

Partial Control of Complex Chemical Plants I Control of a Catalytic Fluid Cracker

A. Arbel, Z. Huang, I. Rinard and R. Shinnar

Dept. of Chemical Engineering, The City College of New York
138th St. at Convent Ave., New York, NY, 10031

I INTRODUCTION

Till recently the main emphasis in the chemical industry has been on the development of new products and processes. However, in recent years it has become evident that control offers an important tool to improve the profitability and competitiveness of the industry. The same applies to refineries which have a large potential for savings by better control, mainly by integration of process models into on-line control. The resulting overall control is in most cases based on adjusting setpoints of individual control loops, the success of which strongly depends on the choice of the proper control loops.

There are several problems facing the control engineer when dealing with steady state control that are different from those facing him when designing or tuning individual control loops, either single loop or multi-variable.

The first is that most of these systems are nonlinear. While one can use linearization for designing individual control loops, efficient steady state control requires nonlinear models. If steady state control requires large changes in setpoints, then the linearized transfer function will also change. The second problem one has to take into account is that exact models are seldom available. The third is that the system itself can undergo fundamental changes when new catalysts are developed, or other process modifications introduced.

The fourth difference and maybe the most critical problem the designer faces is that the number of variables that need to be controlled is often much larger than the number of manipulated variables that are at his disposal. As the number of manipulated variables is determined in the design phase, understanding how this choice affects our ability to efficiently control a process is important for better design. The quality of control will also be affected strongly by the measured variables chosen for control.

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

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(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 matrix, in particular, the set of variables that comprise this matrix. 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 variables that need to be controlled is larger than the number of manipulated variables, then we cannot control them at exact setpoints. All we can hope for is to keep them within an acceptable space. We can do this 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 variables in a square control matrix 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 setpoints of this matrix to keep the required specification variables in an acceptable space. The second method is much more frequently used as a 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. We call 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. Our 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.

The present paper is part of a research effort to better understand the problems of partial control. We use here as an example the control of a Fluid Catalytic Cracker (FCC). Our goal

is twofold. One is to develop a better understanding of the control problems of an FCC in the context of model based control of a refinery itself an important industrial problem. The second is to better understand the general problem of designing partial control systems.

II. PARTIAL CONTROL - DEFINITION AND CONCEPT.

For a chemical reactor, the problem was outlined in Shinnar (1981 and 1986). We will shortly summarize the pertinent aspects.

Let us first define the control problem. A chemical reactor (or any nonlinear system) is described by a nonlinear equation or model:

$$X = M^* (X, U, W, N) \quad (1)$$

where X is the vector of the state variable, U are the manipulated inputs, and W are inputs that can be measured even if they are not manipulated, and N are inputs that are unknown or not measured.

As the dimension of X is very large, we measure normally only a subspace Y of the vector X , Y is given by:

$$Y = CX \quad (2)$$

where C is the measurement matrix.

For complex reactors M^* is not completely known. We try to identify a simplified model M by a combination of identification and mathematical modelling [Aris (1978), Denn (1986), Shinnar (1978)]. This model will be a model in the subspace:

$$Y = M (Y, U, W, N) \quad (3)$$

For simplicity we sometimes redefine N as a disturbance in Y :

$$Y = M (Y, U, W) + N \quad (4)$$

and write the model in terms of Eq. (4).

Y is here the set of all measured state variables and output variables entering our model, and U are all the manipulated variables used for control. However, not all measured state

variables are used in the control system design in an equal way. Some are integrated into the dynamic control, some are used to adjust the setpoints, and some are used mainly for data logging and diagnostic purposes. We therefore divide Y into the following subsets.

Y_d contains the process variables monitored either continuously or with small sampling intervals compared to the overall time scale of the process. Y_s is the vector containing the infrequently measured process variables. Obviously the vector Y contains the elements of both Y_s and Y_d . We also introduce equivalent definitions of U .

There are, however, two other subsets of Y that are important for deriving a control strategy. These are Y_p , the vector or set of process variables that define the product and process specifications, as well as process constraints. Y_p is normally of a much smaller dimension as Y and its elements may lay in both Y_d and Y_s . In chemical reactors it is seldom completely contained in Y_d and part of it lies in Y_s . Y_{cd} is the set of process variables on which we base our dynamic control strategy. Measurements of a large number of variables may be available for information and can be used in the control. Normally we choose a much smaller set of Y as the basis of the dynamic control strategy. If we want to use any integral control, the control matrix must be square and its dimension is limited by the number of manipulated variables available. Its elements are not necessarily made up of Y_p .

Reactor control in general is a strategy to develop a sufficiently large set of U and use it to maintain Y_p within prescribed limits:

$$Y_{pj \min} < Y_{pj} < Y_{pj \max} \quad (5)$$

We note the goal of the control in Eq. 5 is different from the one used in deriving specific algorithms for square matrices in which the goal is to keep a vector Y at a fixed value. This is the distinction between partial control and exact control. The manipulation of U (both U_d and U_s) is based on the information provided by monitoring Y and W .

To achieve the above goal we choose in most cases a small set of measured variables, Y_{cd} and try to keep these at a fixed set of values (or setpoints) by manipulating U_d . The setpoints of Y_{cd} are determined from information obtained by measuring Y_p . Information from other measured variables may also be used and, if available, should be used.

Having chosen this set of basic variables we can now integrate them into a control algorithm G defined by:

$$U = G (Y, W, U) \quad (6)$$

We normally assume that we can decompose G into two parts, a primary dynamic control system operating with a limited set of variables:

$$U_d = G_d (Y_{cd}, Y_d, W, U_d) \quad (7)$$

and the steady state control which determines the setpoints of Y_{cd} , Y_{cd}^s as well as the values of U_s :

$$(Y_{cd}^s, U_s^s) = G_s^s (Y_s, W, U_s) \quad (8)$$

where the superscripts denotes steady state values.

At this point, we should point out that using square primary control matrices, and manipulating the setpoints for steady state control is not the only way to deal with partial control. If a good overall nonlinear model is available one can in theory 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, Ljungquist and Strand (1991)]. In some cases this has been applied quite satisfactorily. However, this requires a much more detailed model information than is normally available. The conventional approach in industry is the one outlined here, our paper will focus solely on this approach.

Let us now define what we mean by U_s . The manipulated variables that enter the basic square control matrix can be manipulated either continuously or frequently (such as in sampled data control). But in many large systems and chemical processes there are often other important variables, that are adjustable at a much slower rate. Thus in an FCC the activity of the catalyst is controlled by daily catalyst removal or withdrawal. It takes days to make a substantial change. The type of feed entering the FCC can also be changed. As an FCC can lose its stability and wind down in hours, it makes sense to distinguish between fast changing manipulated variables entering a dynamic (fast) matrix and those adjusted much less frequently. We note that such a definition is somewhat arbitrary. For example, the feed temperature can be manipulated purely

for purposes of steady state control or, it can be used to continuously adjust reactor temperature in a dynamic feed back circuit. Feed rate can also be adjusted rapidly, but is seldom manipulated continuously. We introduce this difference here as in the control literature as well as in practice, these slow variables are often neglected, despite their being essential tools in proper partial control.

The fast matrix in partial control is used mainly to keep the system stable and take care of disturbances. The fast matrix is also used to keep dominant variables at a desired value, and to prevent the system from exceeding critical process constraints. Thus, for example, when the FCCs are operated at low regenerator temperatures (below 1230°F), as used to be the practice not so long ago, it was critical to control tightly the temperature of the flue gas to prevent erosion of the cyclones by after burning. But flue gas temperature is only a process constraint and has no impact on the process itself. State variables that strongly impact on the total process are called dominant variables.

II.a Dominant Variables and Sufficiency

There is no a priori reason that a specific square matrix of partial control should work in a given system. There is no law that says that with a given limited set of manipulated variables we can keep a system with a large number of state variables stable and that, furthermore, we can keep all desired specifications in an acceptable space. For quite a number of complex systems this is often achievable in a satisfactory way provided we choose the proper set of variables for manipulation and measurement. To put this into a more rigorous framework we have to understand two principles: dominance and sufficiency.

Consider, for example, an FCC. There are three variables that strongly affect all reactions occurring in the reactor. One is the type and activity of the catalyst in the reactor. The other two are the temperature of the regenerator and the temperature profile of the reactor. The temperature in the regenerator is fairly uniform and the dense bed temperature characterizes it quite well. Regenerator temperature impacts the rate of all processes occurring in the regenerator. It also impacts what happens in the reactor as at fixed temperature, it determines catalyst circulation rate. The reactor temperature varies from top to bottom but at fixed feed rate and feed temperature, the profile is an almost unique function of riser top temperature and

regenerator dense bed temperature, so these two temperatures will have a dominant influence on all other kinetic parameters. Inherent catalyst activity changes very slowly, whereas temperatures can change quite rapidly. There is therefore significant hope that by just keeping regenerator temperature and reactor temperature at some given values we can keep the system dynamically stable and furthermore have a strong impact on the total vector of output variables, which for FCCs is very large. In fact, for purposes of stability one can often get away by just controlling regenerator temperature which is the more critical one. But we have to realize that the number of state variables is larger than these three mentioned. Even the simplest model will have two more: coke on regenerated catalyst and coke on spent catalyst.

In many reactors, temperatures are dominant variables. The reason for this is that temperature enters the reaction rate constants of all reactions. This simultaneous effect on a large number of rate constants is one characteristic of a dominant variable and probably the most critical one. The state of the catalyst is another dominant variable. Again changes in the state of the catalyst due to aging affect in most cases all reaction rate constants in a predictable way. Similarly interest rates in economic systems affect a large number of economic variables and interest rate has therefore a dominant effect.

But not in all cases does temperature have a dominant effect on complex chemical reactions. There are a number of complex reactions in which the apparent activation energy is very small. Furthermore, the temperature range of acceptable operation may be so small such that temperature has to be kept constant and cannot be manipulated to control other variables.

The fact that the basic partial control scheme uses dominant variables, both as measured and manipulated variables, is not enough to guarantee that we can meet our goals or stabilize the system.

The number of manipulated variables used must be sufficiently large. This minimum number strongly depends on:

- 1) The nature of the system of M^*
- 2) The nature and size of disturbances
- 3) The nature of the space Y_p (the permitted range of the variables entering the specifications and constraints).
- 4) The choice of the operating region

While Item 4 is not intuitively obvious we will later show that it applies, for example, to conventional operating schemes in fluidized catalytic cracking.

The absolute minimum dimension of the basic control matrix G_d is given by the requirement that the system has to be stable. Our first step in any design of complex systems is to understand the dynamic behavior of the system sufficiently to be able to deal with this problem. Here, we face the problem that accurate information about M^* is often unavailable at the design stage [Shinnar (1978) and (1986)].

One interesting research problem relates to M^* . What are the minimum information and the minimum model complexity necessary to understand the stability and dynamic response problem sufficiently to make a satisfactory choice of the matrix G_d ? As the manipulated variables are chosen in design we have often to choose G_d before we have a unit to study M^* . Pilot plants are mostly used to confirm scale up and are seldom used to build a dynamic model. Luckily for engineers this is easier to deal with in reactor design than in economics.

In practice we have to realize that M^* itself can change. For example, catalyst properties are often contained in M^* . When we introduce a new catalyst, its effect on M^* may not be known. The same applies when introducing a new feedstock. While a proper model should predict such effects, our imperfect model may not account for them.

Our ability to maintain a system at a stable steady state also depends on the nature and size of the disturbances. Manipulated variables have a limited range of action. Furthermore, a set of manipulated variables which is sufficient for a given operation may become insufficient if the disturbances are too large. Partial control therefore always involves limiting the magnitude of feed disturbances (or changes) as well as the rate at which they can change. We achieve this by specifying the composition of the feedstocks and by filtering disturbances using large feed tanks.

If the system can be unstable or has multiple steady states, we also need another type of minimum model information in order to choose a set of setpoints for partial control. We need some approximate knowledge about the space of acceptable setpoints. Quite often we do not have this information, the operator has to search for a stable range of setpoints.

While sufficiency in terms of stability can be rigorously defined, sufficiency in terms of the acceptable space is a much looser concept. We often tend to define Y_p by what we think a

unit can do. A wrongly defined space of Y_p can lead to instabilities in practice, a situation met quite often by the authors (IR and RS).

Y_p may also change in a plant. One of the most important changes affecting refineries today is due to environmental legislature. The specifications on liquid fuels have dramatically changed, which has a major impact on control, the implications of which have not yet been sufficiently digested by the industry.

One way of improving stability and increasing the permissible space of Y_p is to introduce additional manipulated variables. We will later show that FCC went that way [see Avidan and Shinnar (1990)]. But there is an important caveat. Some variables may have strong interactions and in the presence of a highly imperfect model adding more manipulated variables to the primary control matrix may make control more difficult. In fact most people hesitate to use more than two manipulated variables in a primary dynamic control matrix. More are used only if one can clearly separate the time scale. In that case we treat these as inputs. Thus, for example, inlet temperature and pressure can be manipulated fast, and with very little interaction with the primary control matrix. Again this is an area where more research is needed.

Most research on nonlinear control deals with the speed of response. The critical problems of nonlinear systems are however more in the impact of the nonlinear features of the system on steady state control and in the way model nonlinearities affect and change the properties of a control matrix. This will be discussed later using the example of the FCC.

II.b Process Dominated by Specification Vs. Process Dominated by Constraints.

Before dealing with the example of the FCC there is one more important concept that we want to clarify, as it in the past has often clouded the discussion of control problems across industries. Aside from complexity and the quality of model information, there is one critical aspect that distinguishes control practice in different industries and plants. In some industries control is dominated by the desire to optimize process performance while avoiding or meeting process constraints. In others control is totally dominated by trying to meet difficult product specifications. Let us illustrate this by two extreme examples.

The first case is an ammonia plant. There is no great difficulty in a well designed ammonia plant to meet the specification for liquid ammonia. What dominates the control in a

wider sense, aside from maintaining a stable operation, is looking for a more profitable operation while meeting production goals. In a given plant we can either maximize throughput or minimize cost, or some combination of the two. The obstacles we meet are limits on unit and equipment capabilities (compressor throughput, reformer temperature, catalyst activity, etc.). We call this constrained dominated control.

As an example of another extreme, consider a polymer for fiber spinning, in which the fiber has tough specifications for strength and uniform disability. If product specifications are narrow and difficult to meet, the total emphasis of the operator will be in meeting product specifications. If he does not meet them the product may be practically worthless. The operator will insist on a very narrow operating space and tough specification on feedstocks.

We still may deal with partial control in both cases, but the emphasis is different and so is the approach of the operator. Our methodology with proper modifications is suitable for both cases but one has to understand these differences.

In the beginning, the FCC was totally constraint controlled. The main problem was to get it to operate and prevent it from shutting down. In the absence of any real competition any result was acceptable especially as the need for gasoline was still limited. Furthermore, the quality of the gasoline was superior to that from conventional distillation, as it had a higher octane. Gradually all this changed. High conversion became important, as well as the ability to treat heavier feeds. These demands can still be looked at in terms of process constraints as they do not involve concern with tight specifications.

But the recent changes in the industry have caused a much greater concern with product specifications. High octane became important. With tighter limits on variations, limiting the aromatics content in the gasoline fraction has become important. In some cases it is therefore important to increase the feed to the alkylation unit (or maximizing C_4 and C_3 olefines). In others one worries about the quality of the distillate fraction produced (light cycle oil). The industry in general is changing from an industry dominated by process constraints and costs to one in which meeting tight product specification is receiving almost equal consideration. Process constraints have also tightened, as now there are specifications on CO and NO_x and SO_2 in the flue gas. Thus we deal here with a case of mixed control, where Y_p contains important specifications as well as constraints. However, the specs are not hard in the sense that temporary

violations can be mixed out. In the fiber case the polymer not meeting specs may be scrap. In many other processes we have a mixed control system in which both process constraints and product specification enter our control consideration. This introduces new challenges into the control of different units and especially the FCC.

IIc Relations of G_s^s to G_d or Mapping Y_p into Y_d^s

We noted before that the variables Y_d entering G_d and the partial control matrix are not necessary at all part of Y_p . They may contain some of the constraints but seldom the specifications. In many processes such as polymerizations, one of the most difficult tasks is to relate the final product specification to both G_s^s and G_d . If a reliable complete model is available this is relatively easy. But we are also concerned with the much more frequent case where such a detailed reliable model is unavailable. Here, we have to rely on physical insight and data available, from a laboratory experiments. In complex polymers this is the most difficult task. In an FCC this is far easier. Today, there are detailed, reasonably reliable models available for the FCC, but even years ago some of the relations were at least qualitatively well known. Still, the models are far from perfect. Therefore we need feedback and online updating of G_s^s .

In dynamic control we often can get satisfactory control using an experimentally determined response function for the unit. We can do the same for G_s^s . There is a whole literature on design of statistical experiments to do so, though statistical correlations based on a physical model are far superior. We just have to remember that G_s^s normally has to be nonlinear, whereas for G_d a linearized dynamic model is sufficient. However, in a process like FCC in which large process changes occur, G_d will require retuning or even restructuring. The same can apply to G_s^s where process changes, or new catalyst and even new feedstocks may require retuning or reformulation of G_s^s .

There is another critical aspect of the relation between G_d and G_s . The setpoint Y_d^s used in the primary control matrix must be dominant not only in the dynamic model but also in G_s . The latter is a strong function of the properties of Y_p . In FCC it is often possible to check and verify this on the basis of present models, but in other processes such as polymerization it may require a significant research effort.

II.d Dynamic Matrix Control and Quadratic Matrix Control

There have been two important recent developments in multivariable control that are strongly relevant to partial control and in fact deal with some of the problems discussed here. These are Dynamic Matrix Control or DMC [Cutler and Ramaker (1979)] and Quadratic Matrix Control or QMC [Prett and Garcia (1988)] which is a further development based on it.

Both are based on linear models and deal with the partial control problem using a linear model. DMC allows changing a setpoint while staying within all the constraints of the system. It also allows adjusting the setpoints in a way that the system always operates against several constraints. (For an FCC, for example, it will operate at maximum air rate and maximum wet gas compressor loading.) In QMC, one also introduces quadratic objective function for local optimization. Both methods can be further integrated in a model based nonlinear optimization scheme, based on steady state optimization of all inputs including what we call slow variables.

As both methods have been successfully employed in advanced FCC control schemes as well as other refinery units, they have to be considered in any study of partial control. We will deal with them in detail in future publications and in a more limited way in this paper. We will also utilize their methodology when dealing with actual control schemes in the FCC or other processes. The question then arises in what way is our approach different. We here simply ask a broader question about the framework and conditions that allow DMC or QMC to operate. We look at the usefulness and applicability of these linearized methods in terms of the full nonlinear system, and how they can be integrated in the nonlinear G_s^5 .

The relation between the two approaches will become clearer as we go along but here we will make some qualitative clarifications.

1. The success of both QMC and DMC depends on the manipulated variables available as well as their range. The question of sufficiency and adequacy of a given set of manipulated variables to control and stabilize the system is a nonlinear problem that has to be answered separately.
2. While both methods do not require choice of a square control matrix and can use the whole set of measured variables to manipulate the inputs, they often are used within the context of conventional control in which a square matrix is chosen for setpoint control. Dynamic matrix control is then equivalent to setpoint control adjusting the setpoints of the square matrix to

optimize the system within the linearized subspace. However, while there are criteria to choose the measured variables entering the square matrix solely based on the linearized model, these are sometimes not sufficient in a highly nonlinear system. When a system has multiple steady states, a perturbation in feed conditions may cause the loss of the desirable steady state. In this case, if one has the choice between two partial control matrices, one would prefer the one which allows a larger perturbation at a given steady state without losing it.

In such a system one also has to ask the question, what settings are permissible? One also has to avoid input multiplicities [Koppel (1982)]. These are nonlinear questions strongly impacting the success of a DMC controller that have to be asked using a nonlinear model. The question of minimum model information and the impact of design on controllability are also inherently nonlinear questions that require an answer outside the framework of DMC.

As DMC uses linearized local tuning it is important to know the range of operating conditions or controller settings in which such a local linearized model is valid. We will also show that in nonlinear systems that undergoes significant changes, different partial control matrices have to be used for different operating regimes. Thus our approach does not compete with either of those methods but uses them in a more general nonlinear context.

III. CONTROL OF AN FCC

III.a Description of the System.

For a detailed description of the FCC, its historic development and the different types of units in current use, and the design philosophy underlying them see Avidan and Shinnar (1990) and Avidan, Edwards and Owens (1990). Here, we will suffice ourselves with 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 present 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, and 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 column and 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 adsorption 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 and it is therefore important to include their yields in Y_p .

One unique feature of the FCC reaction 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 permanently due to steaming and exposure to higher temperatures. The first problem is solved by the design in which the sojourn time of the catalyst in the reactor between regenerations is measured in seconds. The long range deactivation is taken care of by daily removal and addition of catalyst. The average total sojourn time of the catalyst in the system is two to three months.

In fixed bed reactors catalyst activity can only be changed 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. This also allows a fairly rapid change to a different catalyst.

Modern FCC's have developed another important control strategy based on this feature. Various catalyst additives are 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 olefines and paraffins resulting in higher octane and a higher C_4 , C_3 olefine 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 these modifiers can not be incorporated into the main catalyst. However, adding them separately increases the flexibility of reactor control. Catalyst management is therefore an important and crucial part of slow steady state control.

The design of the reactor varies but basically almost all present reactors are risers with relatively short contact time of both catalyst and feed. The main difference between designs of interest to us 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 fast and accurate control of reactor temperature. It also allows a variation of catalyst flow over a relatively wide range (factor of 3). In some older units catalyst circulation can only be adjusted over a narrow range by the pressure balance, the key variable being the catalyst holdup in the regenerator. While this still allows for some variation of catalyst flow, it is too slow to be useful for dynamic control.

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 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 increased throughput, despite constraints in the maximum catalyst flow rate and air rate.

Some units also have different 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 1200°F)[see Part I*] about 50% of the carbon ends up as CO which has to be combusted in a boiler before being emitted to the atmosphere. As combustion of coke to CO₂ generates three times as much heat (and uses twice the air) compared to its combustion to CO, operating in partial combustion reduces both the heat generated in the regenerator as well as 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.

Controlling feed temperature also allows some dynamic control of reactor temperature in old units with no slide valve. It is much slower than a slide valve (minutes instead of seconds) but is still relatively fast compared to the time scale of regenerator response. Most units have a stripper that removes adsorbed hydrocarbons before they enter the regenerator. Catalyst residence time in the stripper as well as the steam catalyst ratio can be varied adding another control to the heat balance.

*Part I is Huang, Arbel, Rinard and Shinnar, "The Impact of Design and Operating Conditions on the Existence of Multiple Steady States and Input Multiplicities in Fluid Catalytic Crackers", in print.

Regenerator designs have a wider variation. There are two basic designs. One is a fixed fluid bed operating at high velocity (3 ft/sec) and therefore a high recirculation rate through the cyclones (catalyst inventory circulates through the cyclones approximately every 5 minutes). This gives a high catalyst density (0.5 - 1 lb/ft³) in the dilute phase which is important for the regenerator operation. The other uses a circulating fluid bed regenerator such as described in Fig. 2, with multiple air inputs. The division between fixed and circulating fluid beds is here somewhat arbitrary as in the fixed bed there is also a large catalyst circulation. However, in the circulating bed catalyst densities in the riser sector are much higher (10 lb/ft³ versus 1.0) and catalyst recirculation is independently adjustable.

III.b The Dynamic Model

The equations and assumptions of the dynamic model used for both steady state and dynamic computations are identical to those described in our paper on multiple steady states [Part I, The reader is referred to it for details.]. The main differences to previously published models [mainly McFarlane et al. (1993)] are:

1. A more detailed description of the dilute phase in the regenerator.
2. A detailed description of the reaction kinetics in the regenerator especially the combustion of CO to CO₂.
3. A more detailed kinetic model of the reactor. The reactor is also assumed to be in plug flow and therefore has a temperature profile.

We have to point out that this model while quite complex still contains some strong simplifications which we will deal with in future papers. The most important one is that our model assumes a regenerator which is uniform across a cross section. In large fixed bed regenerators this is difficult to achieve and while temperatures are quite uniform, coke concentration is not. This introduces control difficulties as it increases after burning. Also the kinetics of coke formation and combustion are much more complex than described in the literature. We do not distinguish between coke after regeneration and fresh coke in the reactor nor do we deal with other fine details. We have not yet introduced a detailed model of the stripper and therefore neglect the impact of catalyst circulation rate or stripper performance. But our model, as it stands, is complex enough to describe the basic physics. The interactions are so

complex that it is advisable to understand them first with help of a simpler model. However, some of the models published oversimplify the reactors to a degree that some of the main effects are lost.

We use the model in two ways. One is to describe the actual system behavior or, in other words, as a substitute for the real system. Thus, we test control loops and control concepts using the model. In a real design case the exact model of the real system is unknown. We will therefore discuss in a later paper how we can design a control system with information obtainable with conventional identification methods. Before going into a quantitative analysis let us discuss what are the essential features of the process that can be understood from laboratory experiments without a detailed quantitative model.

The most crucial property for control is that the system is adiabatic and has more than one steady state [see Part I]. In fact, when it has only one steady state, this steady state is trivial. That means that only a certain limited space of input conditions or control settings is permissible. Furthermore the heat balance and the variables entering the heat balance are crucial. For the purposes of the heat balance we have to consider separately the reactor and the regenerator.

1. Regenerator

- a. Combustion of adsorbed hydrocarbons is very fast. For simplicity we lump them here with the slow burning coke. The combustion of coke is slower and occurs almost exclusively near the solid surface. It leads to a mixture of CO_2 and CO , the ratio of which is mainly a function of temperature [Weisz (1966), Arthur (1951)]. The surface combustion rate is a strong function of temperature with an apparent activation energy of 37.5 kcal/g mole. The temperature required for sufficiently fast combustion for conventional FCC operation is above 1150°F. Therefore, upper steady states with lower regenerator temperatures are of no commercial interest.
- b. Homogeneous combustion of CO is very slow at temperatures below 1200°F, becomes significant in the 1250°F range, and fast above 1300°F. Combustion promoters allow operation with complete CO combustion at lower temperatures.
- c. Combustion to CO_2 has a heat of reaction of 169,080 Btu/lb-mol versus 46368 for combustion to CO , a ratio of 3.6. Per mole of oxygen the ratio is 1.8. The CO_2/CO

ratio is therefore a crucial parameter affecting the heat balance. A kinetic model is required to predict it. At temperatures below 1200°F this is less important.

As shown in the paper on multiple steady states, the homogeneous combustion of the CO to CO₂ is a potential source of excess steady states at higher temperatures (five steady states instead of three or four high temperatures steady states instead of two), but the likelihood is small.

- d. At temperatures below 1250°F it is hard to get complete oxygen conversion in the regenerator itself. Any unconverted oxygen will react instantaneously with CO present once the catalyst is removed resulting in after burning. The temperature difference across the cyclone is therefore a good indicator of unconverted oxygen and can be used for control at lower regenerator temperatures.
- e. The amount of carbon left in the regenerated catalyst, C_{rc} , has an impact on catalyst activity in the reactor and is therefore an important state variable.

In presently used catalysts the catalyst properties have only a small effect on the rate of coke combustion. The main impact of catalyst properties is on the CO to CO₂ reaction which can be catalyzed by combustion promoters. We can take care of this effect by introducing a catalyst parameter that enters the pre-exponential of the catalytic CO to CO₂ combustion rate. This parameter can be adjusted and maintained by catalyst management, using the CO₂/CO ratio in the flue gas to estimate it.

2. Reactor

For the heat balance, and therefore dynamic control, the main interaction between regenerator and reactor is the amount of coke formed. One can find from laboratory experiments that there are five crucial parameters that determine the coke make.

- a) The time the catalyst spends in contact with oil.
- b) The temperature history of this stay.
- c) The amount of coke on the regenerated catalyst.
- d) The feed composition.
- e) The state and properties of the catalyst.

One can to a good approximation neglect the state of the hydrocarbons and the severity of the reactions. The only way the extent of the reaction enters the heat balance is via the enthalpy of the products. As the difference in enthalpy between different product distributions is not very large we can get away here with strong simplifications. This is a crucial and important feature of FCC catalysts that was discovered quite early [Voorhies (1945)]. It is not at all an obvious property. There are many cracking reactions in which the coking rate strongly depends on the product composition and therefore on the extent of the reactions. This feature may therefore be limited to present catalysts and feedstocks.

Another crucial feature is that for a given catalyst, the extent of the reaction (or the conversion) is strongly correlated with the coke formed, the relation being a function of temperature and initial catalyst state.

Coking, cracking, light gas formation, and product composition are all strong functions of the type and activity of the catalyst. Metals such as Nickel and Vanadium deposited on the catalyst also affect the coking rate. However, only coke on regenerated catalyst enters the dynamic equation. Long range catalyst deactivation by steam and changes due to metal deposition are slow compared to the dynamic response in coke formation due to changes in operating conditions or feed composition. The deactivation of the promoter is also slow. We can therefore decouple the changes in catalyst state from the dynamic model and use the state of the catalyst as an input. For steady state control we need however a separate model describing the state of the catalyst. For coking that can be expressed as a two parameter function, one giving the overall activity of the catalyst, the other a relative coking rate.

Riser top temperature, as well as riser bottom temperature, are critical parameters that affect not only coke formation and conversion, but the whole product slate. As different reactions have different activation energies, temperature is the main variable controlling product composition and properties. High temperatures give more light olefines, and also a higher octane as lower octane components are cracked out.

At higher riser temperatures (above 1000°F), thermal cracking becomes important [Avidan and Shinnar (1990)]. This is minimized by a short residence time. A high bottom temperature achieved by a high T_{rgn} is therefore desirable as it exposes the feed to very high temperatures for ultra short times; the endothermic heat of reaction reduces the temperature very fast. Aside from

catalyst composition control of T_{ms} (top and bottom) is therefore the main tool to control product composition.

The coking capability of the feed is a two parameter function that for present catalysts depends only on feed composition [Weekman and Nace (1976)] and can be decoupled from catalyst properties. One deals with kinetic formation of coke as a function of catalyst residence time and cat to oil ratios. The other parameter gives the contribution of Conradson carbon, the total formation of which is assumed (as an approximation) to be independent of residence time and cat to oil ratio.

The temperature of the oil feed to the reactor also has a strong impact on the heat balance. Increasing feed temperature reduces the catalyst circulation by reducing heat requirements at constant riser temperature. While reduced circulation reduces coke make and thereby air rate, it also reduces conversion. One can also reduce catalyst circulation by increasing regenerator temperature which has a desirable effect on the reactor. It increases the bottom temperature in the riser. This reduces coke make at constant conversion and has other complex effects on the products. In most units feed temperature is used only for slow control and we can consider it as an input. In some older units which have no slide valve it is sometimes used to control reactor temperature.

However, while dynamic control and stability is based on the heat balance, we do not run an FCC as a boiler. Its goal is to crack heavy oil to lighter products. When the FCC was developed any performance was acceptable as it was better than all known alternatives to increase gasoline yield from the heavier fractions. The acceptable product space was therefore very simple.

Today, life is not so simple. Modern catalysts and FCC technology allow processing of a wide variety of feeds and allow changes in product composition over a fairly wide range. And such changes can be achieved over fairly short time spans measured in days. No other unit in the refinery has similar capabilities. Therefore the FCC has become a main tool of the refinery for changing quality product distribution and product specification.

Predicting product composition is a much more difficult job than predicting the heat balance. Several thousand compounds are involved and there are various procedures of lumping involved to make the job feasible. The details of the kinetic lumped reaction rate schemes used

for fitting the pilot plant results depend therefore on our goals and on the required details of the composition space that enter our specifications. We use here a simplified 10 lump model proposed by Jacob et al. (1976). To fully discuss partial control to achieve a desired product composition one has to modify this model to predict octane and gas composition. This modification will be however, the subject of a future paper as here we deal with some more limited aspects of the controller design for which the 10 lump model is sufficient.

For predicting product composition, the quantitative description of the state of the catalyst requires a larger dimensional vector than required for predicting the heat balance. For the heat balance three parameters are sufficient. The number required for product composition depends on our goals and could include following the state of additives separately using a separate dynamic model (with a slower time scale) to follow catalyst management. For the dynamic model used for the primary control matrix, we can use this state vector as an input.

Last, as the residence time in the riser is very fast compared to dynamic changes in the system as a whole, we can assume pseudo-steady states in the riser.

IIIc Manipulated and Measured Variables, and Definition of Specification Space.

In Table 1, we give a list of variables that can be potentially manipulated, both slow and fast, and variables that enter the specifications either as products specifically or as constraints, and state variables that affect the operation and can be either measured or inferred.

The manipulated variables that we use for dynamic control are few. We normally try to avoid use of raw material feed rate in a dynamic control matrix and feed composition can only be varied slowly. Some variations come in as a disturbance. However, one often uses feed rate as a manipulated variable for steady state control.

Catalyst composition and state are also a slow variables by nature. We also want to avoid fast changes in reactor pressure. So we are left with air rate, catalyst circulation rate, feed temperature, and, in units equipped with a catalyst cooler, the rate of heat removal in the cooler. There is, however, a tendency to use control of the cooling rate in catalyst coolers for slow steady-state control. This leaves us with three manipulated variables for fast control: air rate, catalyst circulation rate (which are normally preferred in modern units) and feed preheat. The

question is are these sufficient? We can only answer that by looking at the dynamic behavior of the system.

Table 1: List of Variables in FCC

Manipulated Variables	Fast	Slow
	Air flow rate Catalyst flow rate Feed temperature Catalyst cooler	Catalyst addition rate Feed rate Feed quality Combustion promoter
Specifications	Product	Constraints
	Octane Conversion Alkylation feed Gasoline yield Light cycle oil yield and properties C ₂ Isobutene, propylene to propane ratio	CO, NO _x , SO ₂ in flue gas, Wet gas rate Air rate, Cat. circulation rate Flue gas temperature Riser temperature
State variables	Riser temperature Regenerator temperature Coke on regenerated catalyst Coke on spent catalyst ΔT across the cyclones	

For any nonlinear system we have to first find out if there are multiple steady states. This is discussed in detail in Part I. We deal with this problem in a separate paper due to its

complexity. We showed that there is a chance of five steady states but, for acceptable operation, we need at least three. In the case of only one steady state, this state is trivial (a cold system with no conversion).

It is perfectly possible to find a reasonably wide space of operating conditions in terms of air rate and catalyst circulation rate in which the desirable steady state is open loop stable. It may even be stable for disturbances of quite sizeable magnitude, provided the operating point is far enough from the stability limits. This is shown in Fig. 3. For fixed circulation rates one can lose the upper stable steady state by either losing catalyst activity due to insufficient catalyst additions and withdrawal or by getting a feed with a lower coke making tendency.

However in most cases, the FCC is operated under automatic feedback control. The variables that have been suggested as measured variables in a square 2x2 control matrix using air rate and catalyst circulation rate as manipulated variables are given in Fig. 4a. Different combinations have been suggested and are used for different operating conditions. Other variables such as coke or regenerated catalyst have also been suggested but we limit ourselves to the above forms. Fig 4b shows a schematic control configuration on an FCC, using the above variables.

From the measured variables listed in Table 1 only the reactor and regenerator temperatures are dominant variables. The others are outputs that can be used for inferential control. Reactor temperature is used in all present control schemes either directly or by cascading.

If we look at different choices of the manipulated variables there are several criteria that can be judged based on the linearized transfer function. In addition in a nonlinear system, there are some fundamental criteria that have to be judged separately. It is easier to discuss these criteria by quantitative examples. This we will do later. But, in order to provide a framework for the reader, let us at least define some important criteria.

1. For nonlinear steady state control it is preferable to choose dominant variables for the setpoints. Output variables can be used if they are strongly correlated with the dominant variable desired. This is especially so if the model is imperfectly known since the operator will have more physical insights to choose the setpoint of a dominant variable.

2. It is important that for a reasonable range of changes in the inputs (such as feed conditions or changes in catalyst activity which occur in an FCC), the system stays stable in the sense that the setpoint remains permissible. The stability range can be different for different partial control matrices.
3. It is desirable to avoid input multiplicities in the manipulated variables [Koppel (1982)], at least in the range of achievable inputs. However, such input multiplicities are not always detrimental. If there are only two possible steady states, and those have opposite signs in the controller gain when using linear control with integral action, one would be unstable and would not be an attractor [Morari and Zafiriou (1988)]. Output multiplicities are harder to avoid as the system has multiple steady states and the cold state is unavoidable. The control should not increase the number of stable steady states.
4. It is important to avoid setpoints in which one manipulated variable has either a minimum or maximum when plotted against the setpoint of the measured variable at steady state conditions. If such a minimum (or maximum) occurs in the same region for several variables and especially for a dominant variable the system would be hard to control at those operating conditions regardless of which set of measured variables is chosen for the partial control matrix. In the same way, the system must be integrally controllable at a desired setpoint.
5. It is important that for a given partial control matrix, the changes in Y_p remain acceptable for a significant perturbations in feed properties and catalyst activity. This is related to dominance. This point can also be at least partially investigated in a linearized model.
6. It is desirable that changes in the linearized control matrix are small when operating conditions change. Some choices of G_{cd} will have a wider range of setpoints in which controller settings remain acceptable than other. Finding such sets is useful up to a point. No choice however will be globally applicable. Large enough changes in the operating conditions will, at a minimum, require the system to be retuned and, at worst, require a change in the measured variables entering G_{cd} and perhaps even the manipulated variables. It is important to understand if and when this is needed.
7. There is one more condition that is critical for the choice and, after stability and controllability, the most important one. The variable chosen for the setpoints must relate strongly to Y_p via G_s and the relation should be obtainable by laboratory experiments. The criticality of

the criterion strongly depends on the nature of Y_p . In other words, Y_{ca} has to be mappable to Y_p via G_s .

The reader will note that fast response was not listed as one of the essential criteria. The chance of finding a matrix that fulfills all the above requirements is quite low; one has to set priorities and make compromises. While for some properties fast response could be an important advantage, its priority in the compromises is low unless Y_p is dominated by a single hard constraint.

We also have to be careful when talking about fast response. As mentioned before, if one chooses ΔT as a measured variable one can achieve faster response than for T_{reg} [Hovd and Skogestad (1991)], but the overall system response is still slow. We do not discuss here fast intermediate loops which can be used for stabilization or for smoother response. Thus, ΔT is often used as a fast internal variable to protect the cyclone. This will be discussed under tuning or designing the control algorithm itself as compared to the choice of the matrix. As regenerator temperature is a dominant variable its response time dominates that of most output variables. It is slow as can be seen from Fig. 5.

When designing an adiabatic reactor, one often intentionally increases the time scale of temperature changes by increasing thermal inertia to protect the unit from crashing. The model is always incompletely known and the control imperfect. We want sufficient time to correct mistakes and prevent the unit from wind down. Here, moderately slow response is beneficial. We want at least one to two hours, but not necessarily six. On the other hand academic process control research is so dominated by the concepts introduced from aerospace, where fast control is always essential, that we sometimes lose track of the fact that unlike missiles, the designer introduces slow response intentionally into the system.

There is a good reason for this difference in philosophy. In a missile, the value of fast accurate response and control is high and the price of missing the target acceptable. In an FCC the opposite is true. The cost of one wind down is higher than the potential profit over three months. That does not mean that in an FCC (or any adiabatic reactor) we do not need fast control circuits. Pressure has to be maintained fast and accurately; so must feed temperature. Reactor temperature control is inherently fast. We also need fast control to enforce constraints.

The slow response in Fig. 5 is initially surprising as all residence times are measured in either seconds or minutes. Such response times are typical of systems with internal recycle and also of adiabatic systems with large thermal inertia in which any changes in steady-state temperature will cause a change in the stored thermal energy. Fig. 5 also gives the response to a step input on air when reactor temperature is kept constant by control. This changes the sign of the response compared to the open loop due to the interaction with the reactor.

IV. COMPARISON OF DIFFERENT PARTIAL CONTROL MATRICES

We can now proceed to evaluate different dynamic control matrices for our base case where the on line manipulated variables are air rate and catalyst circulation rate. We gave the different pairs in Fig. 4. We want to point out immediately that from the pairs given only by T_{rgn} , T_{ns} meets condition 7. All mapping of Y_p into Y_{cd} that he can obtain from pilot plant experiments will only have information on T_{rgn} , T_{ris} , C_{rgc} , cat circulation and related variables. ΔT and T_{sg} are strongly model and unit dependant and not mappable.

After the preliminary discussion we can now evaluate the different matrices qualitatively.

IV.a Controllability

There are several aspects of this question. First of all, is the system controllable with air rate and catalyst circulation rate, keeping the other inputs constant? It is a priori obvious that it is not globally controllable in the classical sense. The system has multiple steady states. The only globally controllable region is the one with a single steady state which here is a trivial case (no reaction). There is no way that by changing air rate or catalyst circulation rate, we can move from the cold state to the desired reaction state. All we can hope for is local controllability for one of the useful steady states. For steady state control it is here desirable but not critical that the steady state is stable. We can always devise some scheme that stabilizes it.

Let us therefore look at the controllability of such a local region and we will start with $[T_{rgn}, T_{ris}]$. We could do this by giving a plot of T_{rgn} , T_{ns} as a function of air rate and catalyst circulation rate. This is hard to do as we would require a four dimensional plot. In most cases the priority is given to reactor temperature control. Therefore we look at isoclines of constant T_{ns} . A two dimensional plot using air rate as the abscissa is sufficient. Fig. 6 gives such a plot of

T_{rgn} versus air rate for two isoclines of T_{ris} . The open loop stability of each steady state is indicated in the plot, but does not represent closed loop stability.

For local controllability it is important that

1. For given desired steady state, a set of fixed values for air rate, catalyst circulation rate, T_{ris} and T_{rgn} leads to a unique steady state in all other variables.
2. Fixing T_{ris} and T_{rgn} determines F_{air} and F_{cat} (no input multiplicity).
3. The slope of T_{rgn} versus air rate must be reasonable. Slopes of either zero or infinity lead to loss of control.

The first question cannot be answered by the plot. In our model, the answer is yes. Given a consistent set of inputs, the steady state is unique. However, one can conceptually construct a reasonable catalyst using known phenomena in which this is not true. Isothermal reactor with fixed inputs can have multiple steady states. We are not aware of any present FCC catalyst which has this property. In general, this is one of the most important questions one has to ask when controlling complex systems.

For $[T_{ris}, T_{rgn}]$ as a control variable to give a unique steady state it is therefore sufficient that there are no input multiplicities. If we fix (by feedback control) T_{ris} and T_{rgn} , Fig. 6 shows that at some critical T_{rgn} for any given T_{ris} there is a region of multiplicities close to $T_{rgn} \max$. For a given $[T_{rgn}, T_{ris}]$ we have two pairs of $[F_{air}, F_{cat}]$ which give the same open loop stable high temperature steady state. However as the steady state gains have opposite signs, a controller with integral action is only stable for one of these states, depending how we choose to set it. In that region at or close to complete CO combustion, T_{rgn} is not a suitable control variable as the gains are too small. There are also other reasons which will be discussed later. Here, we have to use excess oxygen control. Actually, the region of multiplicities is theoretically larger than plotted. At very high air rates, we have large values of excess air and can have a stable steady state at T_{rgn} significantly lower than $T_{rgn} \max$. As one always operates an FCC close to maximum air rate (closed snort), these are not reachable steady states and can be disregarded in a real system. For most of the region, a set value of $[T_{ris}, T_{rgn}]$ will completely determine the system. This is not true for any of the other combinations as will be shown later.

The third item can be seen from Fig. 6. There are two regions in which the system is locally uncontrollable. There is a maximum temperature with air rate at which the slope is zero. There is also a minimum air rate at which the slope is infinity.

The minimum regenerator temperature in Fig. 6 is determined by the maximum permissible catalyst circulation rate. At the zero slope point the system is controllable by measuring excess oxygen, as we will see later. However, closer to the region of infinite gain (minimum air rate) the system is not controllable locally with air rate and cat circulation rate. We need here another manipulated variable such as feed temperature. Luckily there is no special advantage to operating in that area. The problem however remains that with limited model information the operator does not know where that inoperable region is, especially as changes in catalyst activity move it quite strongly. Some minimum knowledge of nonlinear model behavior is essential.

T_{rgn} as a function of air rate has one interesting property which is counter intuitive. In the partial combustion region (lower T_{rgn}), T_{rgn} decreases with increasing air rate. In complete open loop (fixed air rate and cat circulation rate) the opposite is true (see Fig. 3). Both T_{rgn} and T_{ris} increase with increasing air rate. This difference in response is caused by the fact that we plot along an isocline of T_{ris} . For steady state control this makes sense as it is important to control T_{ris} closely. If T_{rgn} decreases, F_{cat} has to increase at constant T_{ris} . Increasing F_{cat} will generate more coke. The only way we can combust more coke is to increase air. Conversely, if T_{rgn} increases cat circulation decreases, coke make decreases requiring less air for combustion. At higher T_{rgn} the CO_2/CO ratio increases supplying the required heat with less air and less coke. In open loop cat circulation is constant and high T_{ris} will generate more coke.

While the plots of T_{rgn} versus air rate have a similar form for both riser temperatures the lower riser temperature shows an open loop unstable steady state at a regenerator temperature below 1180°F. Stability will strongly depend on controller tuning. Also the maximum regenerator temperature is lower for $T_{ris} = 950^\circ F$.

IV.b Input and output multiplicities for $[T_{ris}, F_{cat}]$ and $[T_{rgn}, F_{air}]$.

We can now look at other combinations of two variables in the set $[T_{rgn}, T_{air}, F_{air}, F_{cat}]$ in terms of input multiplicities. At first we will look at $[T_{ris}, F_{air}]$. This is a practical control case today since many units try to operate at maximum air.

Fig. 7 gives the possible steady states for T_{ris} and T_{rgn} for a specific case (fixed feed properties, feed rate, etc.) when air rate is fixed and catalyst circulation varied. Open loop stability is indicated in the plot. There is a range of T_{ris} and T_{rgn} for which a steady state exists. In that range, we can always find two catalyst circulation rates giving the same T_{ris} but with different T_{rgn} (for example, point A and B). While both are open loop stable, one cannot find a linear controller with integral action controlling T_{ris} by changing F_{cat} which is stable at both point A and B. The steady state gain of the controller has an opposite sign. We would normally design for point A which is in partial CO combustion since point B is in total CO combustion and one would normally not operate with constant air rate in that region.

Edward and Kim (1988) describe such a case in which a system operating in partial combustion drifted to another steady state, when the controller controlling T_{rgn} by manipulating F_{air} was put temporarily on manual. The reason for the drift is that when both loops are closed, the gain used in tuning the controller for T_{ris} is positive which is correct for constant T_{rgn} . When air control is put on manual the gain become negative and the control loop with integral control becomes unstable at point A and either drifts to point B or leads to a wind down.

We can also look at the reverse case where T_{rgn} is controlled by air rate and catalyst circulation is fixed (fixed $[T_{rgn}, F_{cat}]$). This happens in the old Exxon units (Model IV) in which catalyst circulation is fixed and can only be dynamically varied over a narrow range. A wider variation requires change in the catalyst inventory and is therefore only part of the slow control.

A plot of multiple steady states for the similar conditions as in Fig. 7 is given in Fig. 8 for constant F_{cat} . Again for a given pair $[T_{rgn}, F_{cat}]$ there are two possible values of F_{air} and T_{ris} , one (point A) in partial and the other (point B) in full combustion. Therefore for both cases, specifying $[T_{ris}, F_{air}]$ or $[T_{rgn}, F_{cat}]$ does not specify a unique steady state.

IV.c $[T_{ris}, \Delta T][T_{ris}, T_{sg}] [T_{ris}, O_2]$

The plots of ΔT , T_{sg} and $O_{2,sg}$ corresponding to Fig. 6 are given in Fig. 9. Let us first look at ΔT . Unlike T_{rgn} , a fixed ΔT at a given T_{ris} can correspond to more than one air input over quite a range of ΔT . In fact, in some range there can be three steady states. For $T_{ris} = 1000^\circ\text{F}$, multiplicities disappear at both low and high values of ΔT . Some one would only use ΔT in partial combustion, this gives a reasonable range of high values of ΔT in which ΔT is useful for protecting the cyclones. ΔT responds faster than T_{rgn} as shown by Hovd and Skogestad (1991). However, this fast response is only useful for protecting the cyclone, as the response of T_{rgn} is slow regardless of the variable chosen for the partial control matrix. Therefore, other variables will also change slowly and despite the fast response of ΔT , the system control is still slow.

In the useful region of ΔT control, the input multiplicities are not necessarily detrimental, as the other two steady states have gains of opposite sign. A properly tuned controller will therefore not be attracted to one of these states. At higher values of T_{rgn} (or lower values of F_{air}) there are multiplicities with the same sign close together. Here ΔT is an unsuitable variable. The minimum value also occurs at values of F_{air} and T_{rgn} , which are still in partial oxidation and useful operating regions where T_{rgn} is a good control variable. ΔT is therefore limited to lower operating temperatures. We, however, have to admit that in that region our model is not very good in predicting ΔT . In the region where the homogeneous combustion rate is high, it is hard to get good rate data for CO combustion. At low values of CO, the impact of the CO combustion rate on the heat balance is very small, but ΔT can still be quite large. In that region, it is advisable to use a combustion promoter which will be discussed later.

If T_{ris} is 950°F , then for our case in the useful region of ΔT control, the unit is open loop unstable. There is another caveat. Our model, similarly to other models given in the literature, describes a unit completely uniform across the diameter. It is very hard to have complete mixing of the catalyst over a 35 foot diameter when the total residence time is about five minutes. Depending on design there can be differences in the local carbon content of the catalyst. In regions where it is lower than the average, air conversion will be lower given a higher ΔT in the cyclone above that region. Gross et al. (1976) discuss how to partially alleviate this problem by modifying the air grid.

The nonuniformity strongly depends on the design and the size of unit and a number of designs have much smaller variations. Our model describes the simplest case that was treated by other investigators. When nonuniformities are high, one or more cyclones are hot and in the past one had to use ΔT control to protect them. Process considerations become secondary. In FCC's before modern catalyst were available, two considerations dominated control. One was to prevent snowballing or crashing; the other, to protect cyclones. Today, better materials of construction can tolerate higher cyclone temperatures and the chance of snowballing is low with modern catalysts. If needed, we can also reduce the hot cyclone problem using combustion promoters in partial combustion. If C_{rgc} is low, the excess oxygen will react with CO in the dilute phase.

On the other hand even for cases where ΔT is hard to predict, our model describes the dependence of T_{rgn} on air rate quite reasonably. Even in the cases where ΔT control is essential for protection of the cyclone, it is always used together with controlling T_{rgn} . One either uses closed loop cascading, or controls T_{rgn} in a narrow region:

$$T_{rgn, \min} < T_{rgn} < T_{rgn, \max} \quad (9)$$

by changing the setpoint of ΔT or by using one of the slow control variables ($T_{ns}(\text{set})$, oil feed rate, catalyst activity, feed composition).

Another reason for not relying on ΔT as a main control variable in the partial control matrix is that ΔT is not mappable into G_s . We have no real way to predict its value from laboratory experiments and therefore cannot predict the desired setpoint for a specific unit. This can be seen in Fig. 10 from plots of T_{rgn} versus the setpoint of ΔT .

ΔT is here the input and T_{rgn} , the output. As the $[\Delta T, F_{air}]$ control loop has input multiplicities, this control loop should exhibit output multiplicities and it does. For local control, input multiplicities are more important and these exist only at very high T_{rgn} . However, we note that for $T_{ns} = 1000^\circ\text{F}$ there is an infinite gain near 1300, which reduces to 1250 $^\circ\text{F}$ for $T_{ns} = 950^\circ\text{F}$. Near it ΔT cannot be used for control. As our model is not accurate for ΔT , the region of T_{rgn} controllable by ΔT many vary.

We note in Fig. 10 that O_2 is only useful in the region where T_{rgn} is not controllable due to zero gain. Therefore it is universally used in full CO combustion as none of the other variables

works here. For dynamic control it is excellent. But the range of setpoints is very limited. In complete CO combustion one also needs to maintain minimum value of excess O_2 to control CO below the environmental limits. On the other hand a high excess will increase NO_x . Thus the permitted range is very small.

O_2 control cannot be used to control T_{rgn} , and excess O_2 cannot be varied for partial control in the same way T_{rgn} is used in partial combustion. This is done by changing the setpoints of other inputs in G_s , such as T_F or catalyst activity. This allows modification of T_{rgn} . It is seldom done dynamically as one does not like to introduce a third dynamic circuit unless one has to. It therefore makes no sense to compare control of T_{rgn} versus control of excess O_2 as in an operating region where one of them is useful the other is not.

Fig. 9 and 10 also give the same plots for T_{sg} versus air and T_{rgn} . T_{sg} is here the most problematic one. It has both input and output multiplicities over the whole range. It is equal to $T_{rgn} + \Delta T$ at high values of T_{rgn} . When ΔT is negligible it is equal to T_{rgn} . At low value of T_{rgn} , ΔT increases as T_{rgn} decreases. It therefore has a minimum with decreasing air rate at which it is not controllable. But unlike the uncontrollable region of T_{rgn} this is a useful operating region. The minimum of T_{sg} is also not predictable.

Output multiplicities are obvious for all three variables. Fig. 6 showed that for a given air rate there are two values of F_{cat} , each giving a different regenerator temperature. Input multiplicities are only important if they have the same sign, which occurs here.

However, what is more important is that the range of useful operation is here more limited than for ΔT . If we look at Fig. 10, we note that the region of infinite gain for the control loop [T_{rgn} , T_{sg}], occurs at much lower values of T_{rgn} . There is another problem with T_{sg} . The range of useful values strongly depends on the value of T_{ris} chosen. We will later show that it is also strongly affected by feed and catalyst properties. Thus it is difficult to predict a desirable setting of T_{rgn} , and once chosen it could easily become a nonpermissible setting. It is therefore not surprising that most units that used T_{sg} have switched to other controls especially as present units have moved to higher operating temperatures for which ΔT is lower.

The reason for using higher setpoints for T_{rgn} is not apparent from Figs. 6, 9 and 10. At constant T_{ris} increasing T_{rgn} has several advantages. It reduces both air rate and cat circulation allowing higher throughputs. One pays for the higher production rate by lower conversion as can

be seen from Fig. 11, where we plot conversion versus T_{rgn} . But one can adjust for this by raising catalyst activity. However, there are other advantages for high regenerator temperature. The riser bottom temperature increases which increases the temperature gradient across the reactor. High temperatures in the bottom have advantages that previously (twenty years ago) were not realized. It improves octane, improves olefine selectivity and also gives a lower coke make at constant conversion. The more detailed features of G_s will be discussed in a future paper. Here we simply wanted to give a rationale for a choice which at first seems counter productive. This also shows the need to model the riser properly. We cannot use a stirred tank model, as it would lose these important interactions.

The move to higher T_{rgn} has made T_{sg} control less attractive and we are not aware of any present use. We also have to point out that where T_{sg} control was used, it was used for reasons not apparent from our model. These were cases for which there are large differences between different cyclones as discussed previously. Today, at higher T_{rgn} one has an option to eliminate these local differences by using promoters. Dynamically T_{sg} has advantages [Hovd and Skogestad (1992)], but these are far less important compared to the problems of mappability and multiplicities.

IV.d Impact of Excess Multiple Steady States

We noted in Part I that in an unpromoted system, five steady states could be possible due to the high apparent activation energy of the CO to CO₂ reaction above 1250°F. However, we cannot predict them accurately with presently available data. Nor is there any experimental verification for the existence of such steady states. We therefore omitted that problem from Fig. 9. We can still ask what would happen if such excess steady states occur at the T_{ris} chosen. For local control, the only problem would be if the steady state would be the unstable fourth one. What this would do is shown schematically in Fig. 12. The fourth steady state creates a potential region of unstable steady states between 1250°F and 1350°F. What this could do is to enlarge the temperature region in which the unit is not controllable due to excessive gain for the $[T_{rgn}, F_{cat}]$ loop. No other measured variables would here be helpful. While we can not predict it exactly, if and when this occurs, the knowledge that it could be feasible is important. Here a small

amount of promoter would solve the problem and eliminate excess steady states. The CO_2/CO ratio would increase but the system would remain in partial combustion.

IV.e Impact of Combustion Promoters

We mentioned that in complete CO_2 combustion, one uses promoters. One also often uses them in partial combustion at high values of T_{rgn} . To understand their impact, we give in Fig. 13 and 14, the impact of promoters on the control. To reduce the number of plots, we plot ΔT and T_{sg} versus T_{rgn} , and only T_{rgn} and O_2 versus air. We give two levels of promoter defined by the rate of catalytic CO combustion relative to the unpromoted case. Unpromoted FCC catalyst also promotes the CO to CO_2 combustion [Weisz (1966)] but at a lower level. The promoter containing small levels (ppm) of platinum can increase this base rate by a factor of 10 - 100. Let us first look at T_{rgn} versus air rate. We note that for both levels of promotion temperatures, output and input multiplicities are still maintained. However, the steep region of T_{rgn} with air increases. While one often uses promoters in partial combustion at higher T_{rgn} one uses full promotion only when operating close to $T_{\text{rgn}}(\text{max})$. In full CO combustion with full promotion one always uses excess oxygen for controlling air rate. When one wants to use full promotion at lower T_{rgn} , such as 1270°F in our case, one uses a lower activity catalyst to reduce $T_{\text{rgn}}(\text{max})$. Feedstocks with lower coking rates reduce $T_{\text{rgn}}(\text{max})$, as can be seen from Fig. 17, which is discussed below.

Promoters were initially introduced to eliminate the need for CO boilers. There are advantages of limited promotion in partial combustion that are not apparent in Fig. 13. One was discussed in Part I on multiple steady states. Even in limited promotion, promoters reduce or eliminate the chance of five steady states and therefore guarantee that the operating points in Fig. 13 are stable. They also improve operability as they eliminate or reduce the problem of hot cyclones due to local maldistribution. There is another potential advantage which is hard to model due to lack of data, though it has little impact on the steady state curve. Between 1250°F and 1320°F in the non-promoted case, the homogenous CO combustion rate takes off as inhibitory effects due to the presence of catalyst are decreasing. At lower values of T_{rgn} the catalyst stops the homogenous combustion by intercepting free radicals formed before they continue to react. At higher temperature this effect disappears. Such inhibitory effects are hard

to model and in the region of take off, T_{rgn} is hard to control. Still it is possible to do so as higher temperatures and full CO_2 combustion were introduced before promoters were known. We will discuss this more in detail in the paper on dynamic control. However, there is a penalty one pays for the use of promoter in partial combustion. Conversion is reduced. This can be seen from Fig. 15, where we plot conversion and C_{rgc} vs. T_{rgn} . We note there is no impact in full CO combustion (beyond $T_{rgn}(\max)$). At low T_{rgn} the penalty is large, but in the relevant T_{rgn} range (larger than 1250F) the impact for moderate promotion is small. The amount of promoter has therefore to be carefully adjusted.

If we look at the control of T_{rgn} with ΔT or T_{sg} we note that for high promoter level, T_{sg} becomes practically equivalent to T_{rgn} , and the loop (T_{rgn} , ΔT) with ΔT as input has a very high gain. For partially or fully promoted units, there is no need or advantage to use either ΔT or T_{sg} as a measured variable. Monitoring ΔT is here useful to monitor the activity of the promoter present.

IV.f Robustness of Control

There are several ways to look at robustness for such systems. In the literature [Shinnar (1986), Morari and Zafiriou (1988)] the term robustness is normally defined in terms of the sensitivity of the control loop itself to model uncertainties and changes in model parameters. But the concept of robustness also applies to other aspects. One is the sensitivity of the results and trends to the model itself. The results cannot be overly sensitive to model parameters, and one has to have a sufficiently correct model. Our results follow trends as observed in commercial units and in this sense, it is quite robust in terms of model sensitivity.

Let us first discuss how sensitive are our results to our specific model assumptions. There are two important features in which our model differs from other models. One is the description of the reactor, which currently in almost all units is a riser. Our model assumes plug flow and describing it as CSTR loses many interactions. However the slope of Fig. 9 will be similar. The second and even more crucial difference is the description of the $CO-CO_2$ combustion kinetics. The CO_2/CO ratio has a strong impact on the heat balance. If one defined it as an input to the model [Hovd and Skogestad (1992)] then one assumes an external model that allows one to predict it. By keeping it as a constant value for a given setpoint, one loses the dynamic effects

that exist even in the linear model and which become important at temperatures exceeding 1250°F. Fig. 16 compares our base case (Fig. 9, 1000°F) with two other cases:

- 1) The model used in this paper with decreased rates of CO to CO₂ combustion (both catalytic and homogeneous). We note that the form is preserved while actual values slightly change.
- 2) A case based on a model published by Erazzu, de-Lasa and Sarti (1979) in which the CO₂/CO ratio is given as a function of temperature. The results are similar to our base case up to 1250°F. The Erazzu model cannot reach higher temperatures which require higher CO₂/CO ratios. As it is a correlation over a limited temperature range, it cannot predict complete CO₂ combustion. However, it is well-known that complete CO₂ combustion is achievable without promoters. Assuming CO₂/CO to be a function of temperature only is fundamentally incorrect for a case in which the ratio is kinetically controlled, and while our model maybe not exact it is the only one which contains the essential physics of the kinetic combustion processes in the regenerator.

It has also been suggested to use the intrinsic ratio of CO₂/CO obtained at the catalyst surface as an input neglecting both catalytic and homogeneous CO combustion [the Arthur ratio (1952)]. This would totally reverse trends, but it contradicts industrial experience.

The model published by McFarlane et al. has a very high and nonadjustable rate for the CO to CO₂ combustion, neglecting both intentional and unintentional changes in promoter activity. It is therefore of limited use for control studies even in the area of full combustion.

There is one area where our predictions are model sensitive. ΔT and T_{reg} strongly depend on the assumption of radial homogeneity and are very sensitive to our model assumptions which assume good catalyst mixing and uniform air distribution. T_{reg} and T_{ris} are less affected. Such effects are highly unit dependent and outside the scope of this paper.

Another aspect of robustness required for good control is that the control cannot be overly sensitive to changes in input parameters otherwise the unit cannot be operated. Changes in catalyst activity and coking rate can be quite significant in actual operation. And as a system with 3 steady states is always sensitive to input parameters at the stability limit, we cannot operate it close to the limit, or we may lose the steady state. The problem in FCC control is we never know exactly where the limit is. What makes the unit operable is that its response is very

slow and therefore the operator when he sees that the regenerator temperature drops has time to make corrective actions.

There are several common perturbations facing an FCC. The catalyst properties such as activity, coking, promoter activity, etc. can change. The other common perturbation is a change in feedstock properties. In a specification dominated process we try to eliminate or reduce such perturbations. Refineries have here different policies. Some strongly limit perturbations in feed properties to be able to better optimize the unit or at least filter them. Others use the robustness of the FCC to absorb changes in feed properties and prefer to shelter other units at the cost of optimization of the FCC.

We noted in Section II the importance of dominant variables. Robustness of the total operation in terms of Y_p is therefore strongly related to dominance. At this point we can explain and demonstrate this in quantitative terms.

In Table 2, we give the changes in Y_p as well as the inputs for a step perturbation in coking rate (30% increase) and catalyst activity (30% increase). One can also interpret any of the perturbations as the opposite case and look at the effect of decrease of activity or coking rate. We choose here two stable base operating points labelled in the table Case I and Case II. For each we look at four control strategies: $[T_{rn}, T_{rn}]$, $[T_{rn}, \Delta T]$, $[T_{rn}, T_{sp}]$ and the scheme proposed by Kurihara (1976) $[T_{rgn}, \Delta T]$. We did not discuss the last one in the previous sections as Lee and Weekman (1972) and Shinnar (1976, 1986) showed that it is not attractive as it leaves the reactor uncontrolled. In those papers this claim was not quantified. If we deal with the concept of robustness in general this case is a very useful example, as T_{rn} is the most dominant variable for Y_p . If we are close to instability and stability is our main goal, it is actually quite a good control scheme. For stability it is crucial to make enough coke to keep the unit hot and Kurihara's scheme ensures that better than any other scheme. But as T_{rn} is dominant for Y_p it should be inferior for maintaining Y_p in a narrow space.

Let us first look at an increase in catalyst activity which is compared in the table for both temperatures. All four control schemes maintain a stable operating point, but for some the changes are quite large. The clearly best case for both operating conditions is $[T_{rgn}, T_{rn}]$ as these are the two dominant variables. The Kurihara scheme shows a large loss in conversion and yield. As T_{rn} drops, octane would also decrease.

All other schemes show a small increase in yield, in some cases at a significant increase in air rate and in the case of T_{sg} control, a large increase in F_{cat} which would not be feasible in most units, which limited to cat/oil below ten. T_{sg} control also leads to unacceptable, large changes in regenerator temperature.

For Case II, T_{sg} control would have another problem. This operating point is close to a minimum and therefore hard to control. T_{sg} is hard to control which applies in our case to the whole temperature range of 1230 - 1280. We neglect this problem here as we deal with steady states. Interestingly this does not apply to the new steady states after the perturbation as the perturbation strongly reduces T_{reg} .

The results of a perturbation in coke rate show similar results. Both $[T_{reg}, T_{ris}]$ and $[\Delta T, T_{ris}]$ are acceptable results with $[T_{reg}, T_{ris}]$ being the better case. In both cases conversion drops. T_{sg} control here leads to even bigger and unacceptable increases in F_{cat} . The Kurihara control again has larger changes in conversion resulting from the drop in T_{ris} .

Table 2: Summary of changes to step-change in Activity and Coking Rate (30% increase in both).

Temperature changes are in deg. F, all others are in percent from the base case.

Part I: Base cases values:

Base Values	Case I	Case II
Riser to temp. (T_{ris}) [F]	1000	1000
Regenerator dense bed temp. (T_{reg}) [F]	1184	1251
Temp. rise across the cyclones (ΔT) [F]	222	107
Stack gas temp. (T_{sg}) [F]	1412	1367
Coke on regen. cat. (C_{reg}) [wt%]	0.148	0.0639
Coke on spent cat. (C_s) [wt%]	0.799	0.852
Conversion (Conv.) [%]	76.4	74.2
Gasoline yield (Y_g) [%]	51.1	49.9
Air/Oil (F_{air}) [lb air/lb feed]	0.63	0.58
Cat/Oil (F_{cat}) [lb cat/lb feed]	8.9	6.45

Part II: Changes from the base cases (Temp. in F, other in percent):

30% increase in activity								
	Case I				Case II			
Controlled	T_{ris} T_{rpm}	T_{ris} ΔT	T_{ris} T_w	T_{rpm} ΔT	T_{ris} T_{rpm}	T_{ris} ΔT	T_{ris} T_w	T_{rpm} ΔT
ΔT_{ris} [F]	0	0	0	-33	0	0	0	-40
ΔT_{rpm} [F]	0	-36	-65	0	0	-39	-102	0
$\Delta \Delta T$ [F]	-68	0	68	0	-21	0	48	0
ΔT_w [F]	-71	-40	0	1	-45	-38	0	1
ΔC_{rc} [%]	23.7	82.6	152.8	-12.5	13.5	42.7	119.2	-7.0
ΔC_w [%]	5.6	3.4	4.6	8.9	6.7	2.6	-2.9	10.6
$\Delta Conv.$ [%]	0.8	1.1	1.1	-8.6	1.6	2.7	3.9	-11.3
ΔY_g [%]	0.6	0.9	1.0	-5.3	1.2	2.2	3.3	-7.6
$\Delta Air/Oil$ [%]	-0.2	3.7	7.9	-11.0	1.4	4.6	10.8	-11.9
$\Delta Cat/Oil$ [%]	1.1	24.7	55.1	-22.6	1.0	14.0	46.6	-17.0
30% increase in coking rate								
	Case I				Case II			
Controlled	T_{ris} T_{rpm}	T_{ris} ΔT	T_{ris} T_w	T_{rpm} ΔT	T_{ris} T_{rpm}	T_{ris} ΔT	T_{ris} T_w	T_{rpm} ΔT
ΔT_{ris} [F]	0	0	0	-43	0	0	0	-57
ΔT_{rpm} [F]	0	-45	-79	0	0	-48	-119	0
$\Delta \Delta T$ [F]	-83	0	83	0	-24	0	57	0
ΔT_w [F]	-83	-47	0	6	-51	-46	0	15
ΔC_{rc} [%]	22.0	102.3	195.8	-23.0	12.3	51.3	160.5	-14.6
ΔC_w [%]	5.8	3.4	6.9	10.3	7.0	1.7	-3.1	12
$\Delta Conv.$ [%]	-8.8	-8.3	-8.4	-22.8	-8.7	-6.6	-5.4	-28.6
ΔY_g [%]	-7.0	-6.6	-6.5	-18.0	-7.3	-5.4	-4.2	34.8
$\Delta Air/Oil$ [%]	-4.1	-0.8	6.2	-18.3	-2.7	1.7	8.8	-20.6
$\Delta Cat/Oil$ [%]	-3.4	28.1	68.5	-32.1	-2.8	13.5	56.2	-25.5

With smaller perturbations all control schemes become more acceptable, but the changes assumed are quite reasonable, and they illuminate the robustness of $[T_{rgn}, T_{ris}]$ control. There is one interesting aspect to all cases. Contrary to intuition, the increase in coke on regenerated catalyst is much smaller for the case of $[T_{rgn}, T_{ris}]$ as compared to ΔT or T_{sg} control. As ΔT is proportional to excess oxygen at cyclone inlet, one would have expected that ΔT control is the superior policy to keep C_{rgc} constant. The reason for the increase is the decrease in reactor temperature. The interactions in the FCC are so complex that decoupled arguments can be misleading.

Table 2 demonstrates that the $[T_{rgn}, T_{ris}]$ control is robust and dominant in the sense discussed here. One could get similar results from a linearized analysis, but one would miss the nonlinear stability aspects that are apparent in T_{sg} control.

In Fig. 17 we look at robustness in terms of perturbation in such as coking rate, and catalyst activity over the whole operating region. In Fig. 17 we plot air rate versus T_{rgn} for different levels of coking rate and activity. Fig. 18 gives ΔT and T_{sg} versus T_{rgn} . We note that ΔT in the controllable region changes but the changes are correlated to T_{rgn} . T_{sg} is here more sensitive. The minimum T_{sg} moves up and down with coking rate and if the setpoint is too high an increase in coking rate would make the unit unstable at that T_{sg} .

Fig. 17 also indicates how catalyst activity can be modified to move T_{rgn} in the desired direction. We note that the maximum of T_{rgn} moves strongly when either activity or coking rate change. This has implications when we operate in full combustion or close to the maximum value of T_{rgn} , in partial combustion. Close to the maximum T_{rgn} , the control $[T_{ris}, T_{rgn}]$ is not robust and we have to avoid operating at too high a T_{rgn} . In complete combustion this is less of a problem as we control O_2 concentration at a slight excess. The unit will adjust itself to a new steady state with a lower T_{rgn} .

For a 30% increase in either catalyst activity or coking rate the system remains stable. Air rates at constant $[T_{rgn}, T_{ris}]$ were almost constant for the increase in activity and decreased for an increase in coking rate. But if we decrease activity by 50% the changes are dramatic. As expected $T_{rgn}(\max)$ decreases, but here air rate increases significantly. More importantly, all steady states are open loop unstable. Reducing coking rate has even a stronger effect. There is therefore a range of activity and coking rate for which the control is not robust at this riser

temperature. Increasing riser temperature would allow some stable steady states. But we are here close to the limit where the value of T_{ris} is permissible. A small amount of promoter will make all states for this large perturbation open loop stable, Again we cannot exactly predict this lower limit for catalyst activity, but knowing the form of the topology is useful for bringing back the unit to a stable steady state.

While stability is an overriding concern in an FCC, it is also important to maintain conversion reasonably constant in face of perturbations. Figs. 19 and 20 address this issue plotting conversion and coke on regenerated catalyst respectively versus T_{rgn} and ΔT for the two perturbations. (It makes no sense to plot the unstable case with half catalyst activity.) If we keep T_{rgn} constant conversion will slightly increase for increased catalyst activity and will significantly decrease if coking rate increases. ΔT shows again its multiplicities. In the region where ΔT is an effective control variable, the effects are similar to the case of constant T_{rgn} . On the other hand for coke on regenerated catalyst there is an interesting difference noted better in Table 2. Coke on regenerated catalyst (C_{rgc}) stays fairly constant for both perturbations when T_{rgn} is maintained at a constant level, but increases for both perturbations when ΔT is kept constant. As noted before, this is counter intuitive and again illustrates the complex interactions in this system, especially in closed loops.

In Fig. 21 we replot the three cases for constant excess O_2 control in full CO_2 combustion. For increased activity, conversion stays fairly constant which is surprising as catalyst circulation drops significantly due to higher T_{rgn} . On the other hand while an increase in coking rate also leads to higher T_{rgn} it causes a significant drop in conversion. On the other hand C_{rgc} stays very low for all cases which is expectable, as C_{rgc} has to be low at high temperatures with excess oxygen. As combustion rates at these temperatures are high, it is impossible to get a significant excess oxygen at higher values of C_{rgc} .

The operation in this regime is therefore quite robust. If one wants to adjust T_{rgn} or conversion, one has to use another variable such as preheat.

IV.g Permissible Controller Settings

We pointed out before that in a nonlinear system like an FCC only certain combinations of settings are permissible. Now we have to define what permissible means. Obviously any physically reachable combinations of F_{air} and F_{cat} are permissible. However as we see in Fig. 22 only a limited range will lead to a stable hot steady state. What combination of setting for T_{ris} and T_{rgn} will give a stable and unique steady state? We answered that in Fig. 6 for a given base case looking only at fixed T_{ris} . Answering the question in general is more complex, but we give one plot in Fig. 23. While open loop unstable steady states are permissible and can be stabilized, they are less desirable.

There are however other limitations on permissibility. With present catalyst and operating conditions we are more likely to deal with constraints than inherent instability. Units mostly operate close to maximum air rate. Therefore if for a given steady state $[T_{\text{ris}}, T_{\text{rgn}}]$, the air rate required becomes higher than $F_{\text{air}}(\text{max})$ due to perturbations, there is no operable stable high temperature steady state. Similarly, the permissible steady state cannot exceed wet compressor limits or otherwise the compressor will trip. We need here a default action, or constraint protection, and we have to have a sufficient safety limit. We give the wet gas product (C_4) for the base case together with its perturbation in Fig. 24.

One could plot all these constraints in Fig. 23 but this is strongly unit dependent. As present operation is always in a very narrow region of T_{ris} , plots like Fig. 24 are here more appropriate. Fig. 24 is for an adiabatic unit and is normalized with feed rate. Constraints like F_{air} , F_{cat} and wet gas compressor however are strong function of feed rate and we can move them by either increasing or decreasing feed. Catalyst activity and coking rate are also controllable inputs, and effect these constraints.

IV.h Steady State Control

We mentioned before that the dynamic matrix is part of a larger scheme of steady state control. While we don't have the space left for a detailed discussion of steady state control we will here give a few examples. A more detailed discussion will be given in future parts of the series together with the impact of design on control. We referred in previous sections to several aspects of G_s and the plots of Figs. 6 to 24 give considerable insight on it. Thus Figs. 17 to 24

were used to check robustness to changing inputs. They can equally be used as to how to deal with changes in coking rate and utilize catalyst activity to compensate for other changes. They also show how setpoints can be used to change conversion or move away from constraints. One goal in economic optimization is to find a compromise between throughput and conversion in the face of unit constraints. Fig. 19 and 21 are here useful guides.

In Fig. 6, we gave the impact of riser temperatures on air rate. Figs. 25a and 25b show the impact of riser temperature on conversion and C_{rgc} . We plot these as a function of T_{rgn} for partial combustion. For full combustion these are plotted versus excess oxygen. Again, these plots can be used to see how one can use riser temperature to remove constraints.

Another very important variable is feed preheat. Most units today are equipped with feed preheaters. We give the impact of feed preheat in Fig. 26 to 29. We note that feed preheat strongly reduces air requirements at the expense of conversion. It is therefore a prime tool in optimization strategies.

Today there are catalyst that have low inherent coking rates for the same conversion, though sometimes they have other penalties in Y_p . Again, Fig. 19-21 allow to judge the impact on conversion.

Catalyst coolers have a strong impact when coking rate increases, or when there is a significant amount of Conradson carbon. In steady state we need to burn between 4 to 5% coke to be in adiabatic heat balance. If there is more coke made due to high Conradson carbon, the unit will not have a useful steady state, unless we remove heat by a catalyst cooler. The exact impact again will be evaluated in a future paper.

The discussion of robustness concentrated on steady state control, which is the main subject of this paper. Another aspect of robustness is the impact of changes in operating conditions on the dynamic transfer functions. While dynamic control will be discussed in a future paper, we want to point out here that changes in operating conditions can lead to drastic changes in the transfer functions, which vary depending on the pair chosen. Some examples are given in Table 3.

We note strong changes in the gains and eigenvalues, not only with controller settings but also with changes in coking rates. (similar changes occur with changes in activity). A decrease

in coking rate can make a given setting open loop unstable, and in some cases can lead to unpermissible settings.

When the open loop system becomes unstable one of the eigenvalues changes from negative to positive. Close to the instability the slowest timescale of the response becomes very large but even for smaller changes well in the stable region the timescale can double.

Table 3a: Gains of the Transfer Functions in Parts b and c of this Table

	Case I: $T_{rgn}=1184^{\circ}\text{F}$ $T_{ris}=1000^{\circ}\text{F}$	Case II: $T_{rgn}=1251^{\circ}\text{F}$ $T_{ris}=1000^{\circ}\text{F}$
T_{rgn}/F_{air}	0.73	0.7
T_{ris}/F_{cat}	-0.14	-0.14
$\Delta T/F_{air}$	-14.5	-14.0
T_{sg}/F_{air}	-13.8	-13.3
Coking Rate decrease by 20%		
T_{rgn}/F_{air}	1.1	0.86
T_{ris}/F_{cat}	-0.25	-0.18
Coking Rate decreased by 45%		
T_{rgn}/F_{air}	-1.1	-0.89
T_{ris}/F_{cat}	0.37	-0.31

Table 3b: Transfer Functions for the Base Case

	Case I: $T_{ign}=1184^{\circ}\text{F}$ $T_{ns}=1000^{\circ}\text{F}$	Case II: $T_{ign}=1251^{\circ}\text{F}$ $T_{ns}=1000^{\circ}\text{F}$
$\frac{T_{ign}(s)}{F_{air}(s)}$	$\frac{(s+0.011+i0.0094)(s+0.011-i0.0094)(s+0.012)}{(s+2.3\cdot 10^{-4})(s+0.013)(s+0.01+i0.0088)(s+0.01-i0.0088)}$	$\frac{(s+0.011+i0.0073)(s+0.011-i0.0073)(s+0.0085)}{(s+2.8\cdot 10^{-4})(s+0.009)(s+0.0094+i0.0073)(s+0.0094-i0.0073)}$
$\frac{T_{ns}(s)}{F_{cat}(s)}$	$\frac{(s+0.011+i0.0089)(s+0.011-i0.0089)(s+0.012)(s-3.1\cdot 10^{-4})}{(s+2.3\cdot 10^{-4})(s+0.013)(s+0.01+i0.0088)(s+0.01-i0.0088)}$	$\frac{(s+0.0099+i0.0072)(s+0.0099-i0.0072)(s+0.0084)(s-2.8\cdot 10^{-4})}{(s+2.8\cdot 10^{-4})(s+0.009)(s+0.0094+i0.0073)(s+0.0094-i0.0073)}$
$\frac{\Delta T(s)}{F_{air}(s)}$	$\frac{(s+0.017+i0.0078)(s+0.017-i0.0078)(s+0.012)(s-8.2\cdot 10^{-4})}{(s+2.3\cdot 10^{-4})(s+0.013)(s+0.01+i0.0088)(s+0.01-i0.0088)}$	$\frac{(s+0.032)(s+0.011)(s+0.0089)(s-5.2\cdot 10^{-4})}{(s+2.8\cdot 10^{-4})(s+0.009)(s+0.0094+i0.0073)(s+0.0094-i0.0073)}$
$\frac{T_{sg}(s)}{F_{air}(s)}$	$\frac{(s+0.017+i0.0078)(s+0.017-i0.0078)(s+0.012)(s-7.8\cdot 10^{-4})}{(s+2.3\cdot 10^{-4})(s+0.013)(s+0.01+i0.0088)(s+0.01-i0.0088)}$	$\frac{(s+0.032)(s+0.011)(s+0.0089)(s-4.9\cdot 10^{-4})}{(s+2.8\cdot 10^{-4})(s+0.009)(s+0.0094+i0.0073)(s+0.0094-i0.0073)}$

Table 3c: Transfer Function for Coking Disturbances

	Case I: $T_{rgn}=1184^{\circ}\text{F}$ $T_{ns}=1000^{\circ}\text{F}$	Case II: $T_{rgn}=1251^{\circ}\text{F}$ $T_{ns}=1000^{\circ}\text{F}$
Coking Rate decreased by 20%		
$\frac{T_{rgn}(s)}{F_{air}(s)}$	$\frac{(s+0.011+i0.0097)(s+0.011-i0.0097)(s+0.012)}{(s+1.6\cdot 10^{-4})(s+0.011+i0.0091)(s+0.011-i0.0091)(s+0.013)}$	$\frac{(s+0.012+i0.0069)(s+0.012-i0.0069)(s+0.0087)}{(s+2.2\cdot 10^{-4})(s+0.0093)(s+0.011+i0.007)(s+0.011-i0.007)}$
$\frac{T_{ria}(s)}{F_{cat}(s)}$	$\frac{(s+0.011+i0.0092)(s+0.011-i0.0092)(s+0.012)(s-3.6\cdot 10^{-4})}{(s+1.6\cdot 10^{-4})(s+0.011+i0.0091)(s+0.011-i0.0091)(s+0.013)}$	$\frac{(s+0.011+i0.0089)(s+0.011-i0.0089)(s+0.012)(s-3.1\cdot 10^{-4})}{(s+2.2\cdot 10^{-4})(s+0.0093)(s+0.011+i0.007)(s+0.011-i0.007)}$
Coking Rate decrease by 45%		
$\frac{T_{rgn}(s)}{F_{air}(s)}$	$\frac{(s+0.013+i0.001)(s+0.013-i0.001)(s+0.012)}{(s-1.2\cdot 10^{-4})(s+0.014+i0.0093)(s+0.014-i0.0093)(s+0.013)}$	$\frac{(s+0.0081+i0.009)(s+0.0081-i0.009)(s+0.009)}{(s-5.4\cdot 10^{-4})(s+0.0095)(s+0.025)(s+0.012)}$
$\frac{T_{ria}(s)}{F_{cat}(s)}$	$\frac{(s+0.014+i0.0094)(s+0.014-i0.0094)(s+0.012)(s-4\cdot 10^{-4})}{(s-1.2\cdot 10^{-4})(s+0.014+i0.0093)(s+0.014-i0.0093)(s+0.013)}$	$\frac{(s+0.025)(s+0.013)(s+0.009)(s+1.2\cdot 10^{-4})}{(s-5.4\cdot 10^{-4})(s+0.0095)(s+0.025)(s+0.012)}$

V. SUMMARY AND DISCUSSIONS

Our goals in this paper were twofold:

1. To investigate the control of fluid catalyst crackers
2. To demonstrate and develop the principles of partial control using FCC as an example.

The paper is part of a series and due to the complexity of the problem we were only able to deal with part of the problem. We focused on the required properties of the primary square dynamic control matrix and its relation to the overall steady state control. We showed that despite its complexity a 2x2 matrix is in most cases sufficient for dynamic control and stabilization. But the choice of the variable entering this matrix is crucial.

Most papers dealing with linear multivariable control deal with how to design the algorithm for a given matrix and how to pair the variables. There has been much less emphasis as to how the choice of the matrix effects the overall steady state control which is nonlinear. Our paper demonstrates an approach to identify and test suitable choices. FCC is here a very useful problem to study these effects. Not only is it an important process but it also has very complex interactions and is a highly nonlinear system. Our results can be summarized as follows:

1. In an FCC the choice of the variables in the dynamic matrix cannot be based solely on linear control theory. Additional and primary considerations requiring a nonlinear model are:
 - a. Primary consideration of the choice of the control matrix should be given to evaluate the dominance and sufficiency of the matrix chosen.

Sufficiency is defined as the ability to maintain system stability, and the controlled output Y_p within acceptable limits. The definition therefore depends on model uncertainty and the magnitude of foreseeable disturbances.

The ability to maintain a large set of variables within bounds by a much smaller set of manipulated variables requires that the variables chosen for the primary control matrix have an impact on the whole vector Y_p . Dominance can be evaluated by a linearized model. However, nonlinear models are required to check stability, as in nonlinear systems with multiple steady states we face both linear and nonlinear stability problems

- b. Mappability into Y_p the vector of process specifications and constraints. The setpoints of the variables chosen must be suitable manipulated variables for the overall steady state control

scheme. One must be able to predict their impact based on laboratory measurements. For FCC control only T_{rs} and T_{rp} have this property. ΔT and T_{rg} are strongly sensitive to unit design, and nonidealities and hard to predict from laboratory experiments.

c. The relative gain matrix can change with changes in operating conditions. To avoid frequent retuning the linearized model has to be related to operating condition via a nonlinear model. The control matrix itself may also require change with operating conditions.

d. Input multiplicities

2. Linearized theory is suitable and sufficient for controller tuning and design of algorithm at a given steady state. It is useful for evaluating local stability. The linearized transfer function changes strongly with operating conditions, not only the magnitude of the gains change but also the sign.
3. For evaluating stability considerations and for steady state control and optimization nonlinear models are essential. Linearized theory can be useful for evaluating the stability of a steady state and for stabilizing it. However, the FCC has nonlinear instabilities, and operating close to a bifurcation point is undesirable.
4. While a 2x2 matrix properly chosen is sufficient for dynamic control and stabilization, additional variables available should enter the overall control scheme in a slow mode. Thus catalyst activity is an important and often neglected variable in optimization. So is the composition of the feed, which can be modified. Thus for example when dealing with wind down feed rate may have to be temporarily (or permanently) reduced to remove constraints on the manipulated variables used.
5. When dealing with the impact of disturbances on the process while designing a control circuit emphasis should be given to slow disturbances and changes in inputs. It is also important to minimize changes not just in the variables entering the matrix but also in the crucial variables in Y_p that are normally far more important for the specifications than measured variables chosen for the matrix. This can often be analyzed within a suitable linear model.
6. Fast response, while desirable, is not a main criteria in choosing a dynamic control matrix unless the fast response relates to all crucial variables in Y_p . Fast response can be helped by intermediate control circuits. The FCC is designed to ensure a response slow enough to protect the system from wind down.

NOMENCLATURE

C	- Measurement matrix	
C_{regc}	- Coke on regenerated catalyst	[wt%]
F_{air}	- Air flow rate into the regenerator	[lb air/lb feed]
F_{cat}	- Catalyst circulation rate	[lb cat/lb feed]
G	- Control algorithm matrix	
G_d	- Dynamic control matrix	
G_s	- Steady state control matrix	
M	- Simplified model of M^* by identification and modeling	
M[*]	- Non-linear model of a plant	
N	- Vector of unknown inputs, not measured (disturbances)	
O₂	- Oxygen concentration in the stack gas	[wt%]
T_f	- Feed preheat temperature	[F]
T_{regn}	- Regenerator dense bed temperature	[F]
T_{ris}	- Riser top temperature	[F]
T_{sg}	- Stack gas temperature	[F]
U	- vector of manipulated variables	
U_d	- Vector of manipulated inputs in fast monitoring	
U_s	- Vector of manipulated inputs in slow monitoring	
W	- Vector of measurable inputs, not manipulated	
X	- Vector of state variables	
Y	- Subset of X, measured state variables	
Y_{cd}	- Vector of dynamic control variables	
Y_d	- Vector of process variables in fast monitoring	
Y_p	- Vector of product and process specs and conditions	
Y_s	- Vector of process variables in slow monitoring	
ΔT	- Temperature rise across the cyclones	[F]

Superscript s - variable at steady state

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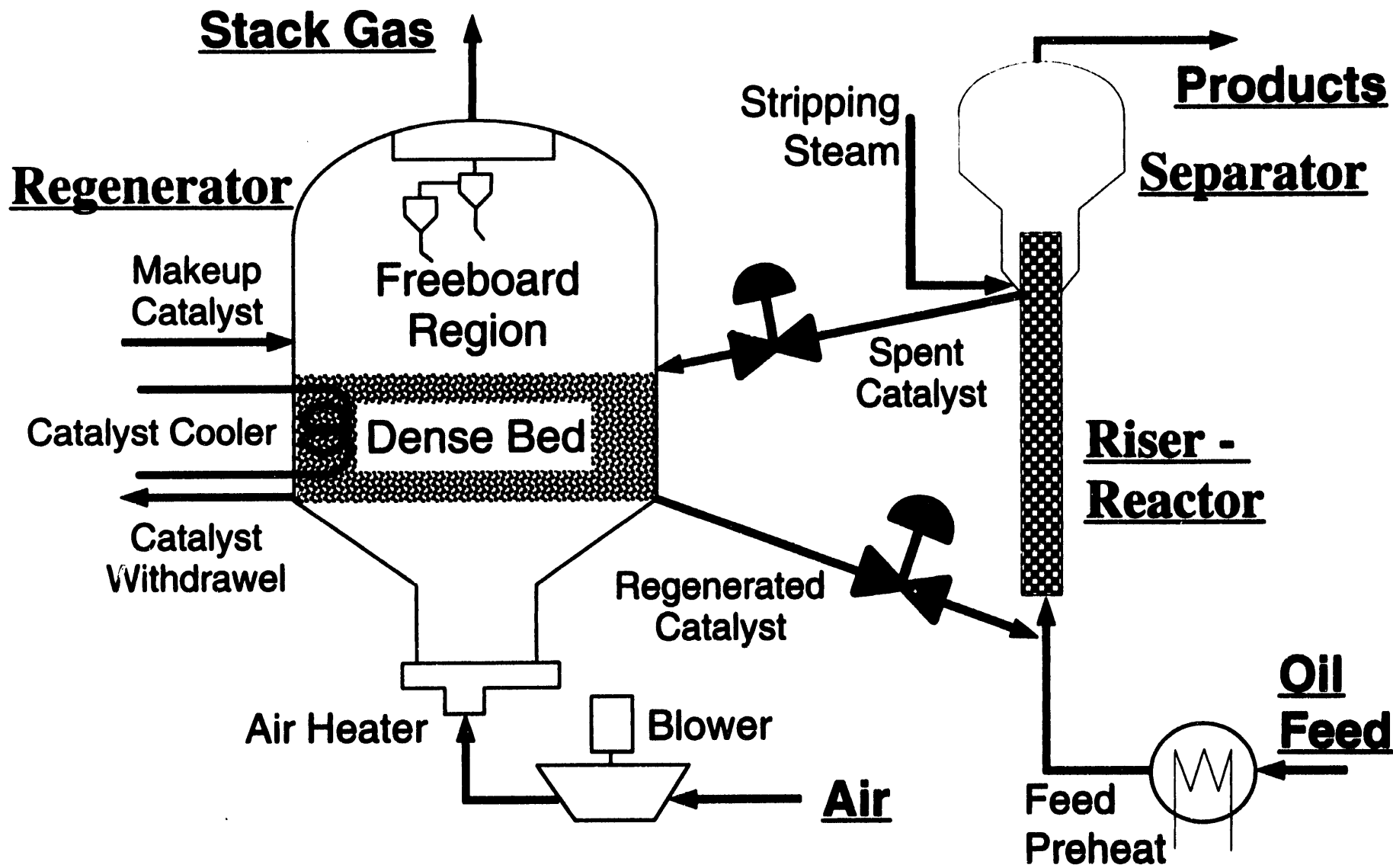


Figure 1: Schematic Diagram of Modern FCC

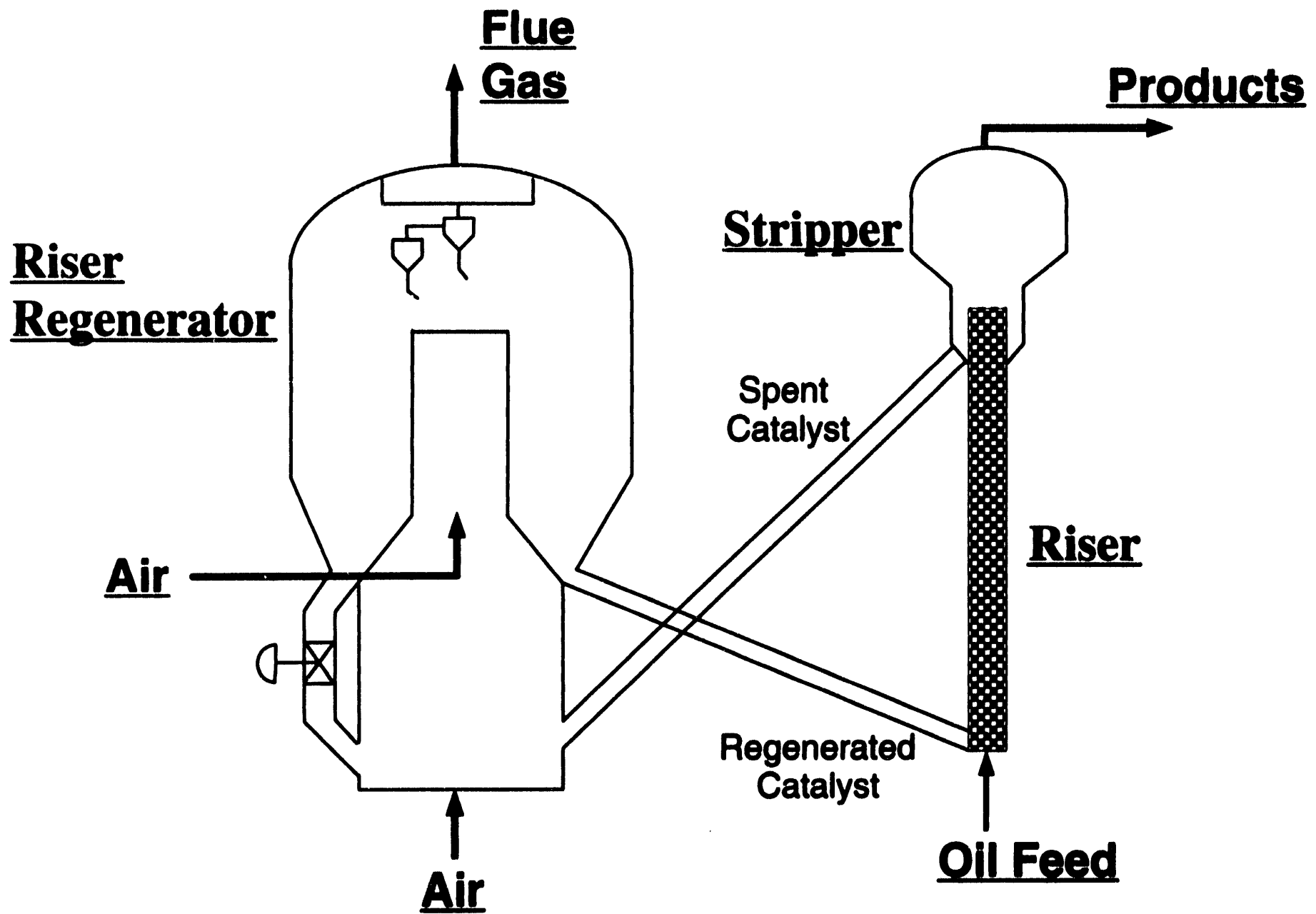


Figure 2: Schematic Diagram of High-Efficiency FCC with Riser-Reactor and Riser-Regenerator

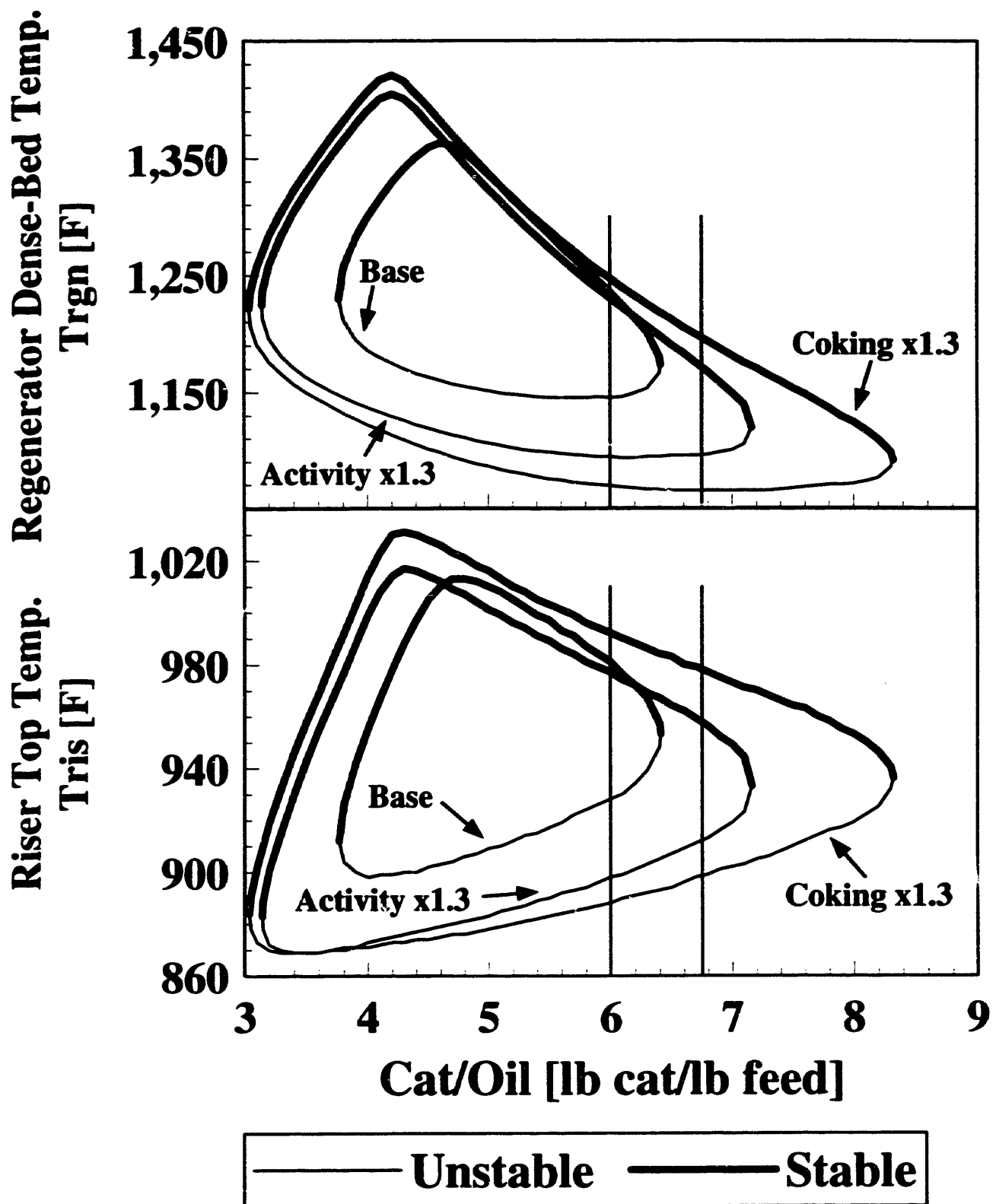


Figure 3: Effect of Activity and Coking Rate on Steady States at Constant Air/Oil=0.55

Manipulated

Controlled

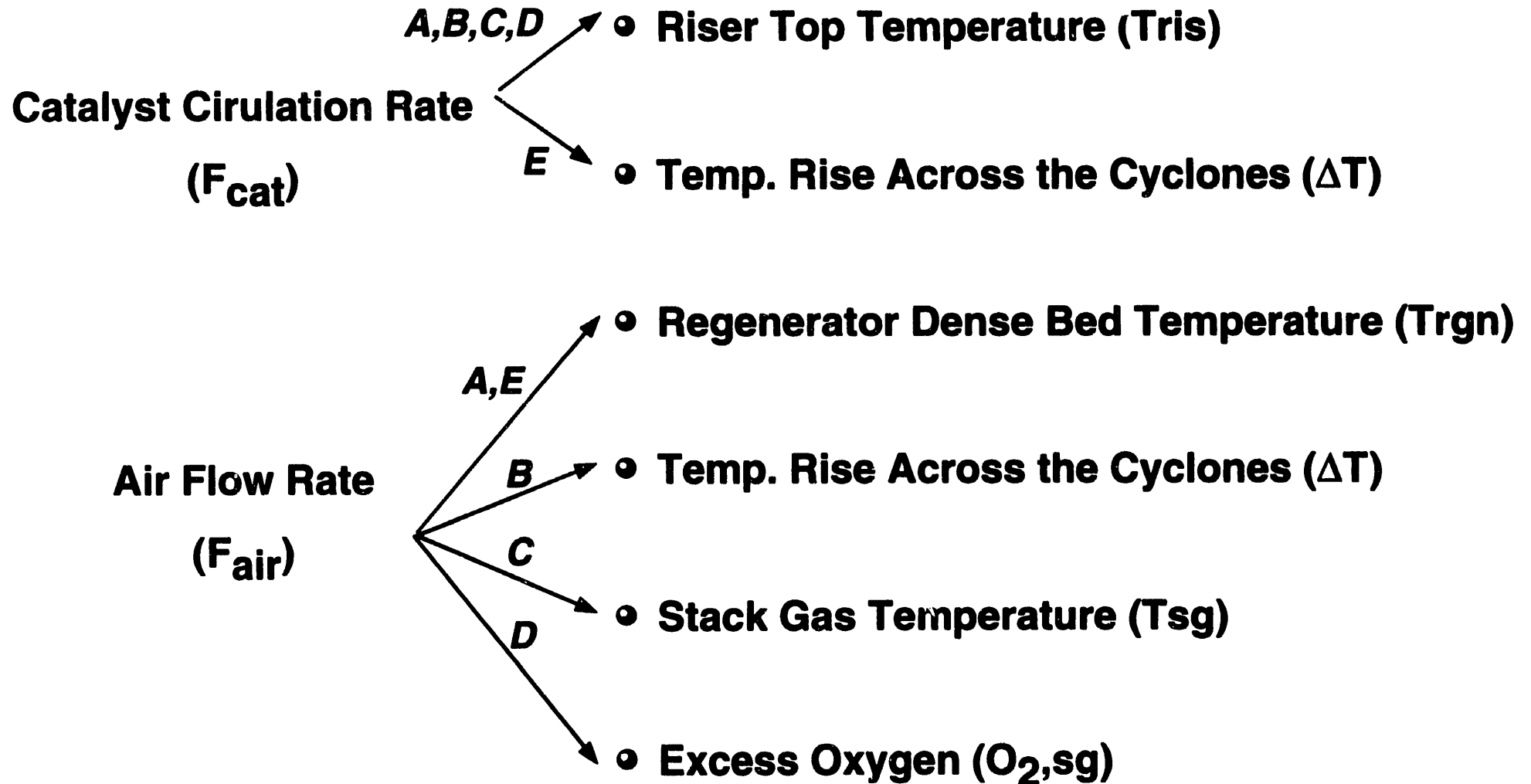
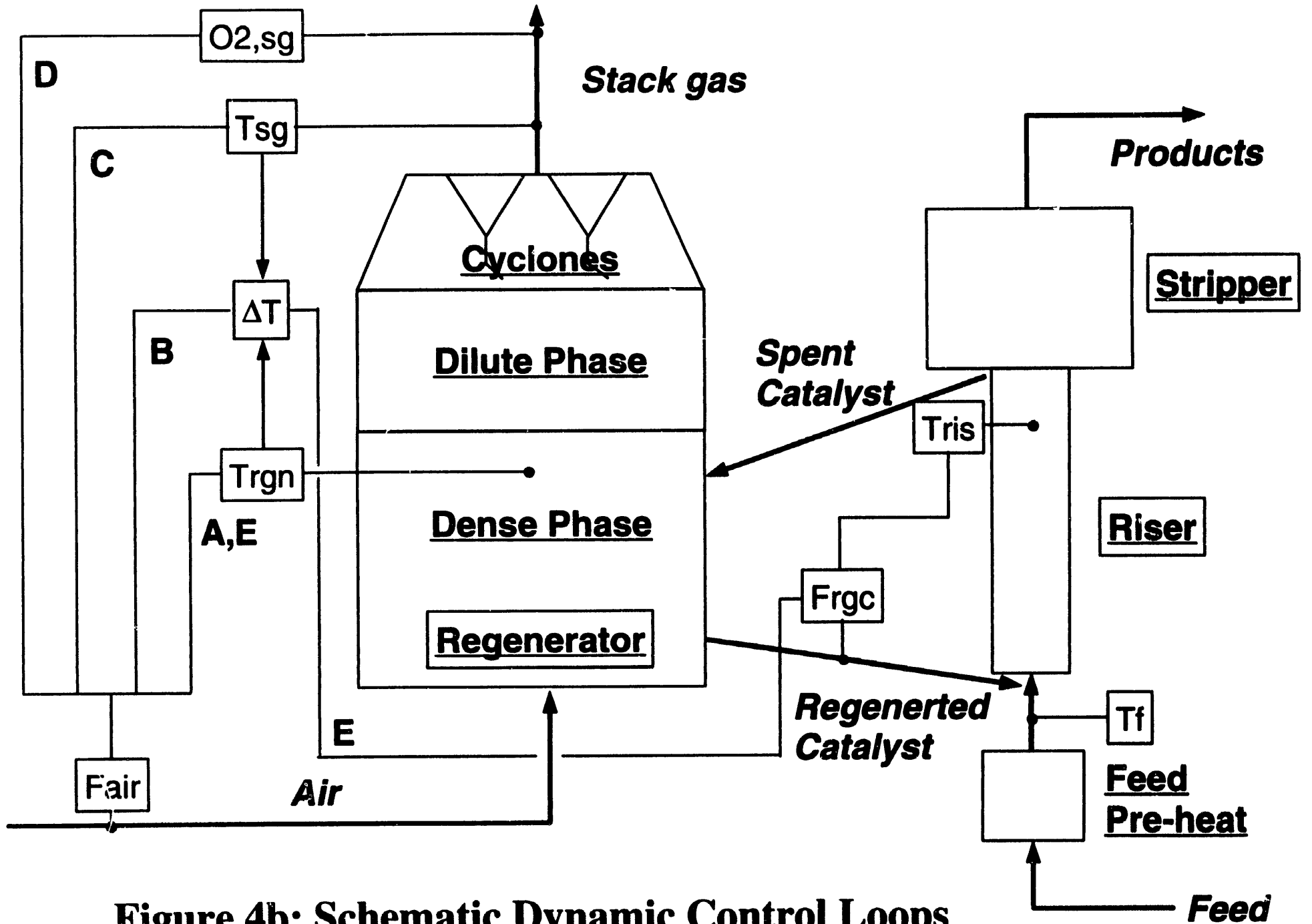


Figure 4a: Primary Dynamic Control - Cases A,B,C,D,E



**Figure 4b: Schematic Dynamic Control Loops
Cases A,B,C,D,E**

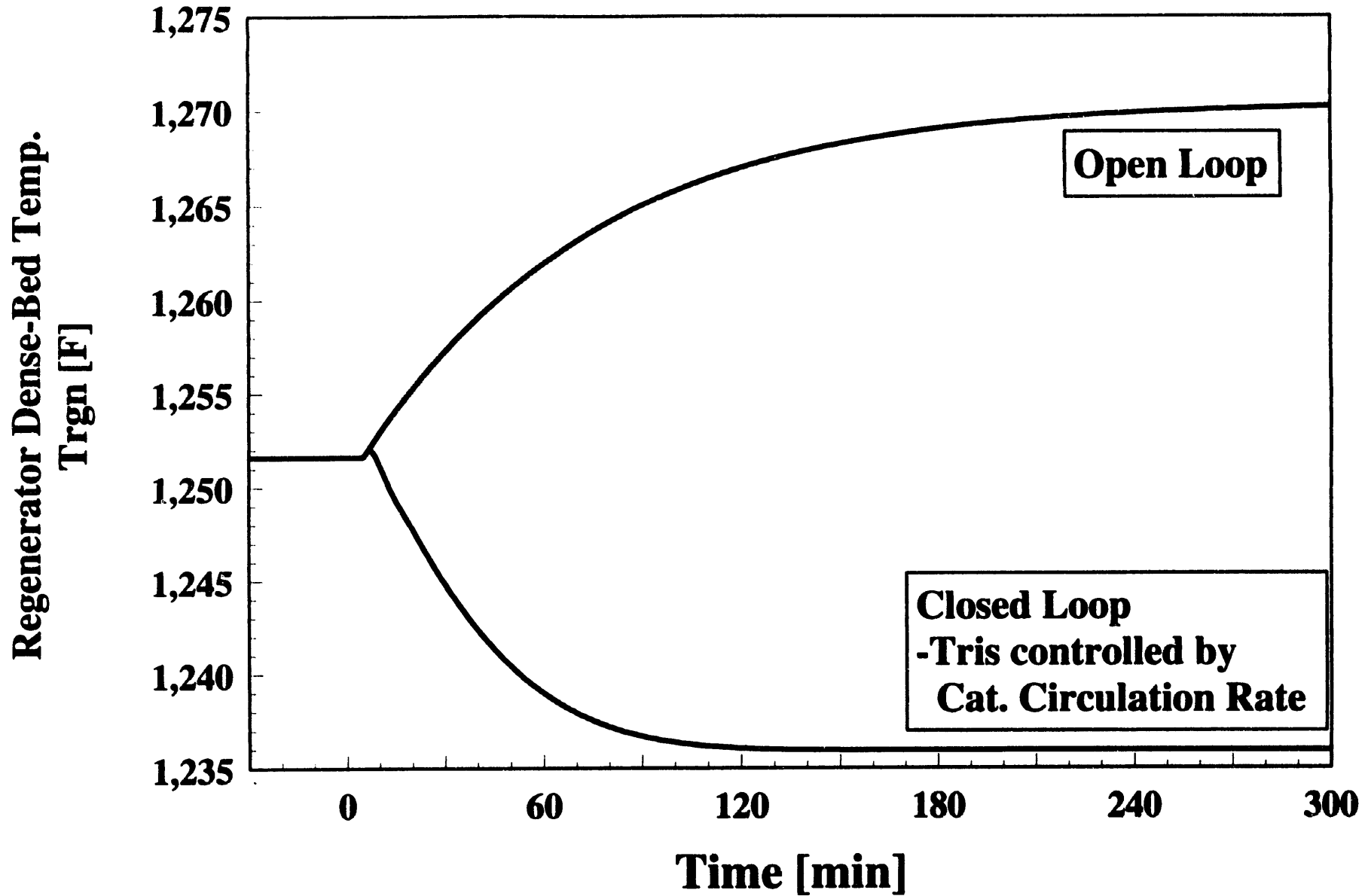


Figure 5: Time Responce of Trgn to 2% Step Change in Air Flow Rate, Open and Closed Loop.

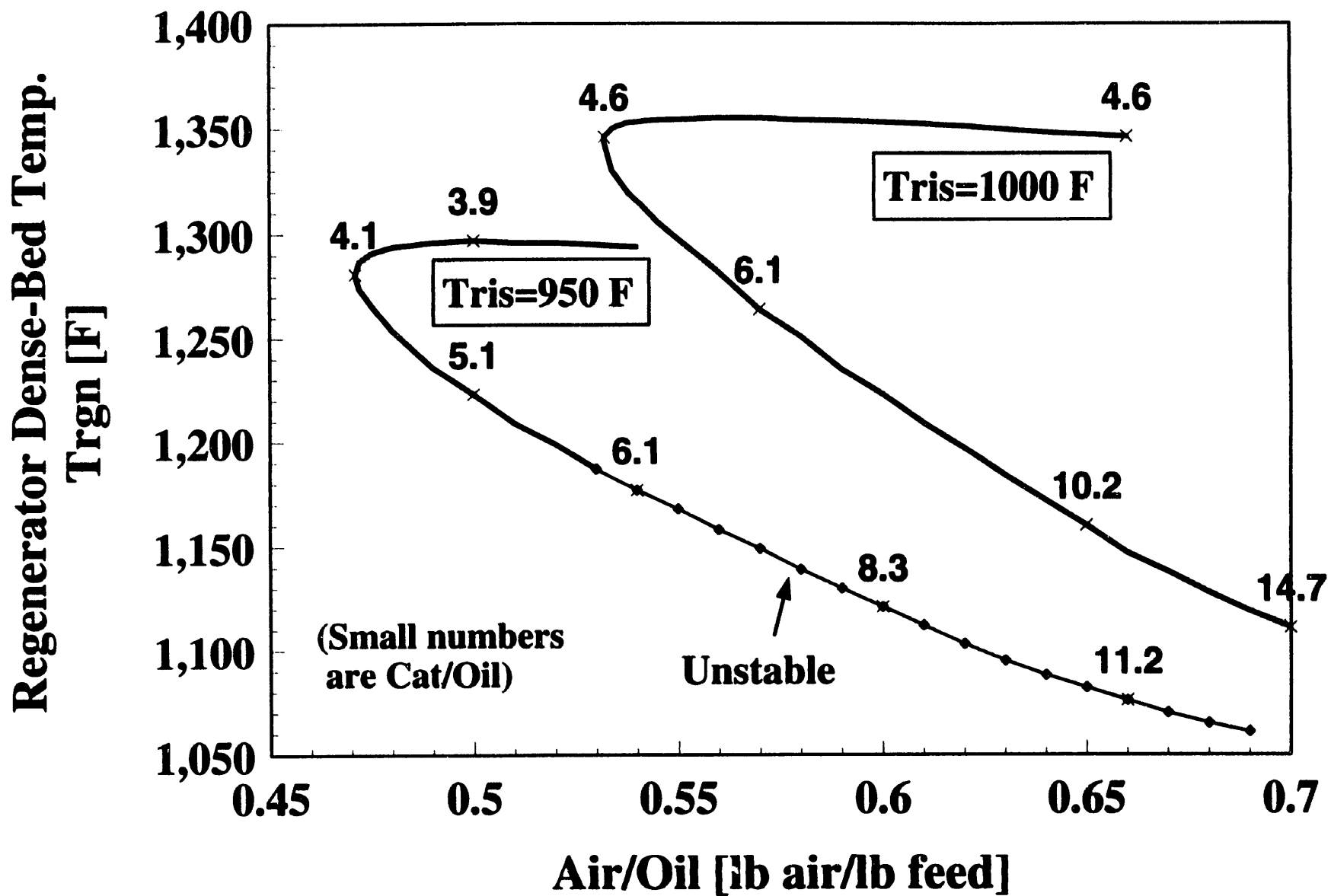
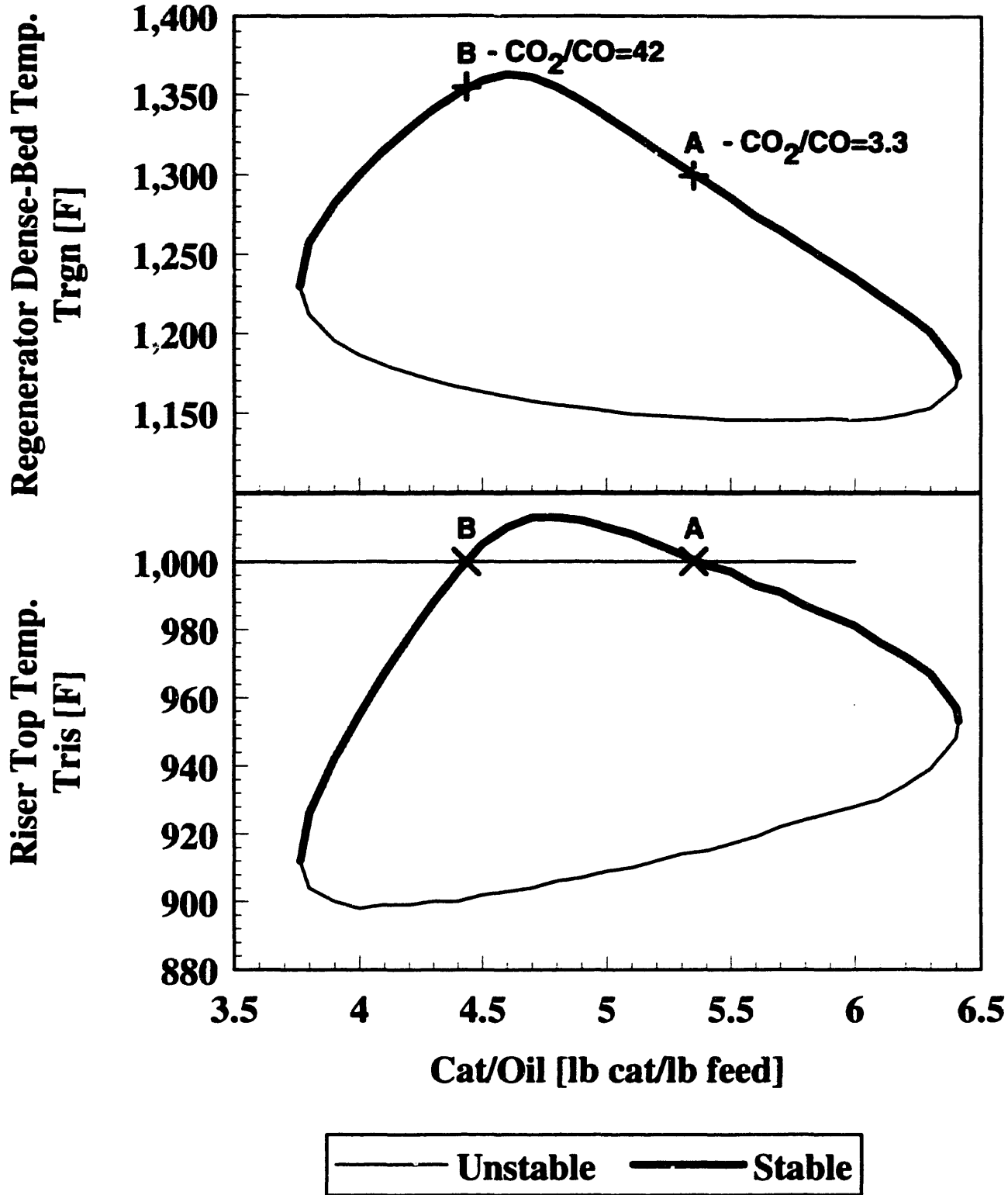


Figure 6: Effect of Riser Top Temp. on Steady States



**Figure 7: Multiple Steady States
at Constant Air/Oil=0.55**

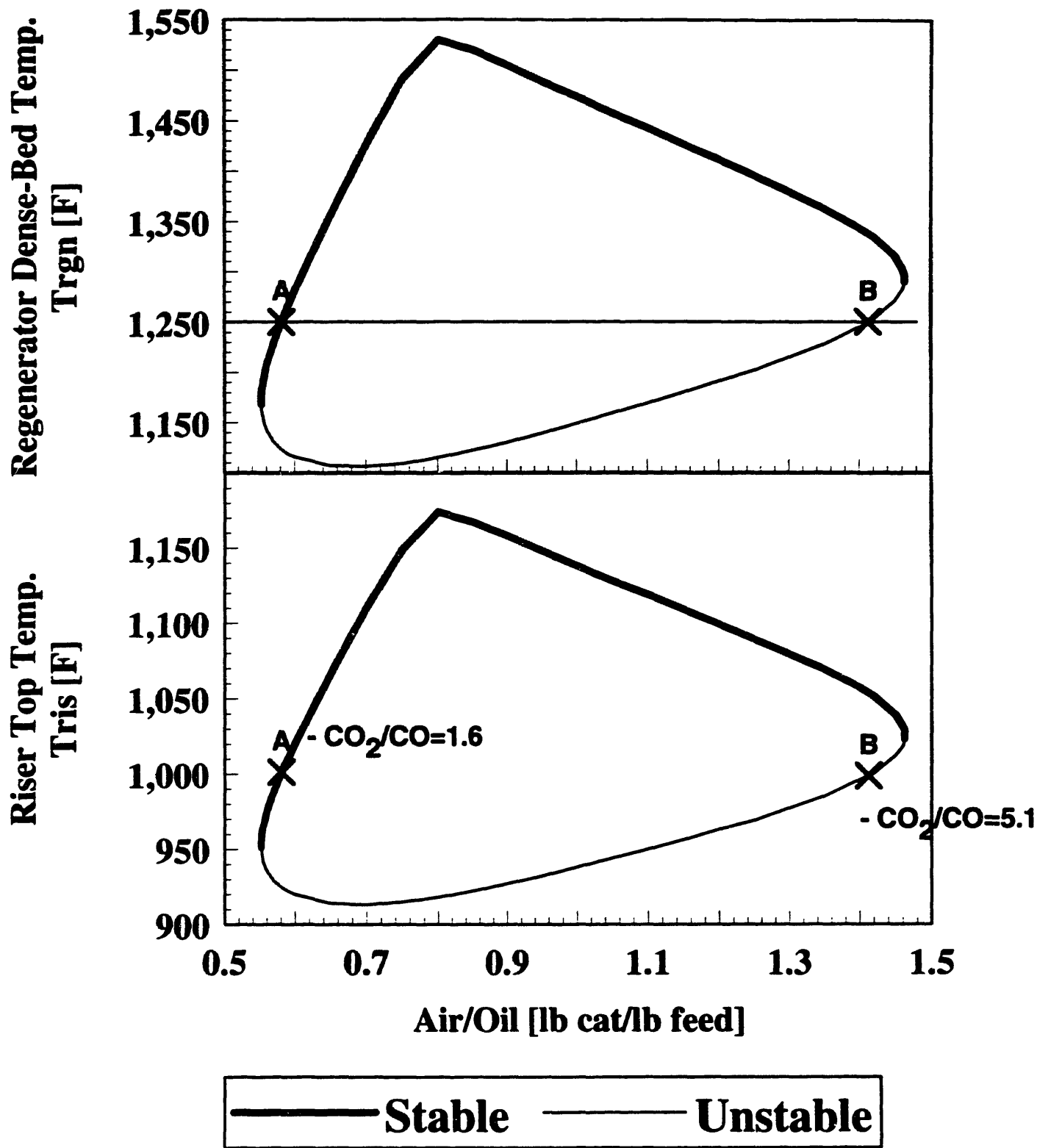


Figure 8: Multiple Steady States at Constant Cat/Oil=6.5

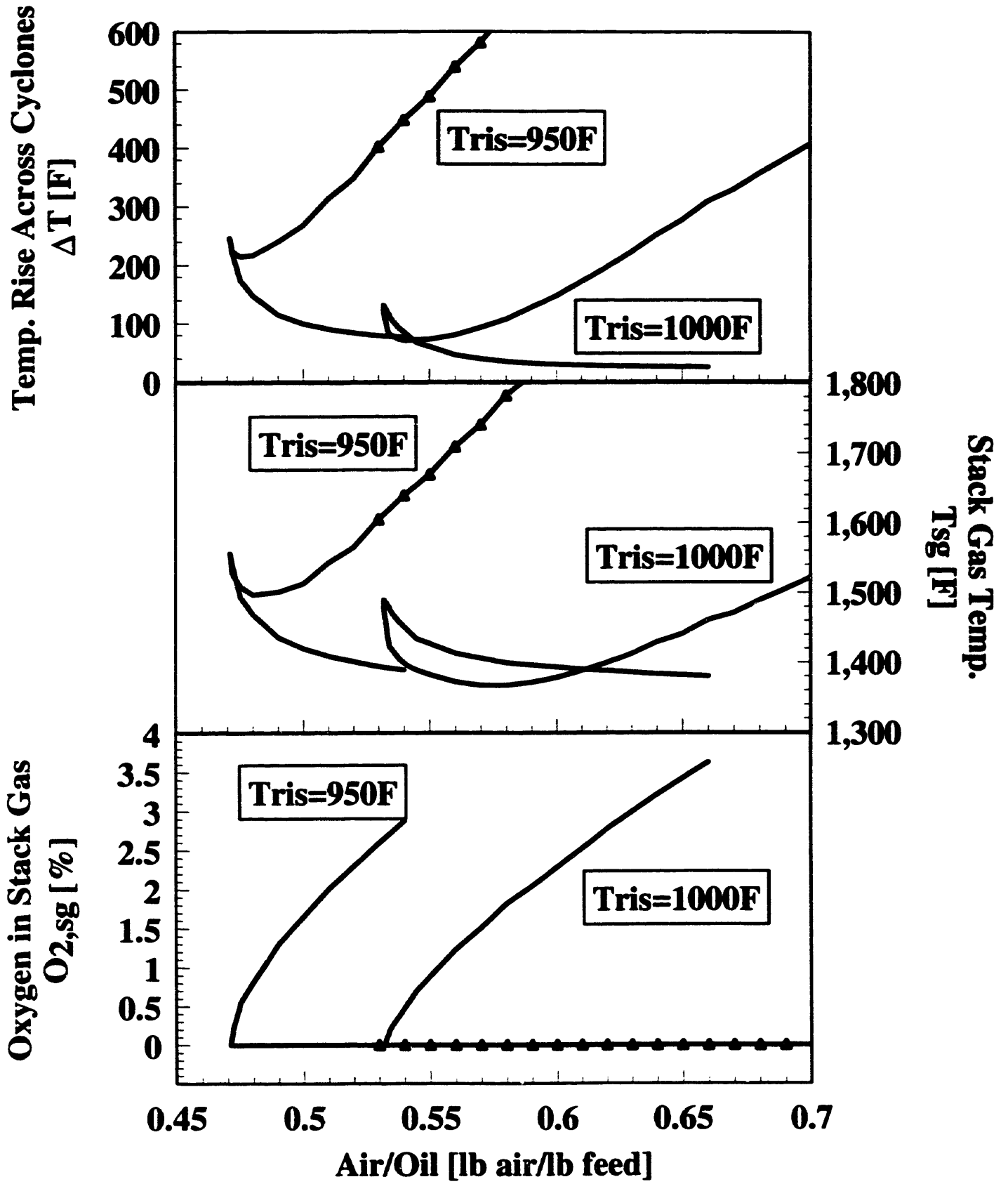


Figure 9: Effect of Riser Top Temp. on Steady States

—•— Unstable — Stable

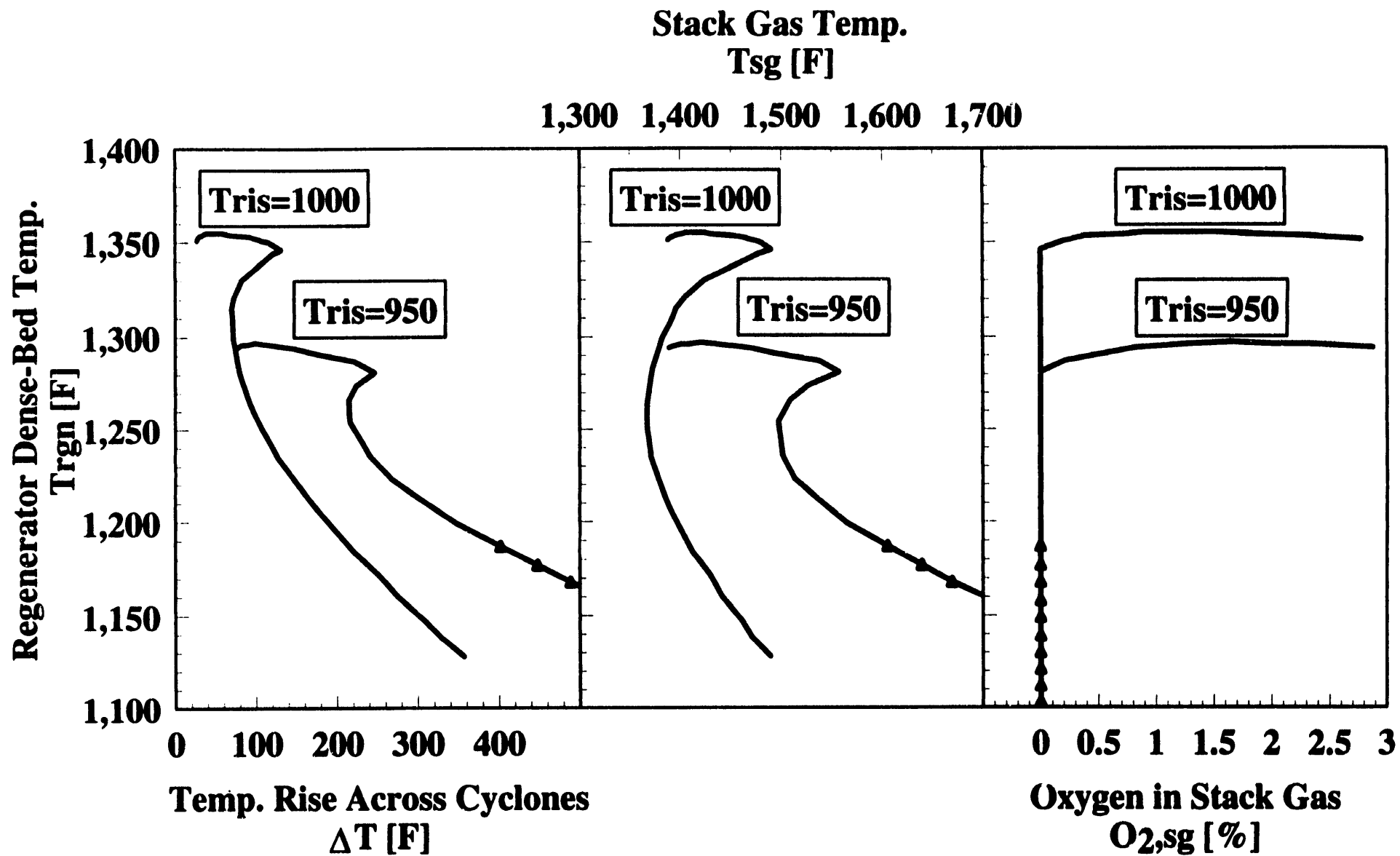


Figure 10: Steady States at Constant Riser Top Temp.

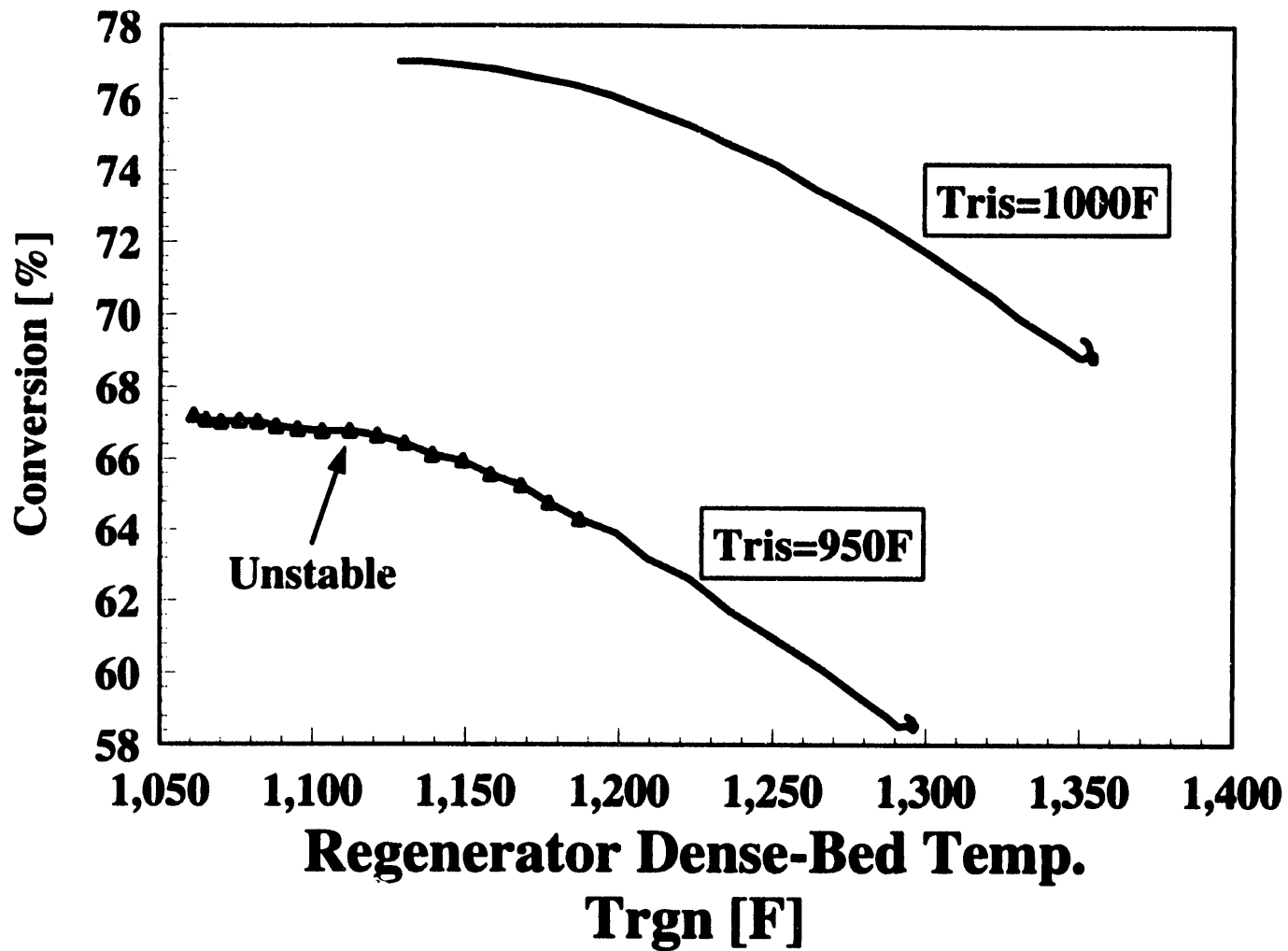


Figure 11: Effect of Riser Top Temp. on Conversion

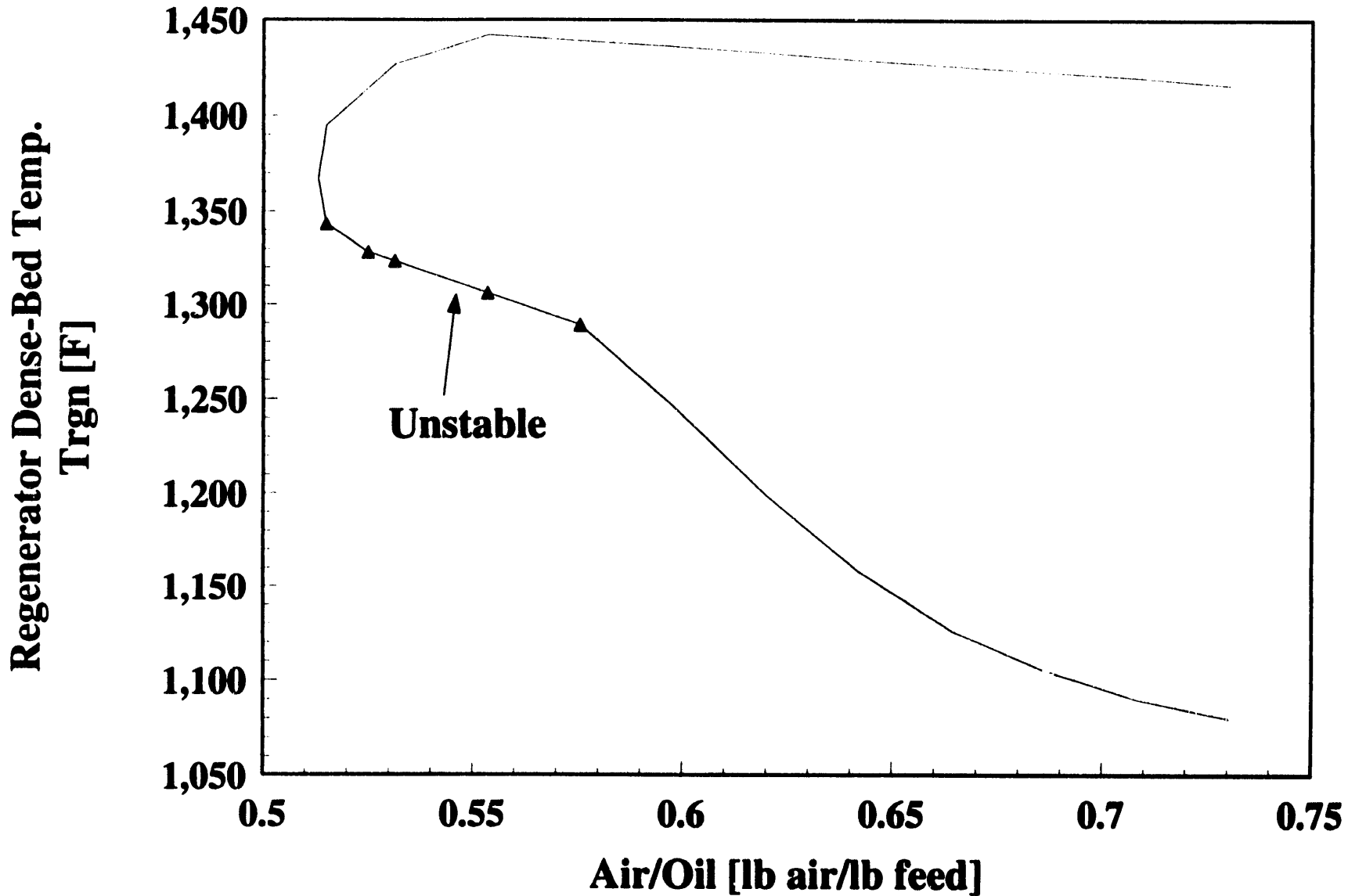
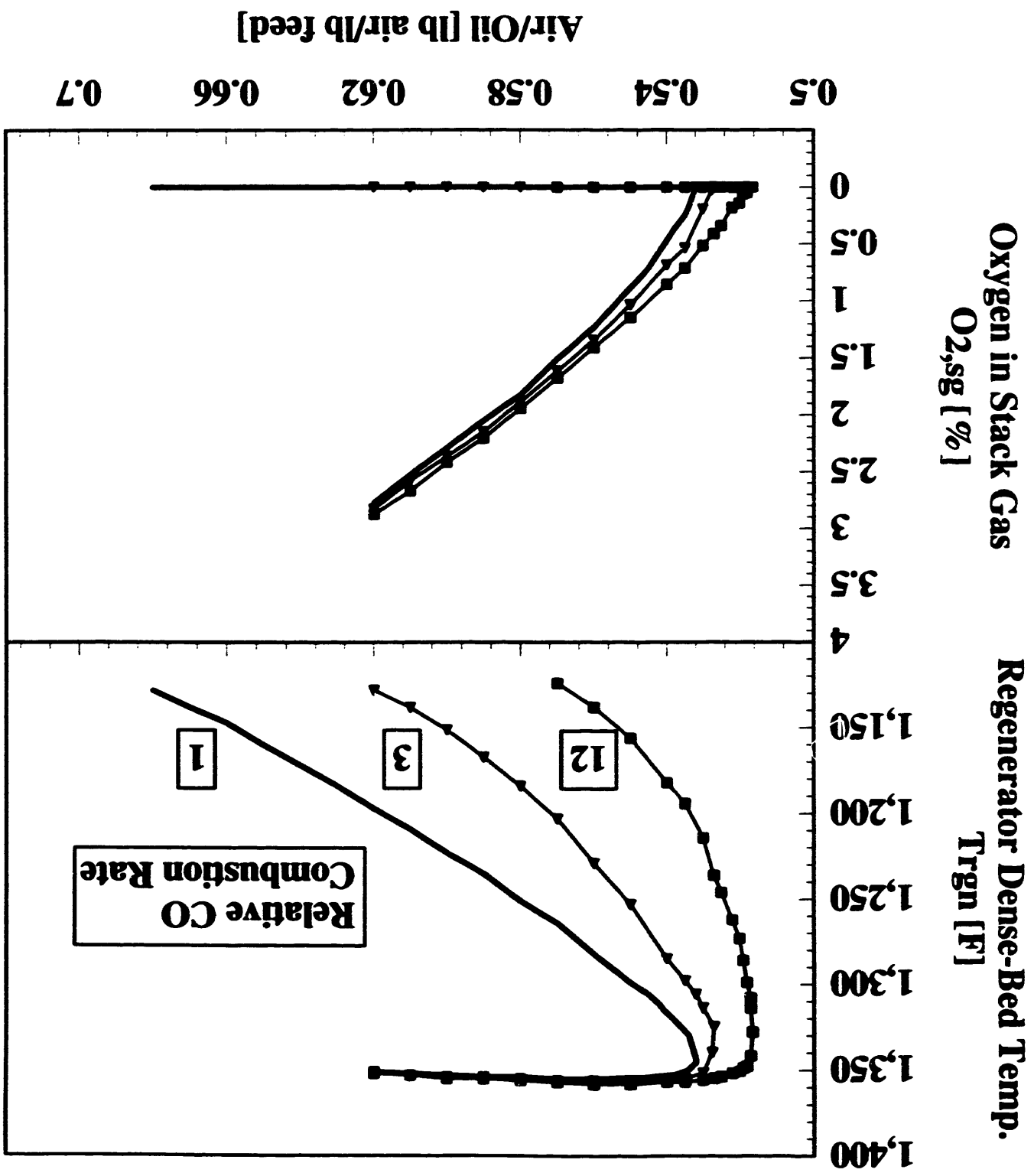
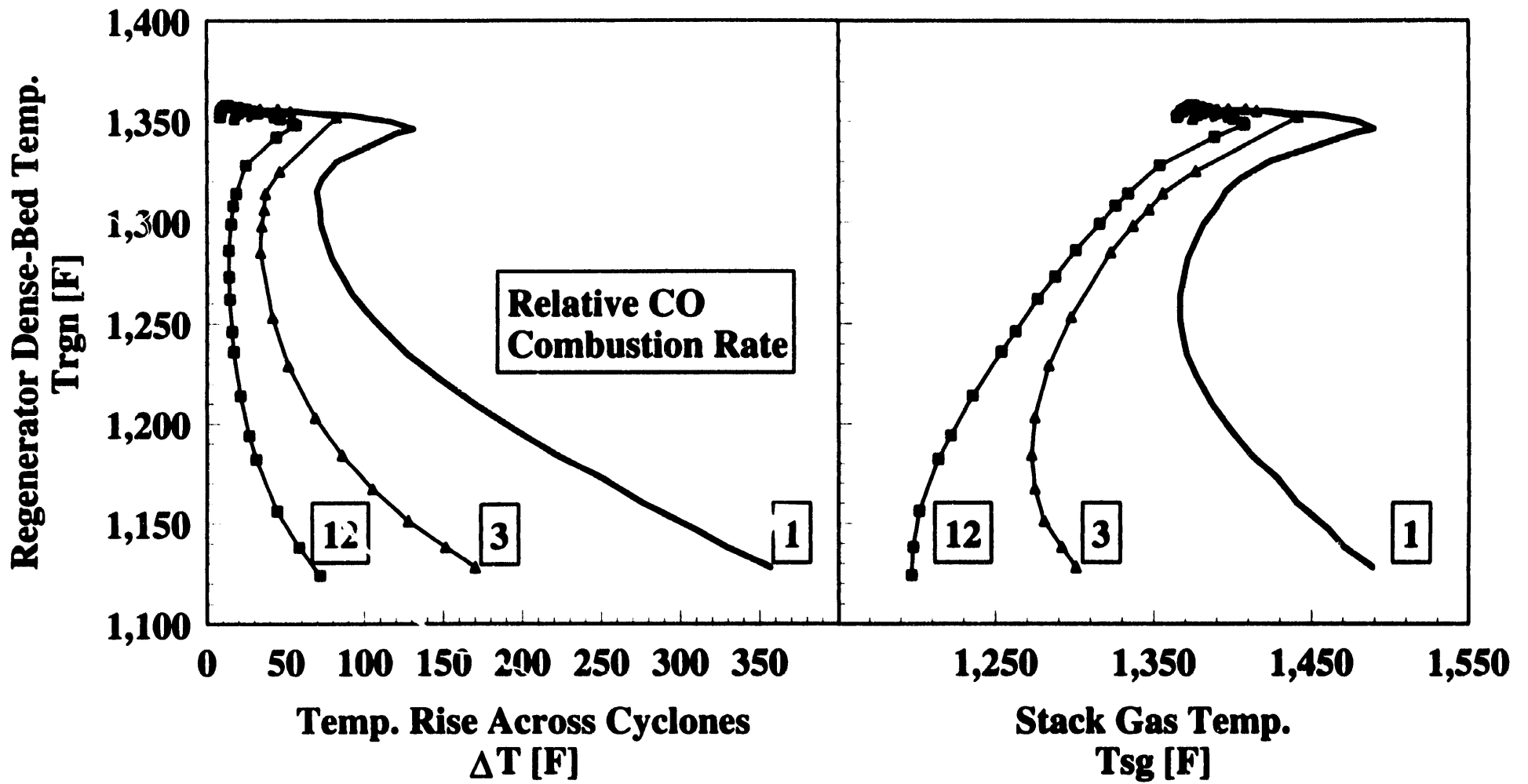


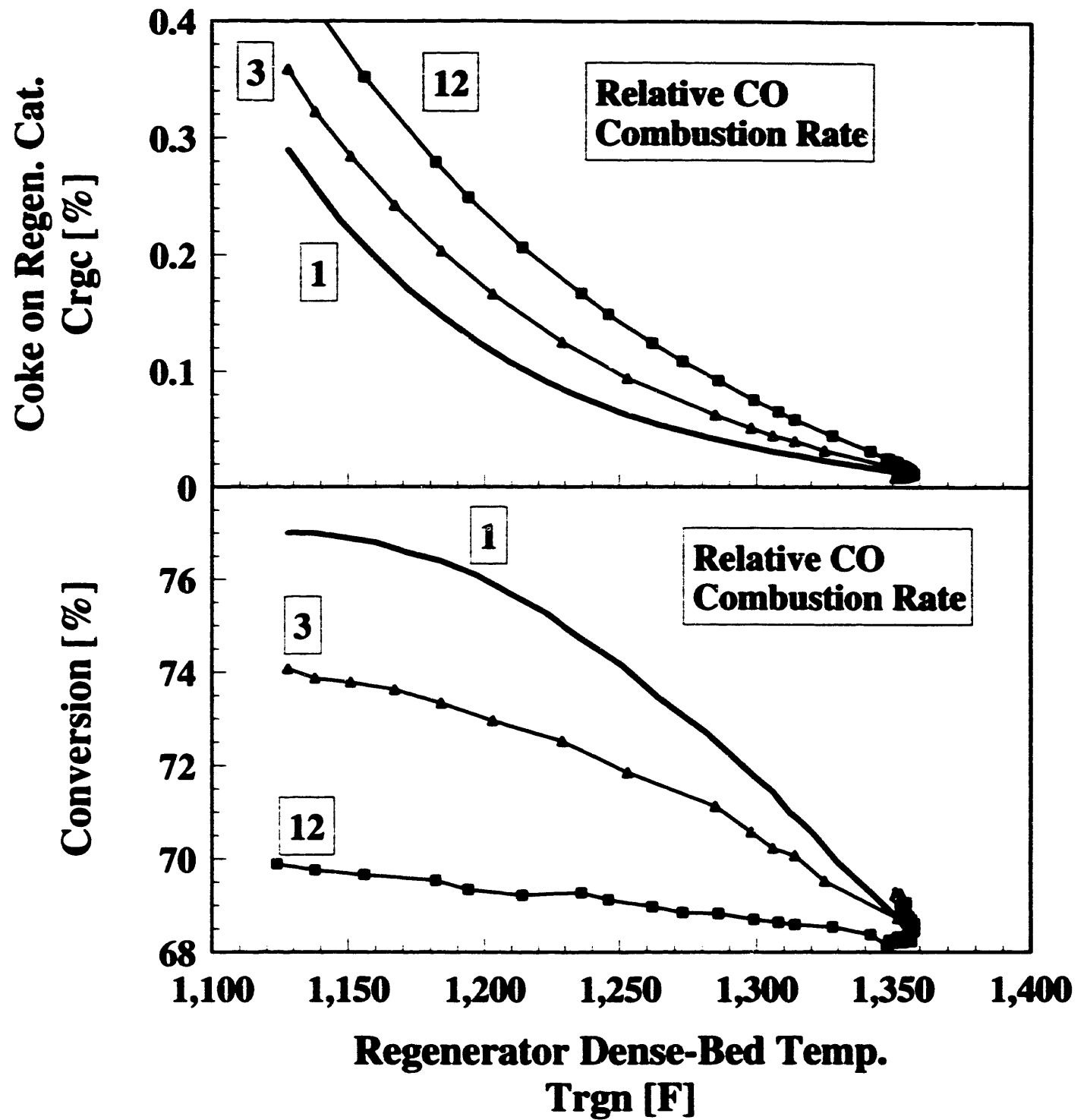
Figure 12: Schematic Figure of Possible Unsteady States at Constant Riser Top Temp.=1000 F

Figure 13: Effect of Combustion Promoter on Steady States. Riser Top Temp.=1000 F





**Figure 14: Effect of Combustion Promoter on Steady States.
at Top Riser Temp.=1000**



**Figure 15: Effect of Combustion promoter
Riser Top Temp.=1000F**

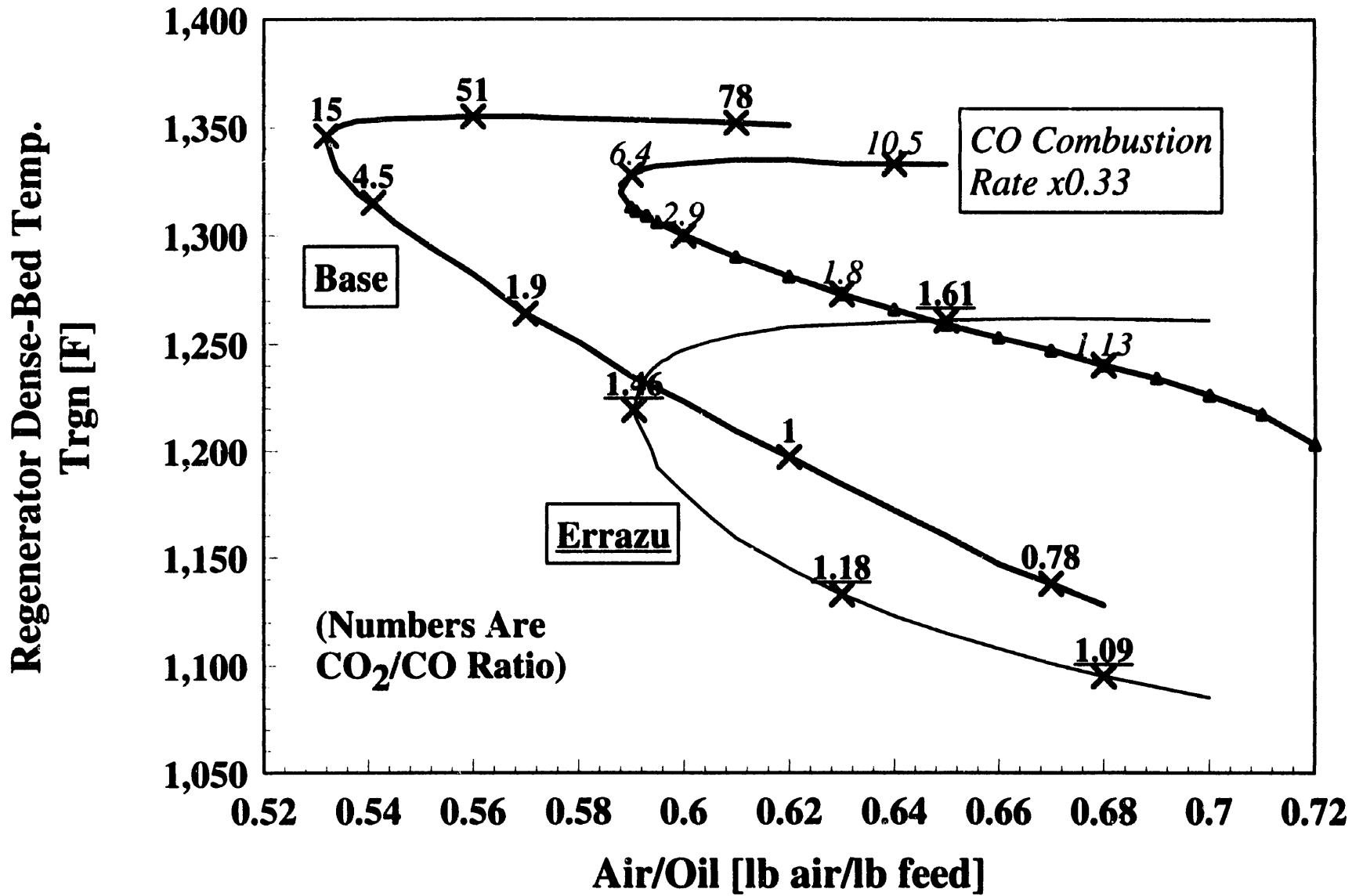


Figure 16: Model Sensitivity to CO to CO₂ Combustion Kinetics
Riser Top Temp.=1000 F

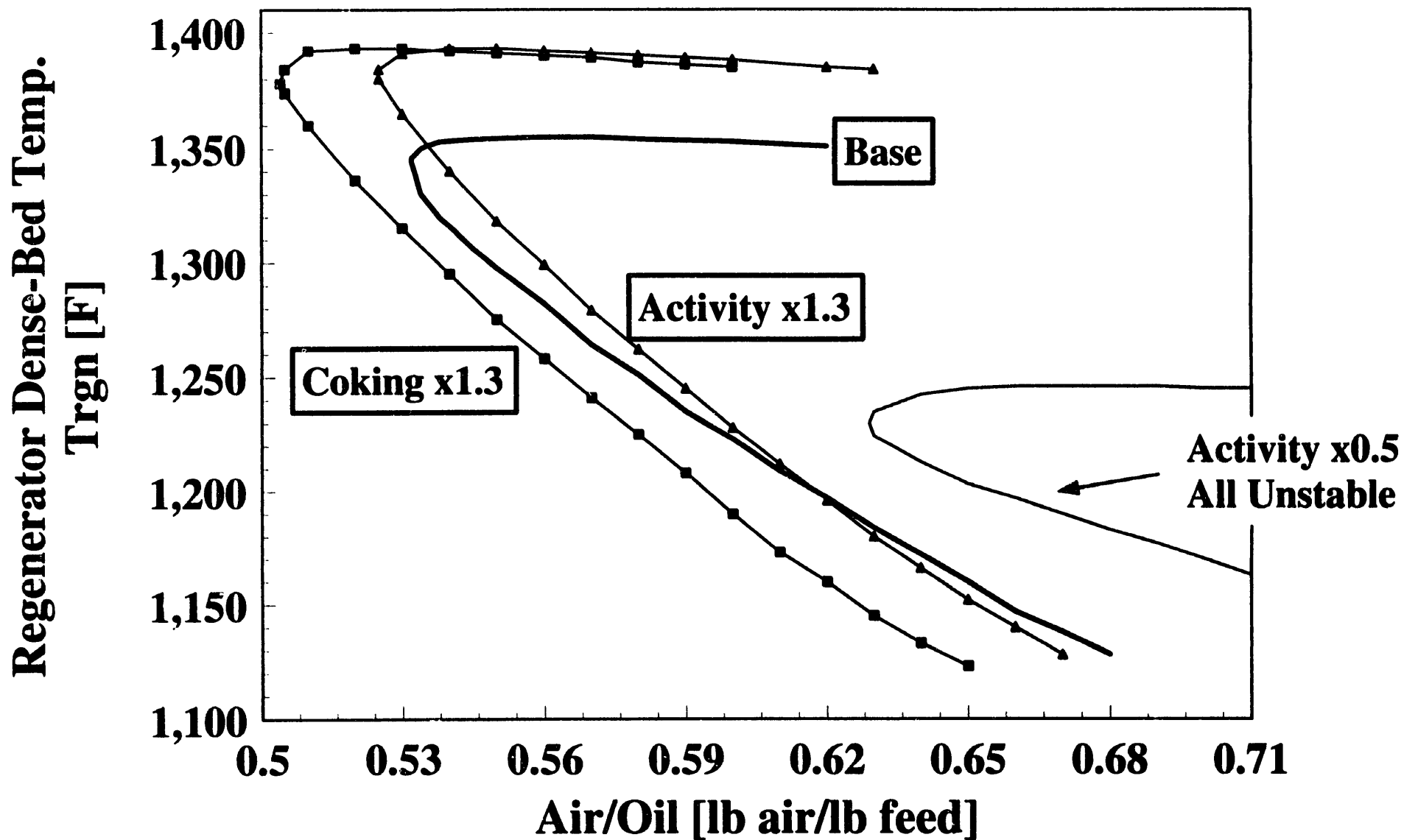
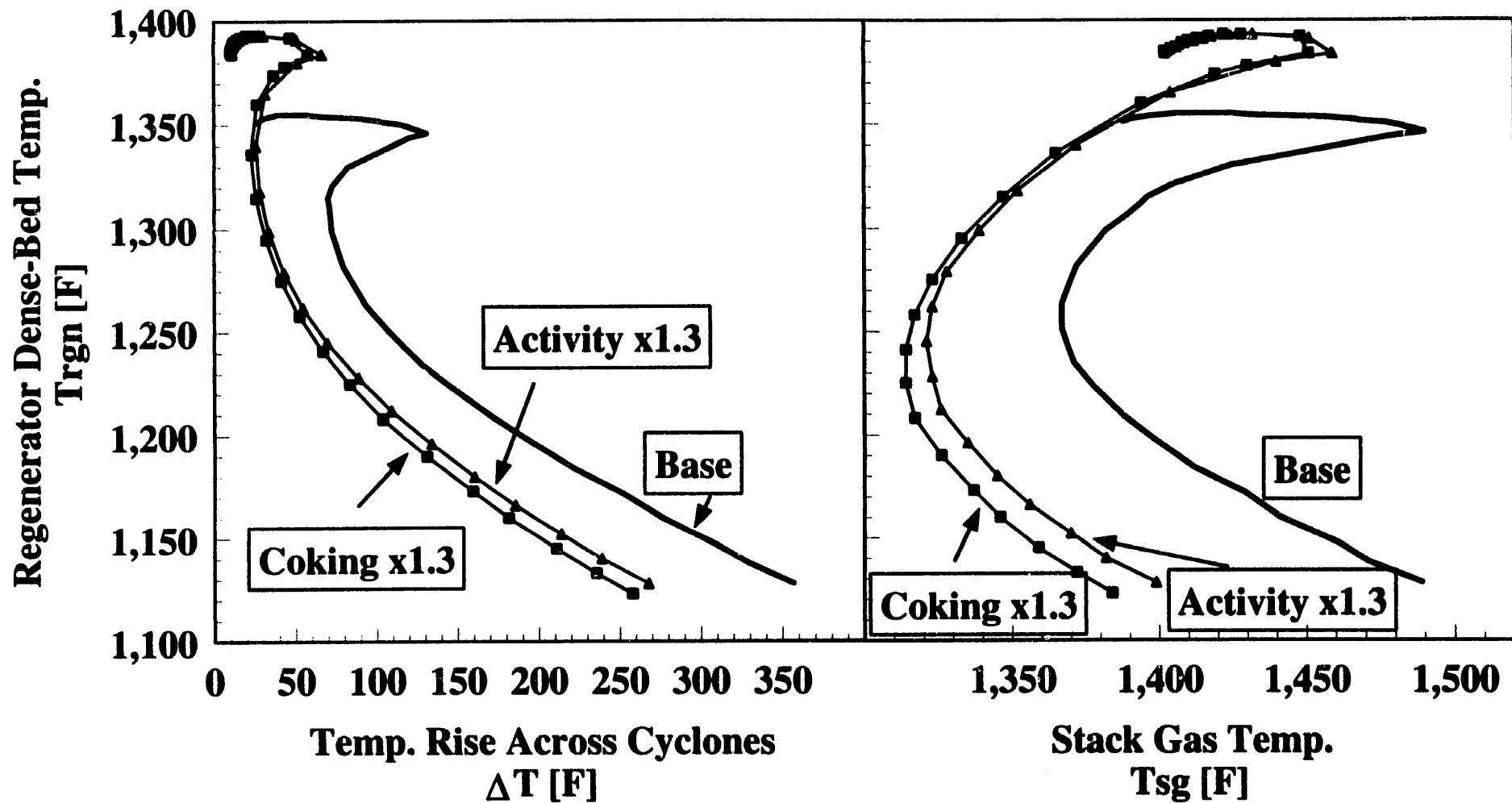


Figure 17: Effect of Activity and Coking Rate on Steady States. Riser Top Temp.=1000 F



**Figure 18: Effect of Activity and Coking Rate on Steady States.
Riser Top Temp.=1000**

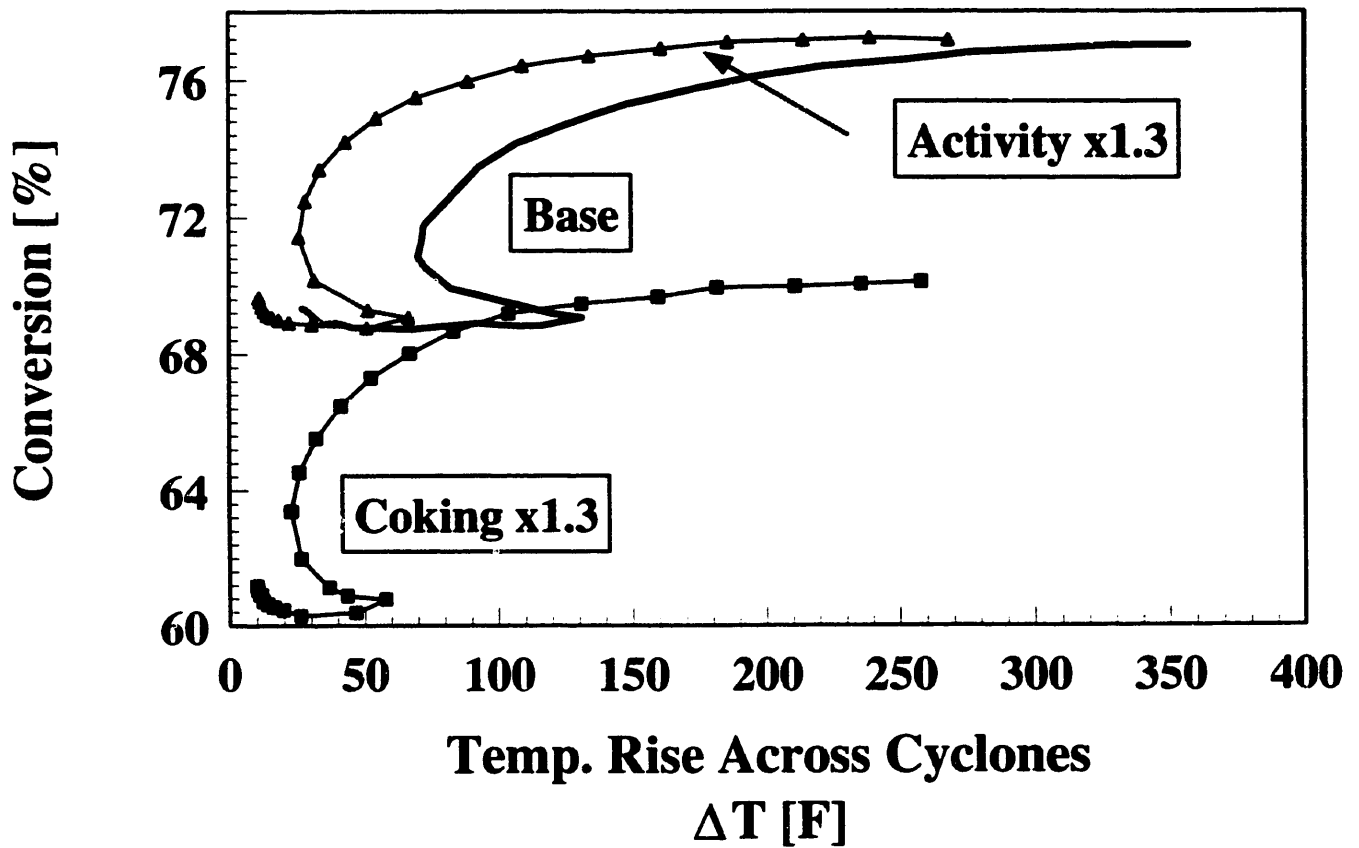
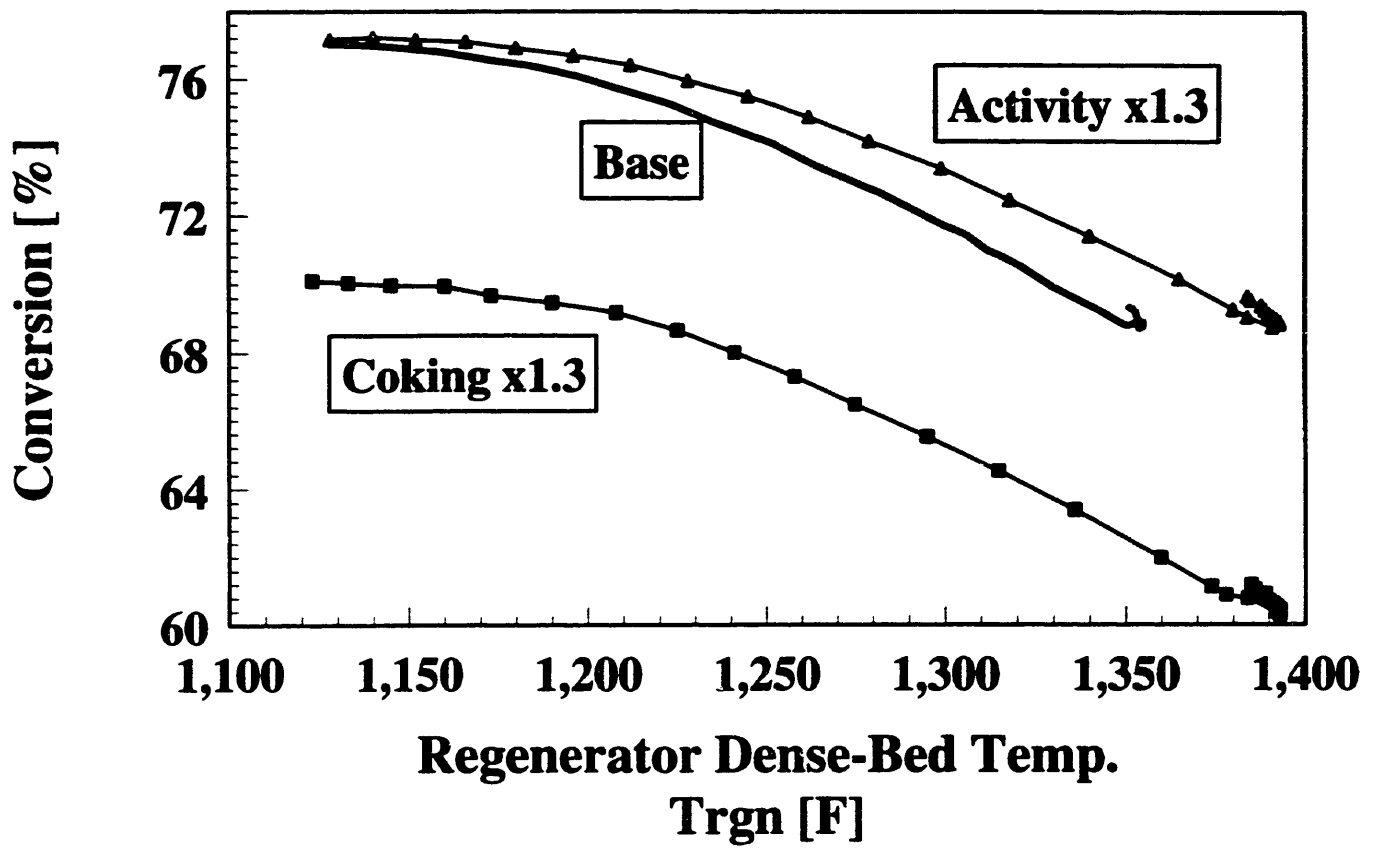


Figure 19: Effect of Activity and Coking Rate
Riser Top Temp.=1000F

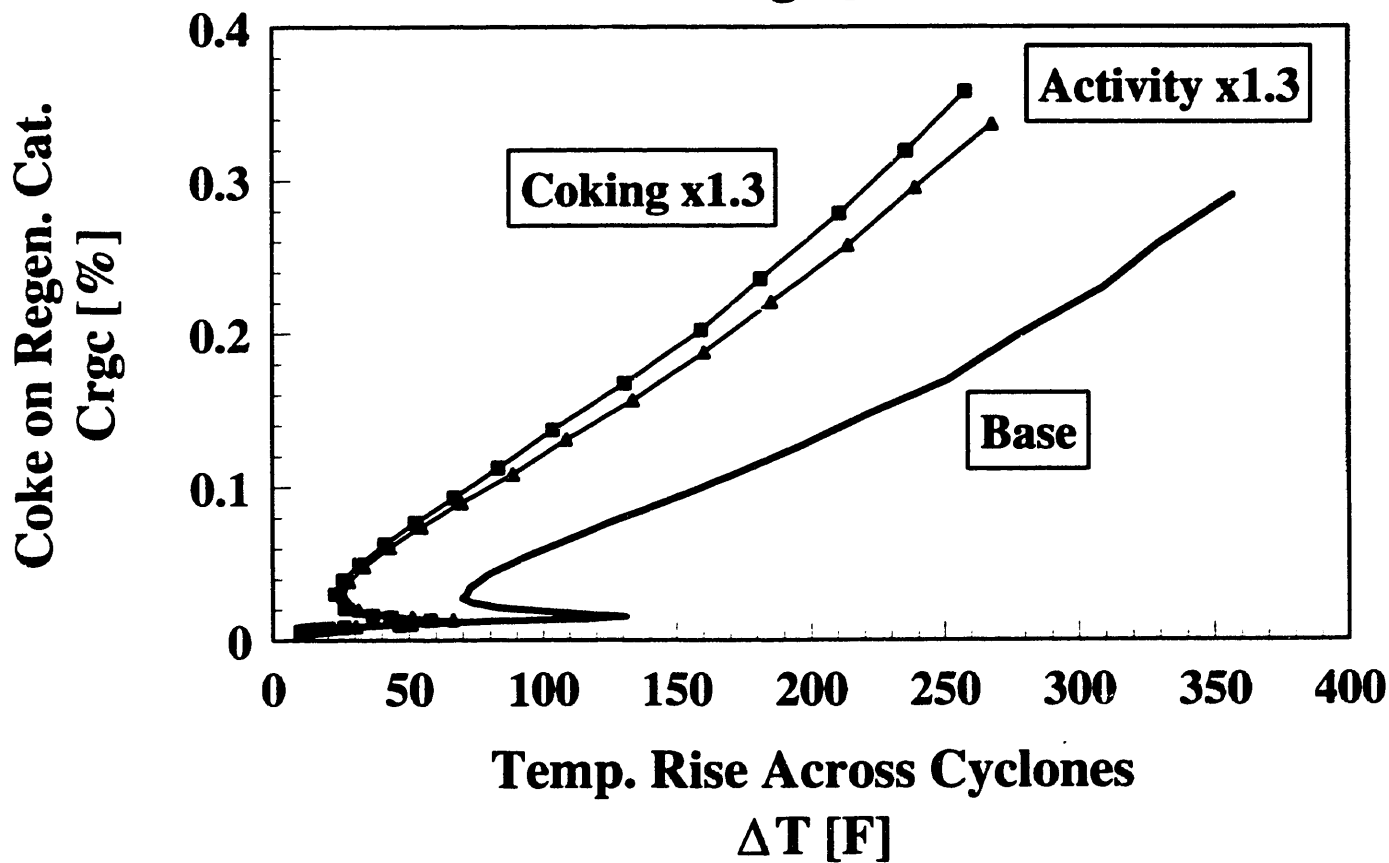
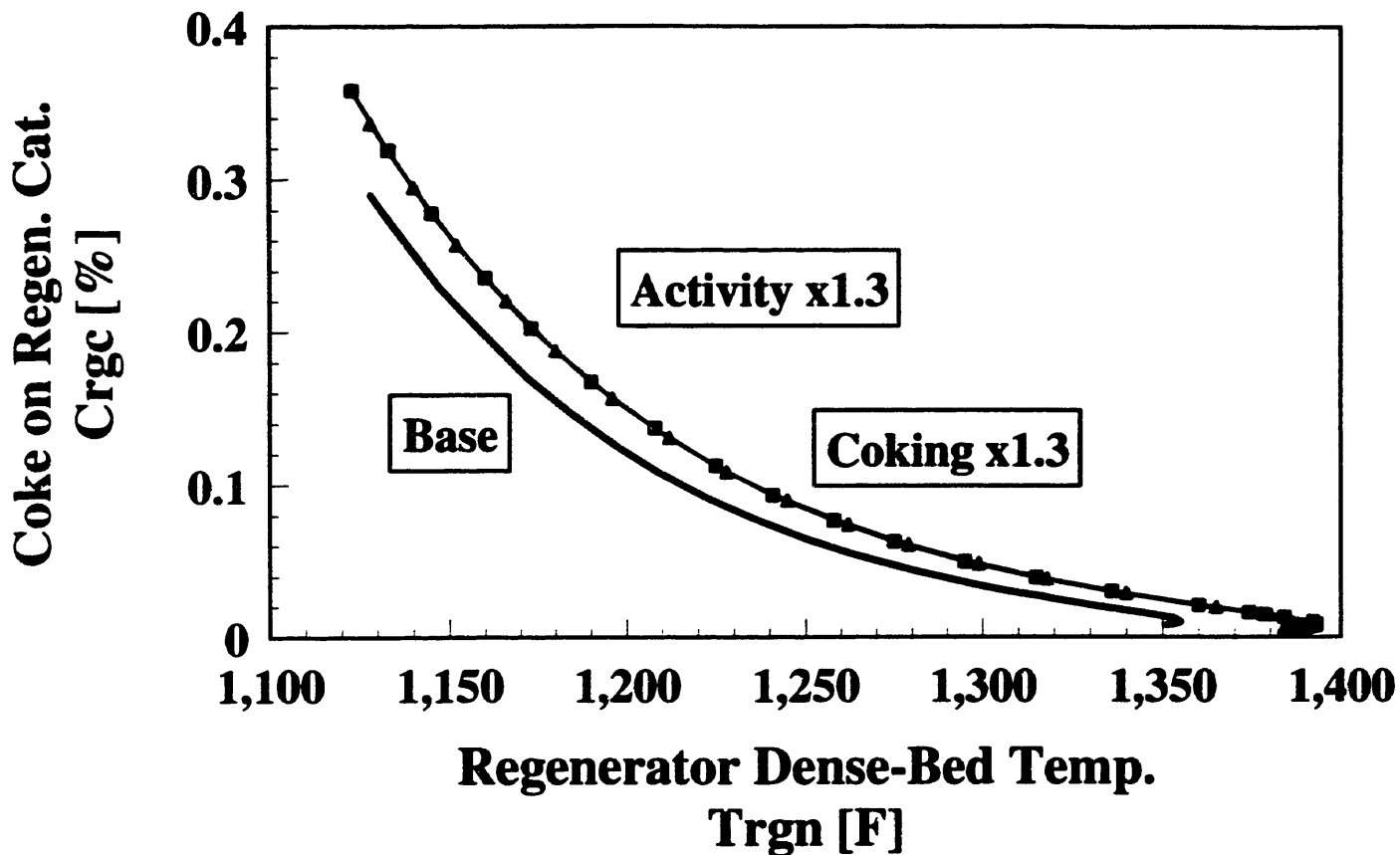
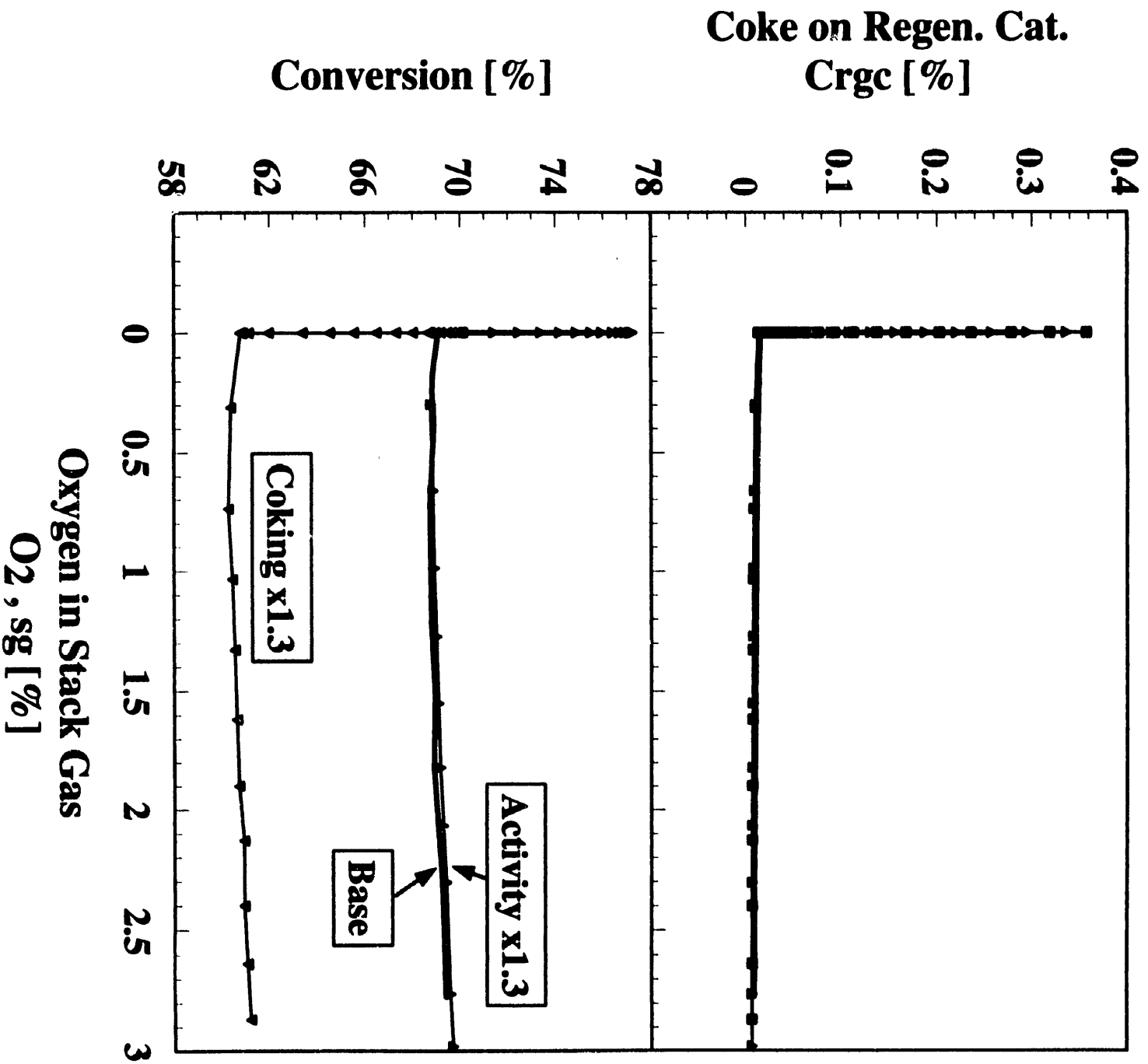


Figure 20: Effect of Activity and Coking Rate
Riser Top Temp.=1000F



**Figure 21: Effect of Activity and Coking Rate
Riser Top Temp.=1000F**

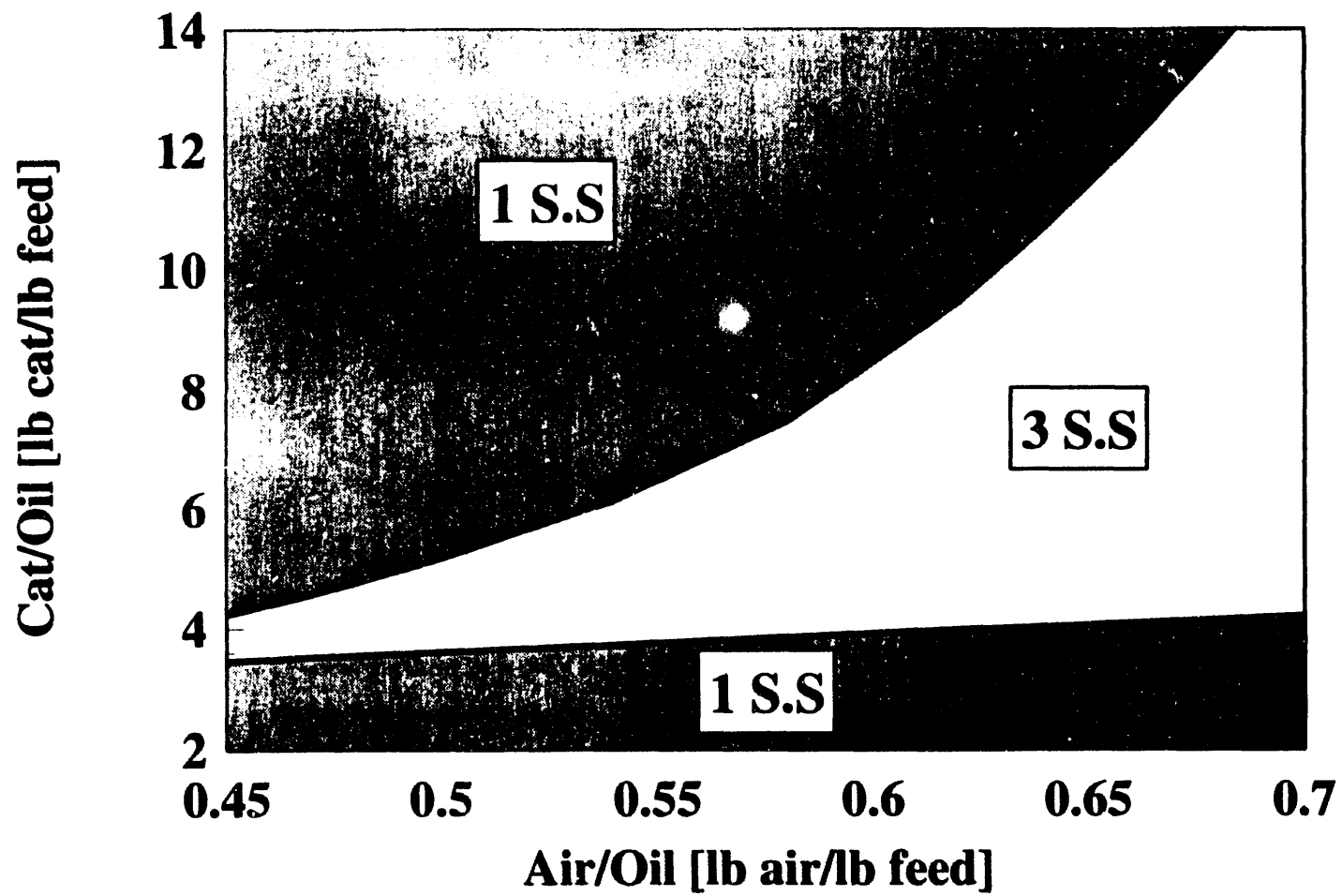


Figure 22: Map of Multiple Steady State in FCC

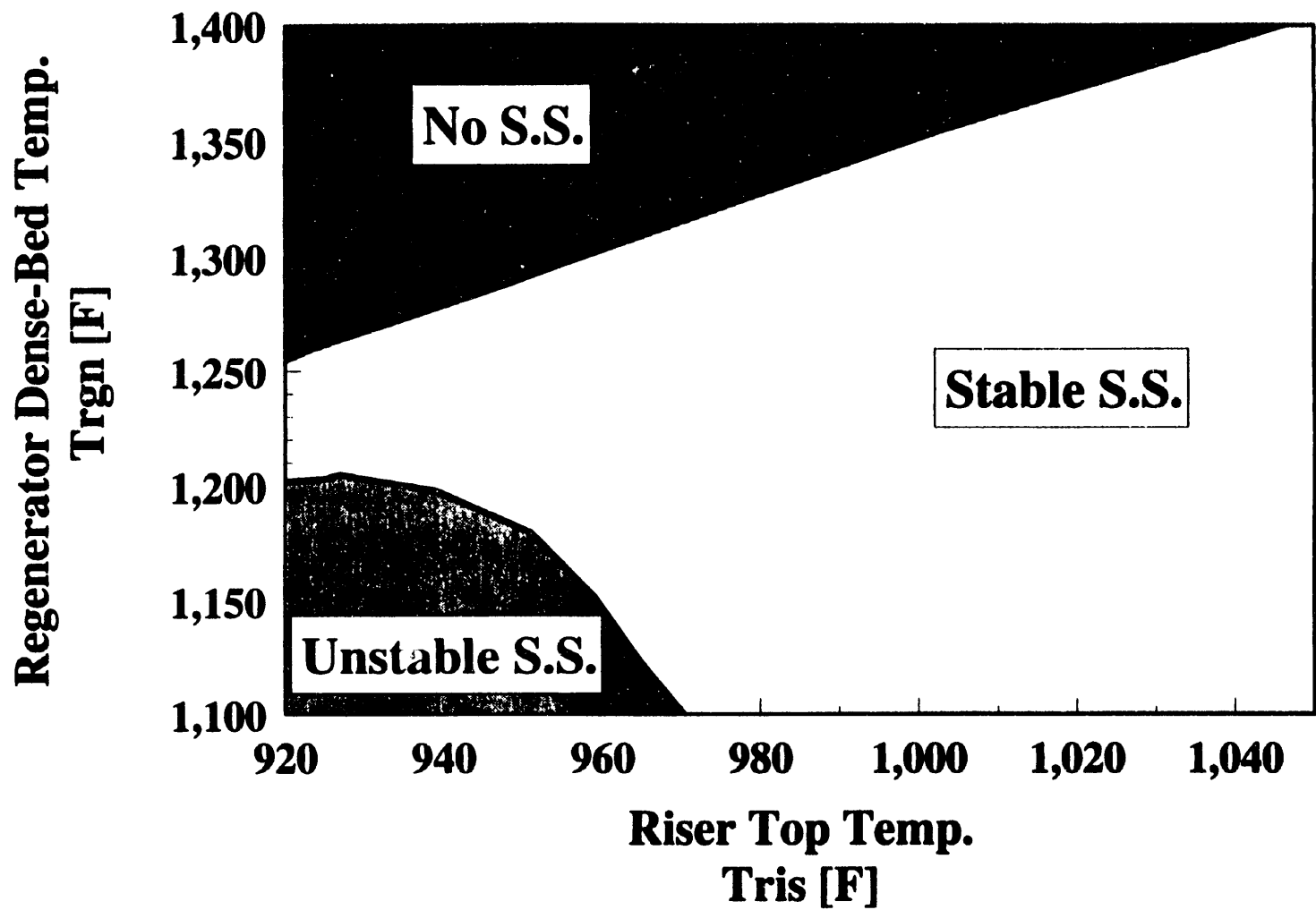
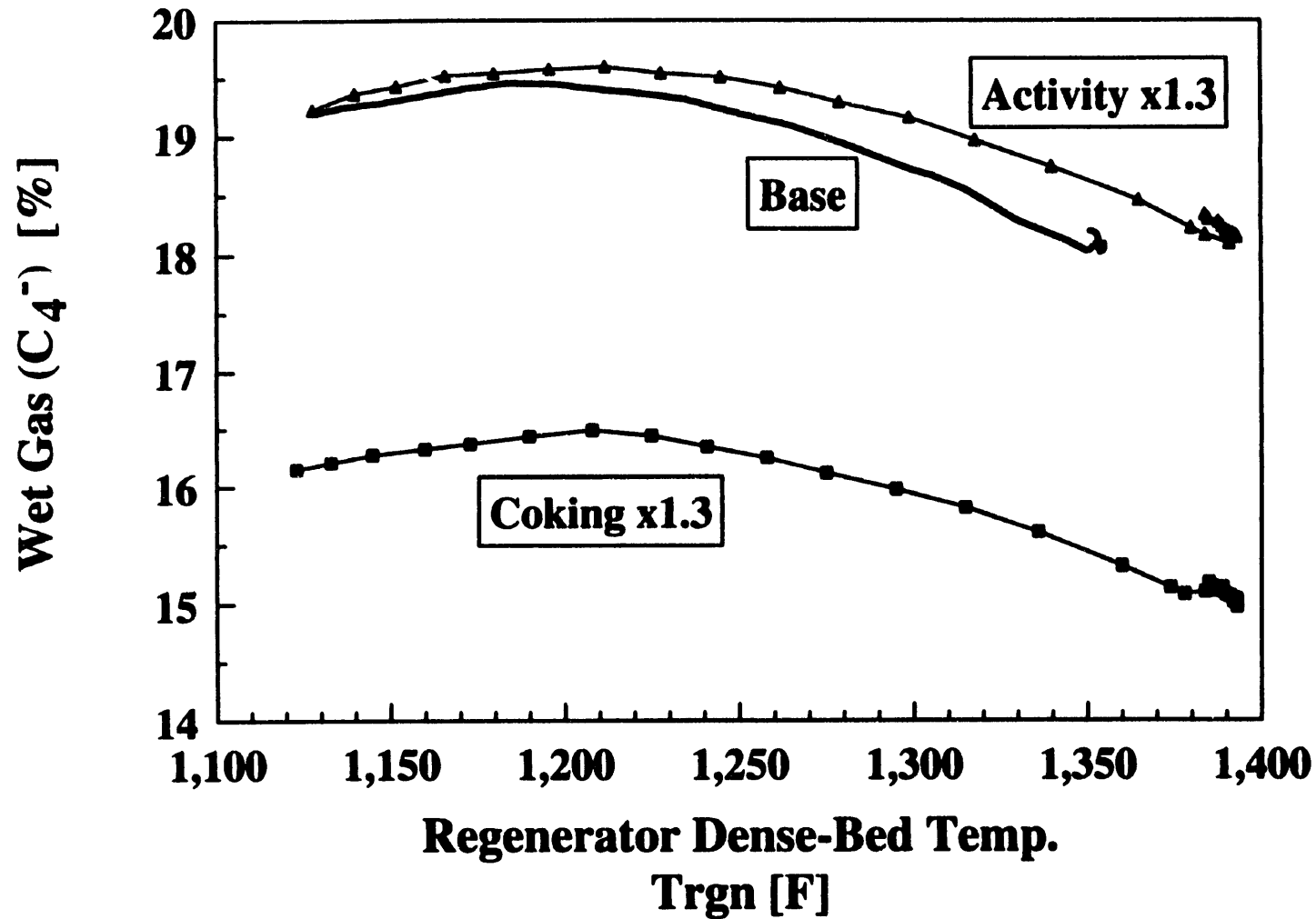


Figure 23: Permissible Operating Points Map



**Figure 24: Effect of Activity and Coking on Wet Gas Production.
Riser Top Temp.=1000F**

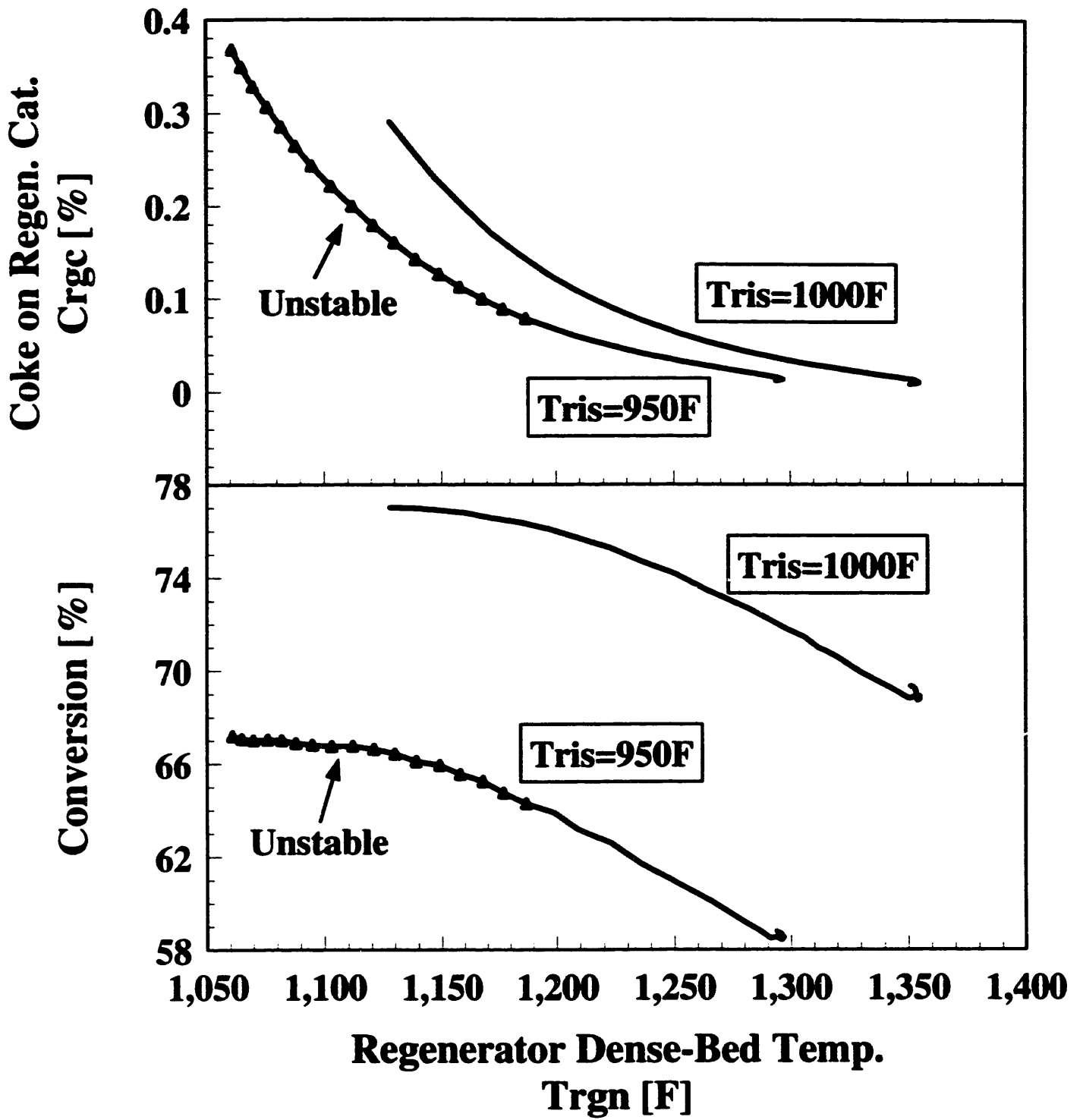


Figure 25a: Effect of Rise Top Temp.

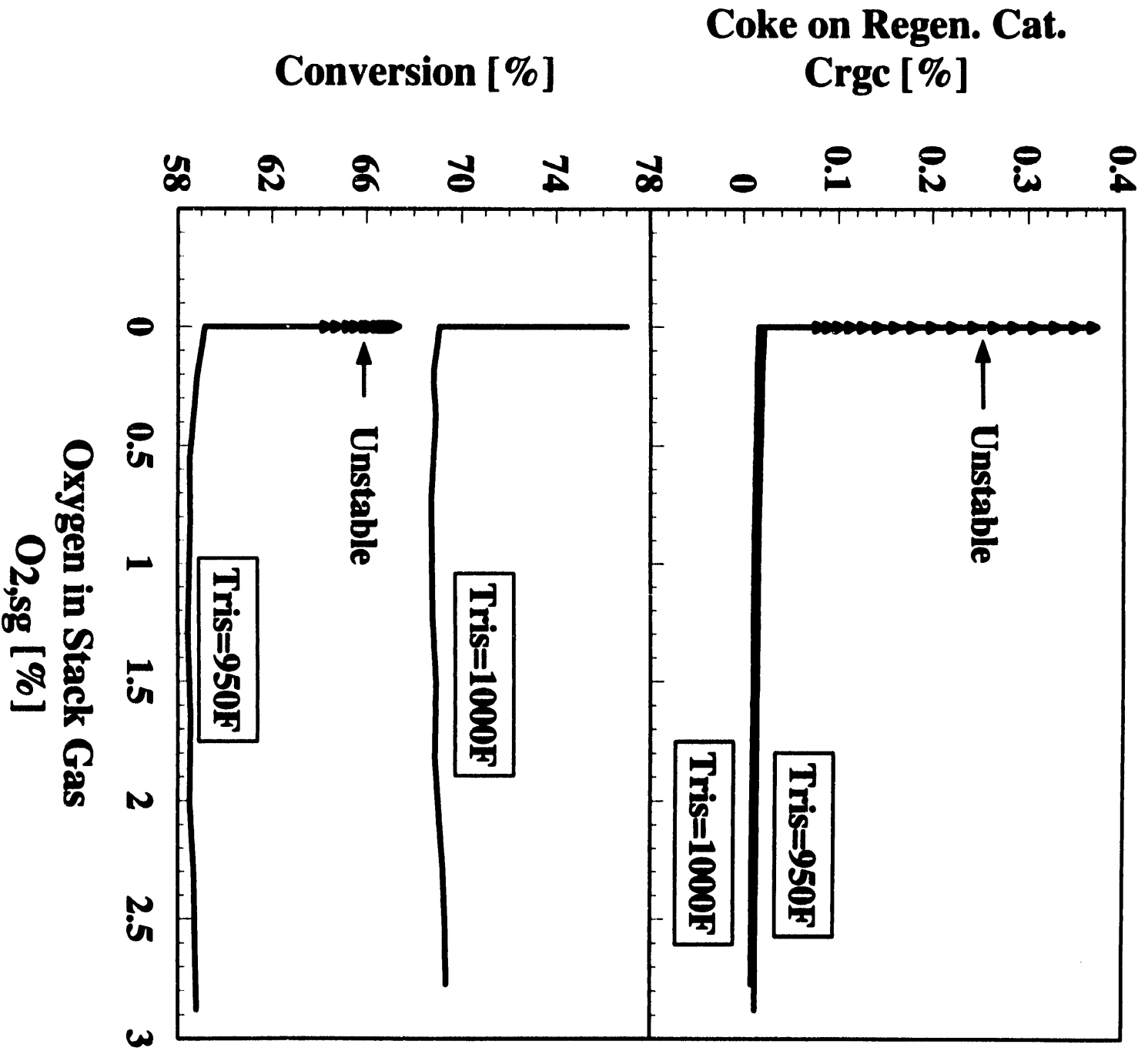


Figure 25b: Effect of Rise Top Temp.

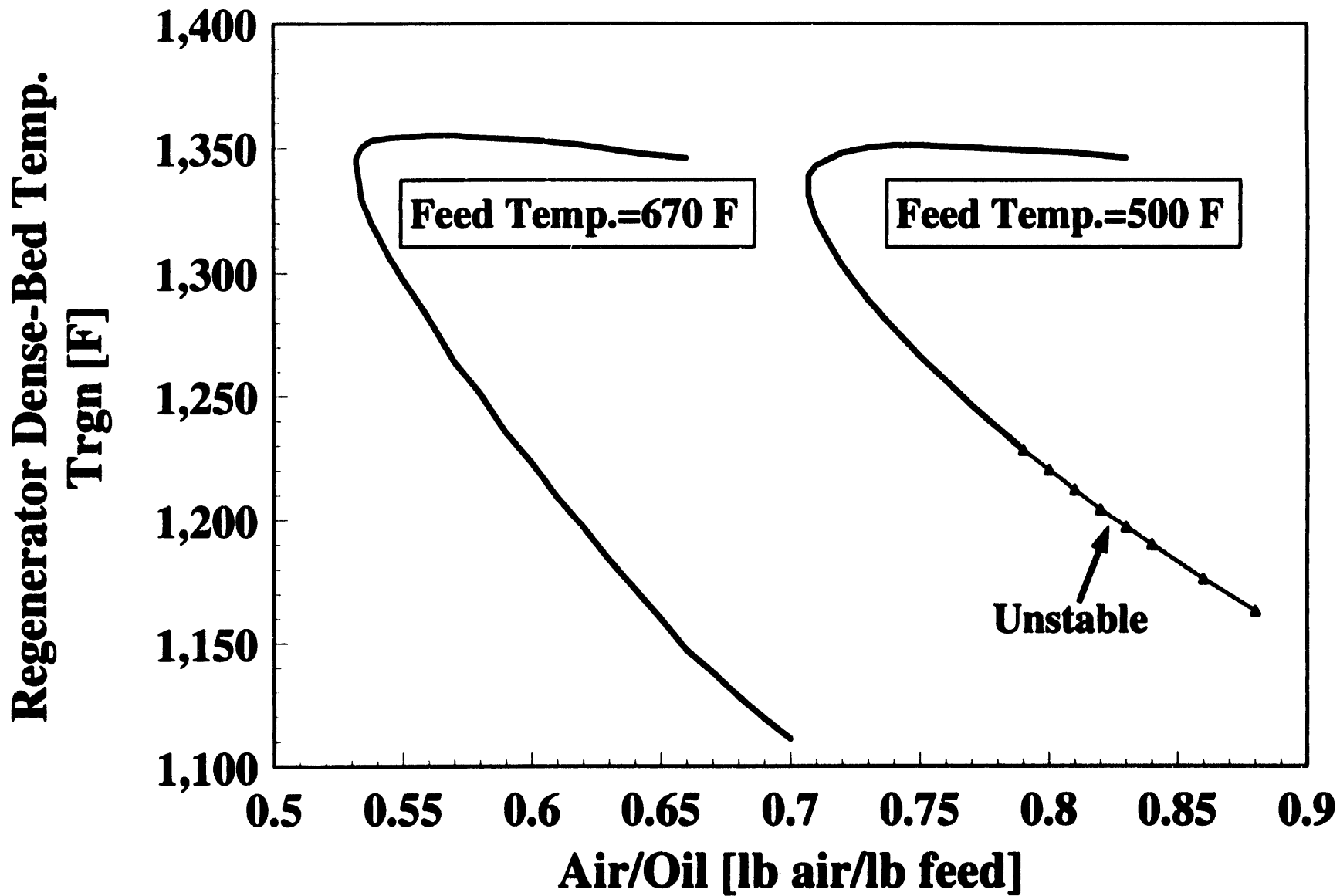


Figure 26: Effect of Feed Preheat on Steady States at Riser Top Temp.=1000 F

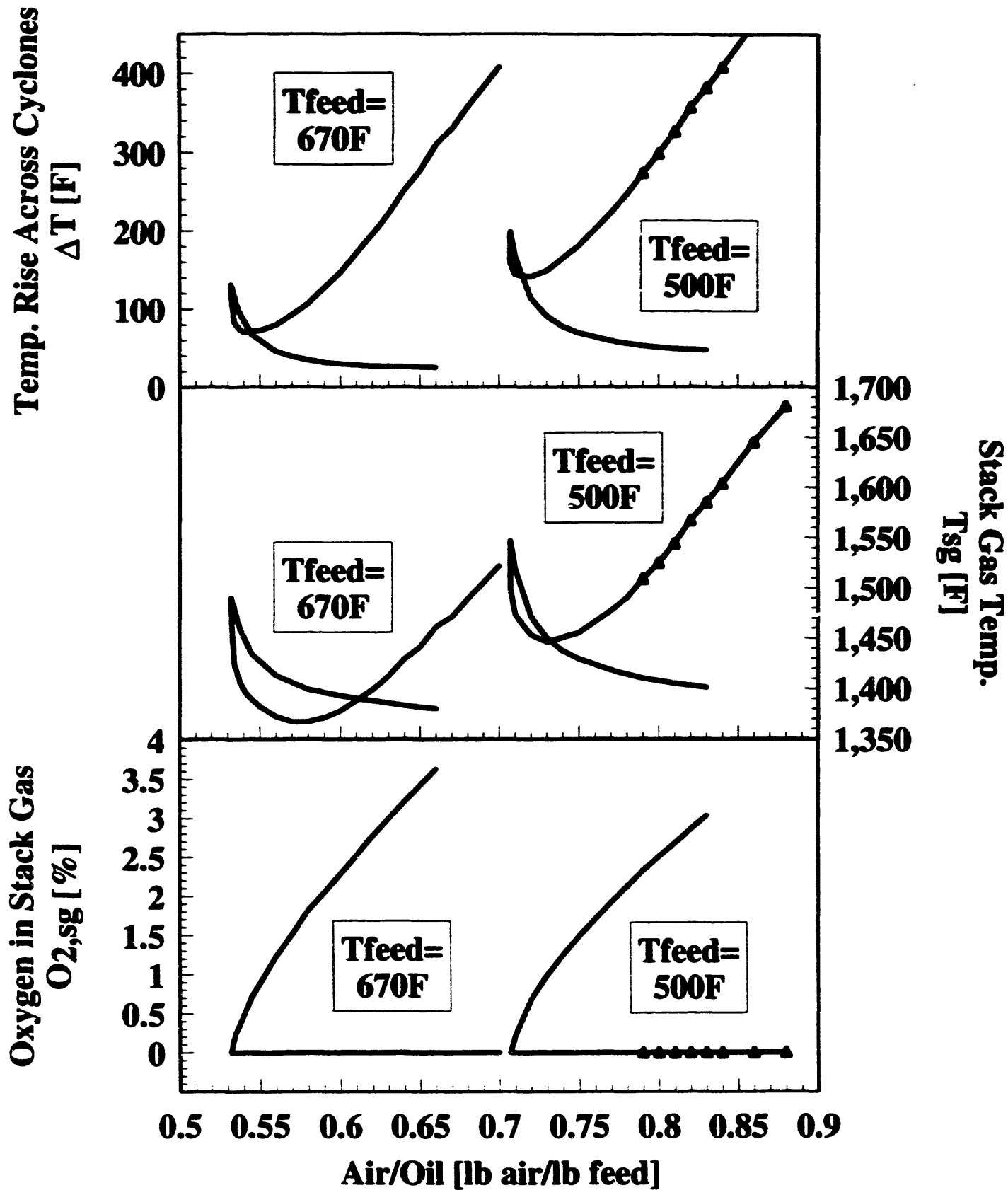
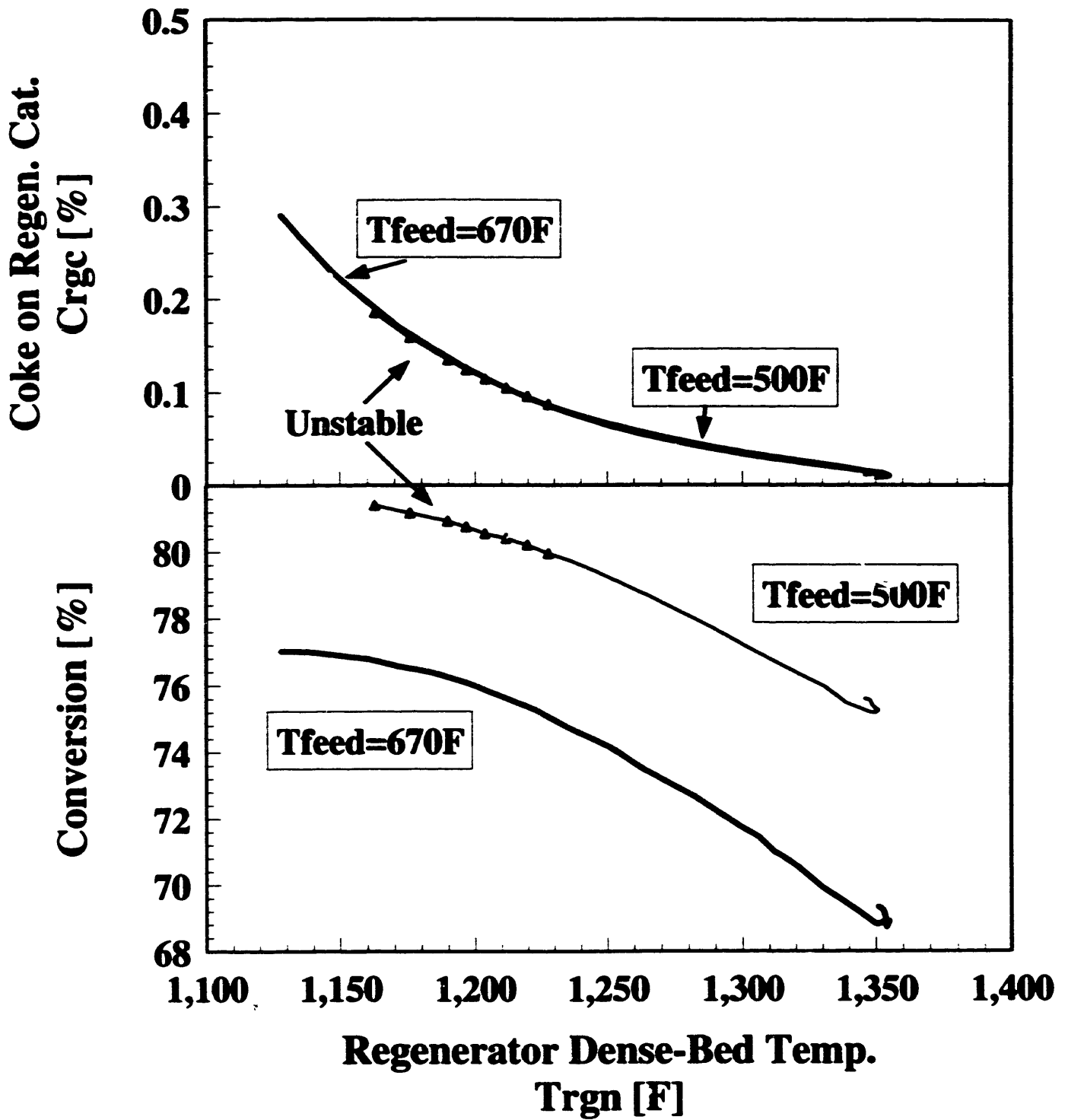


Figure 27: Effect of Feed Preheat on Steady States. Riser Top Temp.=1000 F



**Figure 28: Effect of Feed Preheat.
Riser Top Temp.=1000F**

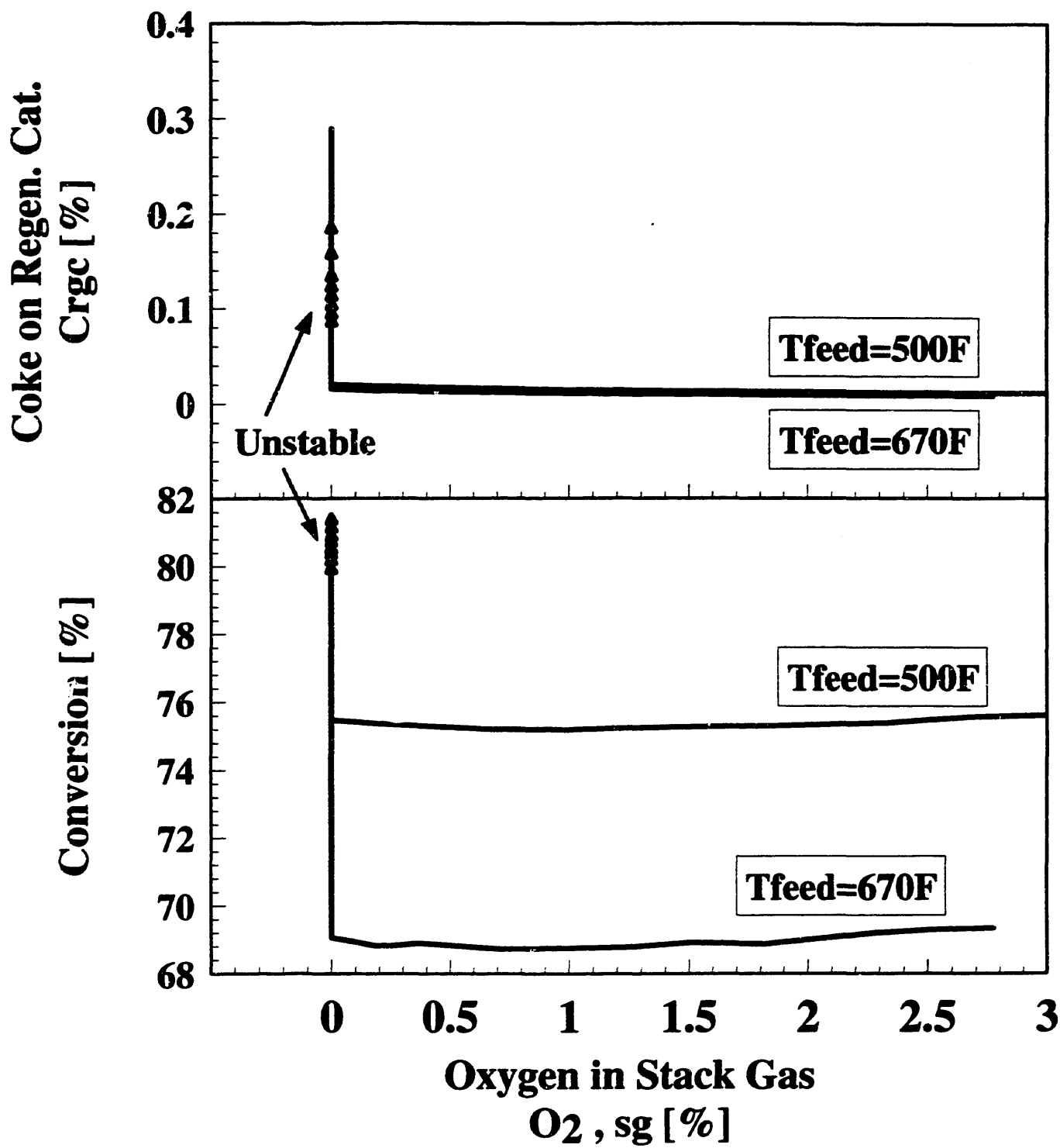


Figure 29: Effect of Feed Preheat.
Riser Top Temp.=1000F

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