PARTIAL CONTROL OF COMPLEX PROCESSING SYSTEMS

PROGRESS REPORT

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II. Description of present research and achievements till now

II.1) A dynamic and steady state model for the FCC

We chose the control of a Fluid Catalytic Cracker (FCC) for several reasons. It is a complex nonlinear system that allows to investigate most of the problems one faces in the control of complex processes.

The number of manipulated variables is much smaller than the number of variables entering the specifications. Both process constraints and product specifications are important (for a schematic description of an FCC see Fig. 1).

- Feed composition varies strongly and so do production goals and product specifications.
- A number of different designs are available allowing to study the impact of design on controllability.
- The FCC is the kingpin of the modern refinery. It is the main tool to absorb and modify the impact of changes in feedstock, and also plays an important role in modifying refinery operations to accommodate tighter product specifications due to environmental legislation. Therefore, improving its control has the strongest impact on refining.
- Based on the wide industrial experience of both investigators, we concluded that there are very few problems that one meets in control of complex chemical plants that one does not also face in FCC control.
- One of the principal investigators (RS) has a 25 year industrial experience in FCC design and operation.

We again want to point out that our goals are more than FCC control, and we are looking for control principles that can be applied on a wide scale.

We were aware from the beginning that, despite claims to the contrary, there is no FCC model available in the literature that would allow a realistic control study suitable for modern FCC operation. Present control studies are greatly hampered by that. However, there is a large amount of data in the literature that permit the formulation of such a model.

Our first task was therefore to develop such a model and test its validity by comparing it to actual industrial results. It was a more difficult and time consuming job than we anticipated, but we finished the first phase successfully in the first year, and our cooperation with Mobil Oil allowed us to check and correct the model on the basis of industrial data.

The main difference between our model and previous published models are:

1) Proper description of the CO-CO₂ combustion kinetics both homogeneous and catalytic, which is especially important in the high temperature region. Earlier models either neglect it [Kurihara (1967)], use a correlation value of over a limited temperature range [Errazu et al. (1979)] or in the case of the Amoco model [McFarlan et al. (1993)] use an unrealistic high value for catalytic combustion.

2) A more realistic description of the reactor which in our case is a plug flow riser reactor with a 10 lump model. Earlier models neglect the details of the cracking reaction and use a stirred tank model for the riser which was reasonable for old versions of the FCC but not suitable for modern FCCs. A detailed description of the model is in the final stages of preparation for publication, and the model will be made available to other investigators. A brief description is given in attachment. Further work on the model is required to give:

- a better description of product composition and properties.
- take care of nonlinear effects in the riser
- allow application to different FCC designs using catalyst coolers or two stage regenerators.

II.2) Multiple steady states and input multiplicities.

A very important and primary concern in the control of nonlinear processes is the potential existence of multiple steady states and other stability problems.

There is a considerable literature volume about multiple steady states in an FCC [Elnashaie and El-Hennawi (1979), Elshishini and Elashzie (1990), Lee and Kugelman (1973)]. Edward

and Kim (1988) report to have observed these experimentally. Regrettably, we found considerable confusion in the existing literature, and our first task was to evaluate this problem, using the large available literature on bifurcations and reactor stability to evaluate this problem. We presented a paper on the subject and the final paper is in preparation. In the quantitative evaluation we used our model, but we also showed that this problem can be evaluated based on laboratory experiments, using heat balances and very limited detailed knowledge. This is important to demonstrate as in many nonlinear systems detailed models are unavailable.

An FCC has always at least three steady states. One can easily show that with available preheat of oil and no preheat of air an FCC will not self ignite and always has one cold stable steady state. Therefore, if there is a hot steady state, there also has to be an open loop unstable steady state and we showed that it could occur in actual operation. The question is are there more than three? Elshishini and Elnashaie (1990) have tried to show the possibility of such steady states due to mass and heat transfer from the bubbles to an emulsion phase, analogous to multiple steady states in catalyst particles, but we showed that this requires totally unrealistic assumptions about transport processes in fluid beds.

We analyzed all the possibilities for more than three steady states and found that the only realistic source is homogeneous CO-CO₂ combustion, which is a consecutive reaction of the CO formed by surface combustion of coke. However, this reaction is inhibited by the pressure of catalyst (or any solid) at temperatures below 1300°F. We showed that if one uses conventional activation energies to model the homogeneous combustion of CO five steady states are not feasible. Their existence requires a very

Figure 2: Effect of Combustion Promoters on Multiple Steady States

large increase in the CO-CO₂ combustion rate over a temperature difference of 100°F. However, it is known that one cannot describe the homogeneous CO combustion reaction in the presence of a catalyst by such kinetics by using a single apparent activation energy. If one uses data published by Upson et al. (1993) to model the CO-CO₂ combustion in the temperature range between 1200° and 1350°F five steady states become likely for some operating conditions. We showed that a small amount of catalytic combustion promoter would eliminate the chance for the additional two steady states. This is illustrated in Fig. 2. We also showed what this would do to control.

However, no industrial or experimental confirmation of additional high temperature steady states has been published.

Figure 3: Multiple Steady States at Constant Air/Oil=0.55

Edward and Kim (1988) report a case where an FCC left on manual control jumped to another steady state. But in their case the system operated with the reactor temperature control loop operating in closed loop and only air rate on manual control. This is a case of input multiplicity as can be seen from Fig. 3. If air rate is fixed and riser top temperature is kept constant by manipulating catalyst circulation rate there are two possible open loop stable high temperature steady states only one of which is

stable with a given linear controller with integral action. Which one is stable depends on the sign of the gain chosen for the linear controller.

The potential for input multiplicities is an important factor when considering the choice of the dynamic control matrix, as was

demonstrated in our evaluation of the control matrix.

The fact that the existence of a cold steady state is unavoidable with present design has, in control, led to a strong emphasis on preventing a wind down. A change in feed stock properties and catalyst activity can lead to the loss of the steady state at a given controller setting. Understanding the impact of such changes is therefore important for efficient control. This is especially so with advanced control methods such as DMC which are based on linearized models. Our work showed how the stability limits depend on the choice of operating conditions, feed and catalyst properties and the level of catalyst CO-CO₂ promotion. See Fig. 4.

Partial Control Dominance and Sufficiency

The concepts and principles of partial control for complex systems were further defined and investigated in our work till now and the results are described in reference 11 (attached). Our work on FCC control demonstrated how conventional approaches in the design of the control system can lead to wrong results.

Classical control theory deals mostly with controller design for a given square matrix. In partial control the number of measured and controlled variables is larger than the number of manipulated variables available (see Table 1 for the FCC). We therefore can control the desired vector of specification of Y_p only in a space

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

To achieve this we choose a smaller square matrix G_d for fast dynamic control and use the setpoint of G_d together with other manipulated variables in a slower steady state control scheme G_s to control Y_p . G_d is commonly based on linear models, whereas G_s requires a nonlinear model.

Figure 4 a: Permissible Operating Range - Base Case

Figure 4 b: Permissible Operating Range - 30% Decrease in Activity

Figure 4 c: Permissible Operating Range - 30% Increase in Coking Rate

Figure 4 d: Permissible Operating Range - Effect of CO Combustion Promoter

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	

We demonstrated some of the principles involved by evaluating several 2x2 control matrices that have been proposed and used in FCC control (see Fig. 5). Conventional approaches using fast and smooth response as a criterion give a preference to the pair $[T_{sg}, T_{ris}]$ (stack gas temperature and reactor top temperature), despite the fact that industry abandoned it.

On the other hand if the mixing goal is to control conversion and product properties, consideration of dominance will favor the pair $[T_{rgn}, T_{ris}]$. For a variable to be dominant, it has to have a strong impact on a majority of Y_p . On the basis of kinetic considerations the pair $[T_{rgn}, T_{ris}]$ has that property. On the other hand, if we are interested mainly in stability or preventing a wind down the pair $[T_{rgn}, \Delta T]$ is dominant. Stability is only a function of the heat balance. Choosing $[T_{rgn}, \Delta T]$ as the controlled variable in the primary fast matrix assures sufficient coke make, as well as sufficient oxygen to burn it down (ΔT is correlated with excess oxygen at the cyclone inlet). The effect of dominance is shown in Table 2.

Table 2 gives the impact of an increase in catalyst activity or coke make on two different steady states, keeping different pairs of controlled variables constant. One notes the superiority of $[T_{rgn}, T_{ris}]$ in minimizing the change in conversion, wet gas make and coke on regenerated catalyst. The case $[T_{rgn}, \Delta T]$ is in this respect the worst as it does not keep T_{ris} constant. T_{ris} is the most dominant variable in the reactor.

Figure 5: Primary Control Pairing

Table 2: Summary of changes to step-change in Activity and Coking Rate (30% increase in both). All changes 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_{rgc}) [wt%]	0.148	0.0639
Coke on spent cat. (C_{sc}) [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 (in percent):

30% increase in activity								
	Case I				Case II			
Controlled	T_{rgn}^{ris}	$T_{ris}^{\Delta T}$	T_{sg}^{ris}	$T_{rgn}^{\Delta T}$	T_{rgn}^{ris}	$T_{ris}^{\Delta T}$	T_{sg}^{ris}	$T_{rgn}^{\Delta T}$
ΔT_{ris}	0	0	0	-3.3	0	0	0	-4.0
ΔT_{rgn}	0	-3.0	-5.5	0	0	-3.1	-8.1	0
$\Delta \Delta T$	-30.6	0	30.6	0	-19.4	0	45.3	0
ΔT_{sg}	-5.0	-2.8	0	0.1	-3.3	-2.8	0	0.1
ΔC_{rgc}	23.7	82.6	152.8	-12.5	13.5	42.7	119.2	-7.0
ΔC_{sc}	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_{rgn}^{ris}	$T_{ris}^{\Delta T}$	T_{sg}^{ris}	$T_{rgn}^{\Delta T}$	T_{rgn}^{ris}	$T_{ris}^{\Delta T}$	T_{sg}^{ris}	$T_{rgn}^{\Delta T}$
ΔT_{ris}	0	0	0	-4.3	0	0	0	-5.7
ΔT_{rgn}	0	-3.8	-6.7	0	0	-3.8	-9.5	0
$\Delta \Delta T$	-37.2	0	37.6	0	-22.6	0	53.3	0
ΔT_{sg}	-5.9	-3.3	0	0.4	-3.7	-3.4	0	1.1
ΔC_{rgc}	22.0	102.3	195.8	-23.0	12.3	51.3	160.5	-14.6
ΔC_{sc}	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

Our results also show that the matrix $[F_{air}, F_{cat}, T_{rgn}, T_{ris}]$ cannot be used at all setpoints. In some range of setpoints the gain is infinity and in another the gain is zero. There is no economic incentive to operate in the range of infinite gains, but the region of zero gain occurs when full CO_2 combustion is used. There, the only choice is to use $[T_{ris}, O_2]$ (riser top temperature and excess oxygen) as controlled variables. In that case one has to control T_{rgn} in G_s by using another manipulated input (T_{feed} , catalyst properties, or feed composition). This is demonstrated in the paper.

Another important consideration is input multiplicities. The impact on the choice of control variables is shown in Fig. 6, where we compare three pairs $[T_{ris}, T_{rgn}]$, $[T_{ris}, \Delta T]$ and $[T_{ris}, T_{sg}]$. We note complex input multiplicities in $[T_{ris}, T_{sg}]$. Also, there is a zero gain in the area of present commercial operations $[T_{rgn}, \Delta T]$ has a wide range where it is applicable, but is less attractive in the range of presently common operating conditions, which involve higher temperatures.

Figure 6: Effect of CO Combustion Promoter on Steady States

Figure 7: Effect of Riser Top Temp. on Steady States

Fig. 6 is given for fixed riser temperature which is kept constant by manipulating catalyst flow rate. All inlet conditions are kept constant. Our work showed the impact of different input conditions, such as feed temperature, feed composition in catalyst activity. Fig. 6 shows that at lower T_{ris} , some choices of $[T_{rgn}, T_{ris}]$ will lead to open loop unstable operating conditions. The likelihood increases if feed temperature or catalyst activity is decreased. While one can stabilize these loops by control, one can also stabilize them by adding a catalytic combustion promoter that catalyzes the CO-CO₂ combustion (see Fig. 7). The impact of the promoter, which is always based on units that operate in full CO₂ combustion, on steady state control was evaluated. It will cause a slight decrease in conversion. Our results explain why 30 years ago stability and prevention of wind down were the major concern in FCC control, whereas they are less of a concern with modern catalysts and operating conditions. Operator strategy has therefore changed.

The use of $[T_{ris}, \Delta T]$ and $[T_{ris}, T_{sg}]$ in cascaded control to protect the cyclone was also evaluated, and $[T_{ris}, \Delta T]$ was again favored by nonlinear steady state considerations [see Fig. 8] despite the fact that analysis of the linear response at some setpoints will favor $[T_{ris}, T_{sg}]$.

Linearized transfer functions can be used to evaluate dominance and for a given setpoint will give results very similar to Table 2. One has to take into account the changes in the transfer function. However, when developing stability one has to consider the nonlinear properties of the model.

Figure 8: Effect of Riser Top Temp. on Steady States

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 to 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 the system stable, 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_{ris} and T_{rgn} have this property. ΔT and T_{sg} are strongly sensitive to unit design, and nonidealities are 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. 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 2×2 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.

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