need to discuss the pairing of the outputs to the inputs. For the structure [T_{ris}, T_{rgn}, F_{air}, F_{cat}] an interesting situation arises. Practice prefers the use of F_{air} to control T_{rgn} and of F_{cat} for T_{ris}. Analysis shows that this corresponds to pairing based on negative RGA diagonal. Such control scheme is known to be sensitive to failure of one of loops. If this loop is the first to be closed, its failure will cause the other loop to become unstable due to change in the sign of the gain. The unit may then drift to another steady state, in the case of FCC either to complete combustion or to crash down to the cold steady state. This might actually explain reports in the literature of similar behavior. The intuitive explanation for the choice of the negative RGA paring is that T_{rgn} is controlled with the direct input that effects it, the air flow. Then the catalyst flow is used to control T_{ris}.

What makes this case even more interesting is that in the case of FCC, for negative RGA pairing , it makes a difference which loop is closed first. It is shown that it is better to close the [T_{rgn}, F_{air}] loop first. This is because it practically decouples the riser from the regenerator by keeping the temperature of the catalyst entering the riser constant. This, in turn, makes the riser much more similar to a laboratory unit with fixed inputs (for fixed T_{rgn} and F_{cat} the variation in C_{rgc} in small). It is now much easier to model and predict the riser behavior. This order of closing, in oppose to closing [T_{ris}, F_{cat}] first, also results with much more relaxed stability limits for the controllers tuning.

It is shown that all of the four structures discussed for partial combustion are sufficient for the stabilization of the unit in face of perturbations only while staying in this operating regime. However, all of the four are inadequate for control in the complete combustion mode. The most severe problem is that none of them have a realistic gain for control of the other variable with F_{air} when T_{ris} is controlled with F_{cat}. Further, it is shown that in order to keep the unit within complete combustion, the excess Oxygen must be controlled.

[T_{ris}, $O_{2,sg}$]: Only this structure is sufficient for the stabilization of the unit in complete combustion. The O_2 must be controlled in order to ensure enough excess air for complete combustion, and T_{ris} must be controlled in order to keep the product within limits. However, it is shown that it is insufficient to control the temperature of the catalyst regeneration, T_{rgn}, which can as a result of some perturbations rise considerably and damage the catalyst. The only alternative here is to add additional control loops. One option is the use of the feed preheat, but this puts a strong demand on the air flow. Another is the use of catalyst cooler if one is available. However, if there is no fast control circuit to add, a slow varying input such a the catalyst activity can be used.

To conclude, a methodical approach to the investigation of nonlinear and complex systems control options under partial control is outlined and demonstrated on the FCC. It is shown that linear considerations, while good for tuning, are not enough for the complete analysis of such systems. A list of criteria is suggested for this investigation.

Appendix - Model Description

Riser:

The riser is modeled as a plug flow reactor using the ten lump kinetic model of Gross et al. (1976). The following assumptions are made: quasi-steady state, no slip, adiabatic operation, no diffusion limits, same temperature of the two phases at each point, constant heat capacities and constant densities. Eq. A-1, A-2 and A-3 are solved using a CST cascade approximation, with initial conditions as indicated.

Coke balance , based on Voorhies (1945), Gross et al. (1976), Krambeck (1991) and Sapre and Leib (1991):

$$\frac{dC}{dh} = \tau_r A z k_{cc0} e^{-\frac{E_{cc}}{RT}} \left(\frac{\Psi}{100}\right)^{\frac{1}{b}} b C^{1-\frac{1}{b}} \quad ; \quad C(h=0) = \alpha C_{rgc} \qquad \text{(A-1)}$$

Oil cracking balance, based on Jacob et al. (1976) and Gross et al. (1976):

$$\frac{dy_i}{dh} = \tau_r A \rho_c \frac{(1-\epsilon)}{\epsilon} \phi \frac{1}{1+k_h y_{Rh}} \sum_{k=1}^{9} a_{ik} y_k K_{k0} e^{-\frac{E_k}{RT}} \quad ; \quad y_i(h=0) = y_i(feed) \qquad \text{(A-2)}$$

Heat balance, based on Jacob et al. (1976):

$$\frac{dT}{dh} = -\frac{F_{tf}}{F_{rgc}Cp_c + F_{tf}Cp_{fv}} \sum_{i=1}^{9} \left|\frac{dy_i}{dh}\right| \Delta H_{ri} \quad ; \quad T(h=0) = \frac{F_{rgc}Cp_c T_{rgn} + F_{tf}Cp_{fl}T_{feed} - \Delta H_{evp}F_{fl}}{F_{rgc}Cp_c + F_{tf}Cp_{fv}} \qquad \text{(A-3)}$$

The lumping scheme and kinetic constants are given in Table A-I.

Table A-1: 10 Lump model kinetic constants

From:		To:		Activation Energy [Btu/lbmole]	Rate @1000°F [1/sec]	Heat of Reaction [Btu/lb]
HFO	Ph	LFO	Pl	26,100	0.196	25
	Nh		Nl		0.196	
	Ah		Al		0.196	
	Ah		Rl		0.489	
	Rh		Rl		0.049	
HFO	Ph	Gasoline	G	9,900	0.611	65
	Nh		G		0.939	
	Ah		G	26,100	0.685	
HFO	Ph	Coke	C	31,500	0.099	225
	Nh		C		0.149	
	Ah		C		0.198	
	Rh		C		0.149	
LFO	Pl	Gasoline	G	9,900	0.282	40
	Nl		G		0.752	
	Al		G	26,100	0.196	
LFO	Pl	Coke	C	31,500	0.099	200
	Nl		C		0.099	
	Al		C		0.050	
	Rl		C		0.010	
Gasoline	G	Coke	C	18,000	0.048	160

HFO - Heavy Fuel Oil; LFO - Light Fuel Oil; Coke - Coke + C_1 - C_4;
A - Aromatic Substitution Groups; R - Carbons Among Aromatic Rings;
P - Paraffins; N - Napthenes; G - Gasoline; C - Coke;
h - Heavy; l - Light;

The catalyst deactivation for cracking goes together with its coking [Krambeck (1991)], then:

$$\phi = C^{1-\frac{1}{b}} \tag{A-4}$$

The feed coking tendency is given by Gross et al. (1976):

$$\begin{aligned}\Psi &= 0.631 y_{Ph}(0) + 0.297 y_{Nh}(0) + 0.773 y_{Ah}(0) + 2.225 y_{Rh}(0) + \\ &+ 0.631 y_{Pl}(0) + 0.11 y_{Nl}(0) + 1.475 y_{Al}(0) + 0.0727 y_{Rl}(0)\end{aligned} \tag{A-5}$$

The bed characteristics are given by:

$$\epsilon = \frac{F_{tf}/\rho_v}{F_{tf}/\rho_v + F_{rgc}/\rho_c} \;\; ; \;\; u = \frac{F_{tf}}{A_{ris}\epsilon\rho_v} = \frac{F_{rgc}}{A_{ris}(1-\epsilon)\rho_c} \;\; ; \;\; \tau_r = \frac{H_{ris}}{u} \tag{A-6}$$

The riser conversion is given by:

$$Conv = 1 - \sum_{i=1}^{8} y_i \tag{A-7}$$

The wet gas yield is given by:

$$Y_{wg} = Conv - Y_g - (C|_{riser\ top} - \alpha C_{rgc})\frac{F_{rgc}}{F_{tf}} \tag{A-8}$$

Stripper:

The stripper is modeled as a well-mixed tank, with a linear stripping function and a constant temperature drop. Note that by setting the catalyst flows equal ($F_{rgc}=F_{sc}$) Eq. A-10 becomes redundant.

Coke balance:

$$\frac{dC_{sc}}{dt} = \frac{F_{rgc}}{W_{st}}[C|_{riser\ top} + (1-\alpha)C_{rgc} + \frac{\beta\, CCR}{F_{rgc}/F_{tf}} + \gamma - C_{sc}] \tag{A-9}$$

Catalyst inventory balance:

$$\frac{dW_{st}}{dt} = F_{rgc} - F_{sc} \tag{A-10}$$

Heat balance:

$$\frac{dT_{st}}{dt} = \frac{F_{rgc}}{W_{st}}[T_{ris} - T_{st} - \Delta T_{st}] \tag{A-11}$$

Stripping function:

$$\gamma = 0.0002 + 0.0018(1 - ks) \tag{A-12}$$

Regenerator:

The regenerator has two regions, a dense bed and a dilute phase. The gas is assumed to flow through the regenerator in plug flow. The catalyst in the dense bed is well mixed, while the entrained catalyst in the dilute phase is in plug flow. The response time of the regenerator is dominated by the large holdup of catalyst, therefore, the response time of T_{rgn} and C_{rgc} is much longer compared to the gas residence time or changes in bed density. Quasi-steady state is assumed, therefore, at any instant of time for gas composition and bed properties. This allows to describe the plug flow as a CST cascade. In each phase three equal CSTRs are used. Further, the pressure is assumed constant by fast control of the flue gas flow rate.

Kinetics:

Four reaction are taking place in the regenerator:

$$I: \quad C+\frac{1}{2}O_2 \xrightarrow{k_1} CO \quad ; \quad II: \quad C + O_2 \xrightarrow{k_2} CO_2$$

$$III_c: \quad CO+\frac{1}{2}O_2 \xrightarrow{k_{3c}} CO_2 \quad ; \quad III_h: \quad CO+\frac{1}{2}O_2 \xrightarrow{k_{3h}} CO_2 \qquad \text{(A-13)}$$

$$IV: \quad H_2+\frac{1}{2}O_2 \xrightarrow{k_4} H_2O$$

The hydrogen combustion, reaction IV, is assumed to be complete and immediate. The hydrogen weight fraction in the coke is assumed constant. The coke combustion, reactions I and II, are proportional to C_{rgc} and P_{O2}. The CO combustion, reactions III, are proportional to P_{O2} and P_{CO} and take place in two parallel paths, heterogeneous and homogeneous. The resulting rate expressions, in [lbmole/ft^3 sec], are:

$$r_1 = (1-\epsilon)\rho_c k_1 \frac{C_{rgc}}{MW_c} P_{O_2} \qquad \text{(A-14)}$$

$$r_2 = (1-\epsilon)\rho_c k_2 \frac{C_{rgc}}{MW_c} P_{O_2} \qquad \text{(A-15)}$$

$$r_3 = k_3 P_{O_2} P_{CO} \quad ; \quad k_3 = x_{pt}(1-\epsilon)\rho_c k_{3c} + \epsilon k_{3h} \qquad \text{(A-16)}$$

Where x_{pt} is a relative combustion rate simulating the addition of promoter. The initial ratio of CO/CO_2 at the catalyst surface is given by Weisz (1966):

$$\left.\frac{CO}{CO_2}\right|_{surface} = \frac{k_1}{k_2} = \beta_C = \beta_{C0} e^{\frac{E_\beta}{RT}} \qquad \text{(A-17)}$$

Then, defining k_c to be the overall coke combustion rate:

$$k_C = k_1 + k_2 \qquad k_1 = \frac{\beta_C k_c}{\beta_C + 1} \qquad k_2 = \frac{k_c}{\beta_C + 1} \qquad \text{(A-18)}$$

Gas phase mass balance:

A molar balance for each gas component in each CSTR will lead to the following:

$$
\begin{aligned}
f_{O_2} &= f_{O_2,in} - V\left[(1-\epsilon)\,\rho_c\left(\frac{1}{2}k_1 + k_2\right)\frac{C_{rgc}}{MW_c}\frac{f_{O_2}}{f_{tot}}P_{rgn} + \frac{1}{2}k_3\frac{f_{O_2}}{f_{tot}}\frac{f_{CO}}{f_{tot}}P_{rgn}^2\right] \\
f_{CO} &= f_{CO,in} + V\left[(1-\epsilon)\,\rho_c k_1\frac{C_{rgc}}{MW_c}\frac{f_{O_2}}{f_{tot}}P_{rgn} - k_3\frac{f_{O_2}}{f_{tot}}\frac{f_{CO}}{f_{tot}}P_{rgn}^2\right] \\
f_{CO_2} &= f_{CO_2,in} + V\left[(1-\epsilon)\,\rho_c k_2\frac{C_{rgc}}{MW_c}\frac{f_{O_2}}{f_{tot}}P_{rgn} + k_3\frac{f_{O_2}}{f_{tot}}\frac{f_{CO}}{f_{tot}}P_{rgn}^2\right] \\
f_{N_2} &= f_{N_2,in} \\
f_{H_2O} &= f_{H_2O,in} \\
& f_{O_2}(0) = 0.21F_{air} - \frac{1}{2}f_{H_2O}(0)\ ;\ f_{CO}(0) = 0\ ;\ f_{CO_2}(0) = 0\ ; \\
& f_{N_2}(0) = 0.79F_{air}\ ;\ f_{H_2O}(0) = \frac{1}{2}F_{cat}(C_{sc} - C_{rgc})C_H\frac{1}{MW_H}
\end{aligned}
\qquad \text{(A-19)}
$$

Where f_i is in [lbmole/sec], $f_{i,in}$ is the inlet to each CSTR, $f_i(0)$ is the inlet to the first CSTR and f_{tot} is the sum of all f_i's.

Carbon mass balance:

Under the assumption that all the entrained catalyst returns to the dense bed, it is possible to write an overall carbon balance for the regenerator:

$$
\frac{dC_{rgc}}{dt} = \frac{1}{W_{rgn}}\left[(F_{sc}C_{sc} - F_{rgc}C_{rgc})(1 - C_H) - (f_{CO} + f_{CO_2})_{cyc}MW_c\right] \qquad \text{(A-20)}
$$

For the dilute phase, in each CSTR at quasi-steady state:

$$
(F_{ent}C_{rgc})_{in} - (F_{ent}C_{rgc})_{out} + (f_{CO} + f_{CO_2})_{in}MW_c - (f_{CO} + f_{CO_2})_{out}MW_c = 0 \qquad \text{(A-21)}
$$

Catalyst inventory balance:

$$
\frac{dW_{rgn}}{dt} = F_{sc} - F_{rgc} \qquad \text{(A-22)}
$$

Note again, that by setting the catalyst flows equal (F_{rgc}=F_{sc}) Eq. A-22 becomes redundant.

Heat balance:

Similar to the carbon balance, an overall heat balance for the regenerator:

$$\frac{dT_{rgn}}{dt} = \frac{1}{W_{rgn} Cp_c}[(Q_C + Q_H + Q_{air} + Q_{sc}) - (Q_{rgc} + Q_{sg} + Q_{wall})] \qquad \text{(A-23)}$$

and for each CSTR in the dilute phase:

$$Q_{ent,in} - Q_{ent,out} + Q_{gas,in} - Q_{gas,out} + Q_{C,CSTR} = 0 \qquad \text{(A-24)}$$

Where:

$$Q_{air} = F_{air} \int_{T_{base}}^{T_{air}} [0.79\, Cp_{N_2} + 0.21\, Cp_{O_2}]\, dT \qquad \text{(A-25)}$$

$$Q_{sg} = \sum_{i=1}^{5} \left[f_i \int_{T_{base}}^{T_{sg,cyc}} Cp_i\, dT \right] \qquad \text{(A-26)}$$

$$Q_{sc} = F_{sc} \int_{T_{base}}^{T_{stripper}} Cp_c\, dT \approx F_{sc} \overline{Cp_c} (T_{stripper} - T_{base}) \qquad \text{(A-27)}$$

$$Q_{rgc} = F_{rgc} \int_{T_{base}}^{T_{rgn}} Cp_c\, dT \approx F_{rgc} \overline{Cp_c} (T_{rgn} - T_{base}) \qquad \text{(A-28)}$$

$$\begin{aligned} Q_C + Q_H = {} & 0.21 F_{air} \int_{T_{rgn}}^{T_{ref}} Cp_{O_2}\, dT + F_{cat}(C_{sc} - C_{rgc}) Cp_{coke} (T_{ref} - T_{rgn}) + \\ & + f_{H_2O} \Delta H_{f,H_2O} + f_{CO} \Delta H_{f,CO_2} + f_{CO} \Delta H_{f,CO_2} + \\ & + f_{H_2O} \int_{T_{ref}}^{T_{rgn}} Cp_{H_2O}\, dT + f_{O_2} \int_{T_{ref}}^{T_{rgn}} Cp_{O_2}\, dT + f_{CO} \int_{T_{ref}}^{T_{rgn}} Cp_{CO}\, dT + f_{CO_2} \int_{T_{ref}}^{T_{rgn}} Cp_{CO_2}\, dT \end{aligned} \qquad \text{(A-29)}$$

Where T_{ref} is the temperature for which ΔH_f is given and T_{base} the base temperature for the heat balance, taken as T_{rgn}. Since all the hydrogen is combusted in the dense bed, Q_H=0 in the dilute phase. Q_{gas} and Q_{ent} in the dilute phase are given similar to Eq. A-26 and A-28, respectively. The heat losses to the atmosphere are modelled as heat transfer to the wall and from the wall to the surroundings:

$$\frac{d\bar{T}_w}{dt} = \frac{1}{d_w Cp_w \rho_w}(Q_{wall} - Q_{lost}) \tag{A-30}$$

Where:

$$Q_{wall} = A_w\left[\frac{1}{U_{in}} + \frac{1}{2}\frac{d_w}{k_w}\right]^{-1}(T_{rgn} - \bar{T}_w) \qquad Q_{lost} = A_w\left[\frac{1}{U_{out}} + \frac{1}{2}\frac{d_w}{k_w}\right]^{-1}(\bar{T}_w - T_{atm}) \tag{A-31}$$

The bed Characteristic :

The bed characteristics are calculated based on the air molar flow rate into the regenerator. Gas molar density, from ideal gas law [lbmole/ft^3]:

$$\rho_g = P_{rgn} / RT_{rgn} \tag{A-32}$$

Superficial linear velocity [ft/sec]:

$$u = F_{air} / \rho_g A_{rgn} \tag{A-33}$$

Void fraction in the dense phase from King [1989]:

$$\epsilon_{den} = (0.305u + 1)/(0.305u + 2) \tag{A-34}$$

Catalyst density in the dense phase [lb/ft^3]:

$$\rho_{den} = \rho_c(1 - \epsilon_{den}) \tag{A-35}$$

Catalyst density in the dilute phase, from McFarlane et al. (1993) [lb/ft^3]:

$$\rho_{dil} = \mathrm{MAX}\,[0, (0.582\,u - 0.878)] \tag{A-36}$$

Void fraction in the dilute phase:

$$\epsilon_{dil} = \rho_{dil}/\rho_c \tag{A-37}$$

Dense bed height [ft]:

$$z_{bed} = \mathrm{MIN}\left\{z_{cyc}, \frac{W_{rgn} - \rho_{dil}A_{rgn}z_{cyc}}{A_{rgn}(\rho_{den} - \rho_{dil})}\right\} \tag{A-38}$$

Entrained catalyst flow rate [lb/sec]:

$$F_{ent} = \rho_{dil}A_{rgn}u \tag{A-39}$$

Cyclones:

When the solids are removed from the gas in the cyclones, the CO reacts with any residual O_2 in a self ignite reaction. The ignition depends on the temperature of the gas, the temperature of the cyclones and the gas composition. For purpose of modelling, A lower temperature limit of 1120 °F is set for the ignition of this reaction. It should be noted that during start-up, when the cyclones are cold, the reaction may start at a higher temperature, but when the cyclones are hot it may exist at lower temperatures. The heat balance over the cyclones is given by:

$$Q_{gas,in} - Q_{gas,out} = 0 \qquad \text{(A-40)}$$

Where Q_{gas} is given by Eq. A-26, and the flow rates are calculated based on the limiting reactant.

References

Arandes, J. M.; de Lasa, H. I. Simulation and Multiplicity of Steady States in Fluidized FCCUs, *Chem. Eng. Sci.*, **1992**, *47*, 2535-2540.

Arbel, A.; Huang, Z.; Rinard, I. H.; Shinnar, R.; Sapre, A. V. Dynamic and Control of Fluidized Catalytic Crackers. 1. Modeling of the Current Generation of FCC's. *Ind. Eng. Chem. Res.*, **1995a**, *34*, (4), 1228-1243.

Arbel, A.; Rinard, I. H.; Shinnar, R.; Sapre, A. V. Dynamic and Control of Fluidized Catalytic Crackers. 2. Multiple Steady States and Instabilities. *Ind. Eng. Chem. Res.*, **1995b**, *34*,(4), 3014-3026.

Aris, R. Introduction to the Analysis of Chemical Reactors, *Prentice-Hall*, Englewood Cliffs, New Jersey, **1965**, pp. 173-190.

Aris, R. Mathematical Modelling. Pitman, **1978**.

Arthur, J. R. Reactions Between Carbon and Oxygen, *Trans. Faraday Soc.*, **1951**, *47*, 164.

Avidan, A. A.; Shinnar, R. Development of Catalytic Cracking Technology. A Lesson in Chemical Reactor Design, *I&EC Res.*, **1990**, *29*, (9-11), 931-942.

Avidan, A. A.; Edwards, M.; Owen, H. Innovative Improvements Highlight FCC's Past and Future, *Oil and Gas J.*, **1990**, *88*, (1), 33-58.

Balakotaiah, V.; Luss, D. Global Analysis of the Multiplicity Features of Multi-Reaction Lumped-Parameter Systems, *Chem. Eng. Sci.*, **1984**, *39*, (5), 865-881.

Balakotaiah, V.; Luss, D. Input-Multiplicity in Lumped-Parameter Systems, *Chem. Eng. Commun.*, **1985**, *39*, 309-322.

Balchen, J. G.; Ljungquist, D.; Strand, S. State-Space Predictive Control, *Chem. Eng. Sci.*, **1992**, *47*, 787-807.

Bozicevic, J.; Lukect, D. Dynamic Mathematical Model of the Fluid Catalytic Cracking Process, *Trans. Inst. M. C.*, **1987**, *9*, (1), 8-12.

Bristol, E. H.; On a New Measure of Interactions for Multivariable Process Control. *IEEE Trans. Automat. Control*, **1966**, AC-11, 133.

Chester, A. W.; Schwartz, A. B.; Stover, W. A.; McWilliams, J. P. Catalyzing the energy balance of Cat Cracking, *CHEMTECH*, **1981**, *1*, 50-58.

Cutler, C. R.; Ramaker, B. L. Dynamic Matrix Control - A Computer Control Algorithm. In *Proc. Automatic Control Conf.*, San Francisco, CA, **1980**, Paper WP5-B.

de Lasa, H. I.; Errazu, A.; Barreiro, E.; Solioz, S. Analysis of Fluidized Bed Catalytic Cracking Regenerator Models in an Industrial Scale Unit, *Can. J. Chem. Eng.*, **1981**, *59*, 549-553.

Denn, M. Process Modelling. W. H. Freeman, New York, **1986**.

Edwards, W. M.; Kim, H. N. Multiple Steady States in FCCU Operations. *Chem. Eng. Sci.*, **1988**, *43*, 8, 1825-1830.

Elnashaie, S. S. E. H.; El-Hennawi, I. M. Multiplicity of the Steady State in Fluidized Bed Reactors-IV. Fluid Catalytic Cracking (FCC), *Chem. Eng. Sci.*, **1979**, *34*, 1113-1121.

Elnashaie, S. S. E. H.; Elshishini, S. S. Digital Simulation of Industrial Fluid Catalytic Cracking Units - IV. Dynamic Behaviour, *Chem. Eng. Sci.*,**1993**, *48*, (3), 567-583.

Elshishini, S. S.; Elnashaie, S. S. E. H. Digital Simulation of Industrial Fluid Catalytic Cracking Units: Bifurcation and Its Implications, *Chem. Eng. Sci.*, **1990**, *45*, 553-559.

Elshishini, S. S.; Elnashaie, S. S. E. H. Digital Simulation of Industrial Fluid Catalytic Cracking Units: Bifurcation and Its Implications, *Chem. Eng. Sci.*, **1990**, *45*, (2), 553-559.

Errazu, A. F.; de-Lasa, H. I.; Sarti, F. A Fluidized Bed Catalytic Cracking Regenerator Model. Grid Effects, *The Canadian J. of Chem. Eng.*,**1979**, *57*, 191-197.

Ewell, R. B.; Gadmer, G. Design Cat Crackers by Computer, *Hydrocarbon Processing*, **1978**, *4*, 125-134.

Farag, H.; Ng, S.; Lasa, H.; Kinetic Modeling of Catalytic Cracking of Gas Oils Using in Situ Traps (FCCT) To Prevent Metal Contaminant Effects, *Ind. Eng. Chem. Res.*, **1993**, *32*, 1071-1080.

Felipe, L. I.; Richard, R. M. Dynamic Modelling of a Fluid Catalytic Cracking Unit, *Presented at the AIChE Ann. Meet.*, Los Angeles, CA, **1991**.

Ford, W. D.; Reineman, R. C.; Vasalos, I. A.; Fahrig, R. J. Modeling Catalytic Cracking Regenerators, Presented at the *NPRA Ann. Meet.*, San Antonio, TX, **1976**, pap. AM7629.

Giona, M.; Palidino, O. Bifurcation Analysis and Stability of a Controlled CSTR, *Comp. & Chem. Eng.*, **1994**, *18*, (9), 877-887.

Grosdidier, P.; Morari, M.; Holt, B. R. Closed-Loop Properties from Steady-State Gain Information, *Ind. Eng. Chem. Fundam.*, **1985**, *24*, 221-235.

Gross, B.; Jacob, S. M.; Nace, D. M.; Voltz, S. E. Simulation of Catalytic Cracking Process, US Patent 3960707, **1976**.

Guigon, P.; Large, J. F. Application of the Kunii-Levenspiel Model to a Multistage Baffled Catalytic Cracking Regenerator, *Chem. Eng. J.*, **1984**, *28*, 131-138.

Hovd, M.; Skogestad, S. Procedure for Regulatory Control Structure Selection with Application to the FCC Process. *AIChE J.*, **1993**, *39*, (12), 1938-1953.

Hovd, M.; Skogestad, S. Pairing Criteria for Decentralized Control of Unstable Plants. *Ind. Eng. Chem. Res.* , **1994**, *33*, 2134-2139.

Hovd, M.; Skogestad, S. Controllability Analysis for the Fluid Catalytic Cracking Process. *Ind. Eng. Chem. Res.*, **1992**, *31*, 1589.

Huq, I.; Morari, M.; Sorensen, R. C. Design of operational Modification to Model IV FCCUs to Improve Dynamic Performance. *AIChE J.*, **1995**,

Iscol, L., The Dynamics and Stability of a Fluid Catalytic Cracker, Presented at the *ASME 11th Jt. Automat. Contl. Conf. Amer. Automat. Contl. Council, Atlanta, GA*, **1970**, Pap. 23-B.

Jacob, S. M.; Gross, B.; Voltz, S. E.; Weekman, V. M. Jr. A Lumping and Reaction Scheme for Catalytic Cracking, *AIChE J.*, **1976**, *22*, (4), 701-713.

Kestenbaum, A.; Shinnar, R.; Thau, F. F. Design Concepts for Process Control. *I&EC Proc. Des. Dev.*, **1976**, *15*, 2.

King, D. F. Estimation of Dense Bed Voidage in Fast and Slow Fluidized Beds of FCC Catalyst, *Fluidization VI, Procs of the Int. Conf. on Fluidization*, R. Grace et al. Eds., Alberta, Canada, **1989**, 1-8.

Kokotovic, P. V.; O'Malley Jr., R. E.; Sannuti, P. Singular Pertubations and Order Reduction in Control Theory -- An Overview. *Automatica*, **1976**, *12*, 123-132.

Koppel, L. B. Input Multiplicities in Nonlinear Multivariable Control Systems, *AIChE J.*, **1982**, *28*, (6), 935-945.

Krambeck, F. J. Continuous Mixtures in Fluid Catalytic Cracking and Extensions, *Mobil Workshop on Chemical Reaction in Complex Mixtures*, Van Nostrand Reinhold, New York, **1991**, 42, 42-59.

Krishna, A. S.; Parkin, E. S. Modeling the Regenerator in Commercial Fluid Catalytic Cracking Units, *Chem. Eng. Prog.*, **1985**, *81*, (4), 57-62.

Kunii D.; Levenspiel, O. Fulidization Engineering, *Wiley*, New York, **1969**.

Kurihara, H. Optimal Control of Fluid Catalytic Cracking Process. Ph.D. Dissertation, MIT, **1967**.

Larocca, M.; Ng, S.; Lasa, H. Fast Catalytic Cracking of Heavy Gas Oils: Modeling Coke Deactivation, *Ind Eng. Chem. Res.*, **1990**, *29* 171-180.

Lee, W.; Kugelman, A. M. Number of Steady-State Operating Points and Local Stability of Open-Loop Fluid Catalytic Cracker, *Ind. Eng. Chem. Process Des. Dev.*, **1973**, *12*, (2), 197-204.

Lee, L.; Chen, Y.; Huang, T.; Pan, W. Four-Lump Kinetic Model for Fluid Catalytic Cracking Process, *Can. J. Chem. Eng.*, **1989**, *67*, 615.

Lee, E.; Groves, F. R. Jr. Mathematical Model of the Fluidized Bed Catalytic Cracking Plant, *Trans. Soc. Comput. Simulation*, **1985**, *2*, 219-236.

Lee, L.; Yu, S.; Cheng, C. Fluidized-bed Catalyst Cracking Regenerator Modelling and Analysis, *Chem. Eng. J.*, **1989**, *40*, 71-82.

Liss, B.; Shinnar, R. The Dynamic Behavior of Crystallizers in Which Nucleation and Growth Depend on Properties of the Crystal Magma, *AIChE J.*,**1976**, *72*, (153), 28-40.

Ljungquist, D. Online Estimation in Nonlinear State-Space Models with Application to Catalytic Cracking. Dr. Eng. Thesis, Norwegian Instit. of Technol., **1990**.

Luyben, W. L.; Lamb, D. E. Feedforward control of a Fluidized Catalytic Reactor-Regenerator System, *Chem. Eng. Prog. Symposium. Ser.*, **1963**, *46*, (59), 165-171.

Martinez, E. C.; Drozdowicz, B. Multiple-Scale Approach to Real-Time Simulation of Stiff Dynamic Systems. *Computers Chem. Engng.*, **1989**, *13*, 7, 767-778.

McAvoy, T. J.; Interaction Analysis. *Instrument Society of America*, Research Triabgle Park, USA, **1983**.

McDonald, G. W. G.; Harkins, B. L. Maximizing FCC Profits by Process Optimization, *NPRA Ann. Meet.*, San Antonio, TX, **1987**.

McFarlane, R. C.; Reinman, R. C.; Bartee, J. F.; Georgakis, C. Dynamic Simulation for Model VI Fluid Catalytic Cracking Unit, *Computers Chem. Eng.*, **1993**, *17*, (3), 275.

McGreavy, C.; Isles-Smith, P. C. "Modelling of a Fluid Catalytic Cracker," *Trans. Inst. M. C.*, **1986**, *8*, (3), 130-136.

Meadowcroft, T. A.; Stephanopoulos, G.; Brosilow, C. The Modular Multivariable Controller: I: Steady-State properties. *AIChE j.*, **1992**, *38*, 8, 1254-1278.

Morari, M.; Lee, J.H. Model Predictive Control: The Good, the Bad, and the Ugly. In *Chemical process Control CPC IV, proceedings of the 4rd International Conferance on Chemical Process Control*, **1991**.

Morari, M. Robust Stability of System with Integral Control. *IEEE 22nd CDC*, **1985**, 865-869.

Morari, M.; Arkun, Y; Stephanopolous, G. Studies in the Synthesis of Control Structure for Chemical Processes, Part I, II and III. *AIChE J.*, **1982**, *28*, 60.

Morari, M.; Zafiriou, E. Robust Process Control. Prentivce-Hall, Englewood Cliffs, NJ, **1989**.

Morbidelli, M.; Varma, A.; Aris, R. Reactor Steady-State Multiplicity and Stability *in Chemical Reactor and Reaction Engineering*, Carberry, J.J., A. Varma Eds., Marcel Dekker, New York, **1987**, Chapter 17.

Moro, L. F. L.; Spandri, R.; Odloak, D. Constrained Multivariable Control of Fluid Catalytic Cracking Converters - A Practical Application. Pres. at the AIChE National Meeting, Houston, TX; April **1995**; paper 84b.

Paraskos, J. A.; Shah, Y. T.; McKinney, J. D.; Carr, N. L. A Kinematic Model for Catalytic Cracking in a Transfer Line Reactor, *Ind. Eng. Chem. Process Des. Dev.*, **1976**, *15*,(1), 165-169.

Pellegrini, L.; Biardi, G. Chaotic Behavior of a Controlled CSTR, *Comp. & Chem. Eng.*, **1990**, *14*, (11), 1237-1247.

Prater, C. D.; Wei, J.; Weekman, V. W., Jr.; Gross, B. A Reaction Engineering Case History: Coke Burning in Thermofor Catalytic Cracking Regenerators, *advances in Chemical Engineering*, Wei, J., Ed., Academic Press, New York, **1983**, *12*, 1-

59.

Prett, D. M.; Garcia, C. E. Fundamental Process Control. Butterworth: Stoneham, MA, **1988**.

Quann, R. J.; Jaffe, S. B. Structure - Oriented Lumping - Describing the Chemistry of Complex Hydrocarbon Mixtures, *Ind. Eng. Chem. Res.*, **1992**, *31*, (11), 2483-97.

Razon, L.F.; Schmitz, R. A. Multiplicities and Instabilities in Chemically Reacting Systems - a Review, *Chem. Eng. Sci.*, **1987**, *42*, (5), 1005-1047.

Rosenbrock, H. H. Computer-Aided Control System Design. Academic Press, London, **1974**.

Ross, R. A.; Fong, P. Catalytic Conversion of Wood Barks to Fuel Gases, *Ind. Eng. Chem. Prod. Res. Dev.*, **1981**, *20*, (1), 197-203.

Sapre, A. V.; Leib, T. M. Translation of Laboratory Fluid Cracking Catalyst Characterization Tests to Riser Reactors, *ACS Symposium Series 452*, Fluid Catalytic Cracking II-Concepts in Catalyst Design, M. L. Occelli, Ed., **1991**, Chapter 9, 144-164.

Seko, H.; Tone, S.; Otake, T. Operation and Control of a Fluid Catalytic Cracker, J. Chem. Eng. Jpn, **1978**, *11*, (2), 130-135.

Seko, H.; Tone, S.; Otake, T. Criterion for Stability of Steady States and its Prediction in a Fluid Catalytic Cracker, *J. Chem. Eng. Japan*, **1982**, *15*, (4) ,305-310.

Shah, Y. T.; Huling, G. T.; Paraskos, J. A.; Mckinney, J. D. A Kinematic Model for an Adiabatic Transfer Line Catalytic Cracking Reactor, *Ind. Eng. Chem. Process Des. Dev.*, **1977**, *16*, (1), 89-94.

Shinnar, R. Chemical Reactor Modeling. The Desirable and the Achievable. *Chem. React. Eng. Rev.*, **1978**, *29*, 1-36.

Shinnar, R. Chemical Reactor Modelling for the purposes of Controller Design. *Chem. Eng. Commun.*, **1981**, *9*, 73-99.

Shinnar, R. The Interface Between Industry and the Academic World. *AIChE Symposium Series*, **1976**, *159*, 72, 155-166.

Shinnar, R. Chemical Reactor Modelling. The Desirable and the Achievable, *Chemical Reaction Engineering Reviews*, IEC Research, **1978**, *29*, 1-36.

Shinnar, R. Impact of Model Uncertainties and Nonlinearities on Modern Controller Design. Present Status and Future Goals. In *Chemical Process Control - CPCIII, proceedings of the 3rd International Conferance on Chemical Process Control*, Morari, M., McAvoy, T. J., Eds.; Elsvier: New York, **1986**, 53-93.

Shinseky, F. G.; Process Control Systems. McGraw Hill, New York, **1967**.

Shinseky, F. G.; Distillation Column Control. 2nd ed., McGraw Hill, New York, **1984**.

Shnaider, G. S.; Shnaider, A. G. Kinetic Models of Catalytic Cracking of Oil Fractions in Single-stage and Multistage Reactors with Fluidized Beds and Interpretation of Experimental Data of Catalytic Cracking Performed in these Reactors, *Chem. Eng. J.*, **1990**, *44*, 53-72.

Silverstein, J.; Shinnar, R. The Effect of Design on the Stability and Control of Fixed Bed Catalytic Reactors with Heat Feedback 1. Concepts. *I&EC Proc. Des. Dev.*, **1982**, *21*, 241-256.

Takatsuka, T.; Sato, S.; Morimoto, Y.; Hashimoto, H. A Reaction Model for Fluidized-Bed Catalytic Cracking of Residual Oil, *Int. Chem. Eng.*, **1987**, *27*, 107-116.

Uppal, A.; Ray, W. H. On the Dynamic Behavior of Continuous Stirred Tank Reactors, *Chem. Eng. Sci.*, **1974**, *29*, 967-985.

Upson, L. L.; Hemler, C. L.; Lomas, D. A. Unit Design and Operational Control: Impact on Product Yields and Product Quality, *Fluid Catalytic Cracking: Science and Technology*, J. S. Magee and M.M. mitchell, Jr. Eds, Studies in Surface Science and Catalysis, Vol. 76, Elsevier, New York, **1993**.

Voorhies, A. Carbon Formation in Catalytic Cracking, *Ind. Eng. Chem.*, **1945**, *37*, 318.

Weekman, V. W., Jr. Lumps, Models, and Kinetics in Practice, *AIChE Monograph Series*, **1979**, *75*, 1-29.

Weekman, V. W., Jr.; Nace, D. M. Kinetics of Catalytic Cracking Selectivity in Fixed, Moving, and Fluid Bed Reactors, *AIChE J.*, **1970**, **16**, 397-404.

Weisz, P. B.; Goodwin, R. D. Combustion of Carbonaceous Deposits within Porous Catalyst Particles, II. Intrinsic Burning Rate, *J. Catal.*, **1966**, *6*, 227-236.

Weisz, P. B. Combustion of Carbonaceous Deposits within Porous Catalyst Particles, III. The CO_2/CO Product Ratio. *J. Catal.*, **1966**, *6*, 425-430.

Zhao, X.; Lu, Y. Nonlinear Dynamic Model and Parameter Estimation for Fluidized

Catalytic Cracking Unit, *IFAC Proceedings Series*, **1988**, *2*, 1079-1083.

Zheng, Y. Dynamic Modeling and Simulation of a Catalytic Cracking Unit, *Computers Chem. Engng*, **1994**, *18*, (1), 39-44.