Advanced Control Practice in the Chemical Process Industry: A View from Industry

This paper presents a commentary on the status of application of modern control theories to industrial chemical processes. The difficulties encountered in the implementation of advanced control to chemical reactors are critically examined in order to determine the reasons for its limited success to date. Chemical processes, very often poorly understood because of the physicochemical complexities, are nonlinear, highly interacting, spacially dependent, and continuously disturbed by many uncharacterized noises. It is time-consuming and expensive to develop advanced control methods for these systems because few of the available theories can be applied directly. Thus, we often find it difficult to economically justify advanced control projects.

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Modern control methods have not made the expected impact on the control of chemical processes as originally anticipated in the first flush of enthusiasm. This has been the subject of many discussions and publications (for example, Rosenbrock, 1962; Denn, 1967; Foss, 1973; AIChE Research Subcommittee Report on Reaction Engineering, 1974; Kestenbaum et al., 1976). The theoreticians argue that those involved in control applications are unable to understand the theory, while the practitioners argue that much of the theory does not apply to real processes. Nevertheless, all agree that there exists an unmistakable gap between theory and practice. The leading periodicals in process control are replete with papers on optimal control of various chemical processes; however, many are academic exercises and hardly represent the complexities and difficulties in the real world. It is encouraging to see that the academicians are questioning, in a constructive way, the directions and values of the latest developments in the process control theory (Foss, 1973).

In the present paper we examine, strictly from the viewpoint of practitioners, the inherent difficulties in the application of advanced control theory to chemical processes. The term advanced control here is loosely defined as any control techniques beyond the linear theory of single input-single output control; however, our discussion will be directed toward more sophisticated control theories such as optimal or noninteracting control. The term chemical process implies broadly any typical plant common in the chemical process industry; however, our emphasis will be on chemical reactors. Control of unit operations, such as distillation, evaporation, or adsorption columns, has been treated by many others (for example, Fisher and Seborg, 1973; Shinsky, 1967) and will not be discussed here.

Those of us involved in the control of chemical processes are inclined to think that our problems are unique among all other control problems. Rightly or not, we have many reasons to believe that this is the case.

The difficulties encountered in chemical process control may be grouped into two broad categories, the first of which is related to understanding the process itself and the second to applying the theory to commercial plants. The difficulties inherent in the process characterization are reasonably well recognized; however, those associated with the implementation have received less attention. In the ensuing discussion we will rely heavily upon our own experience in the petroleum industry, but we are assured through numerous consultations with our colleagues in other process industries that these difficulties are common throughout the process industry.

DIFFICULTIES IN CONTROL DESIGN

Unknown Process Characteristics

The single most difficult problem to be overcome is understanding the process itself. Chemical processes are generally characterized by large dimensionality, strong interaction among process variables, and nonlinearity. Typical examples encompass multicomponent reaction system with catalyst decay and Arrhenius temperature dependency, heat and mass interactions in distillation column control, or propagation of burning fronts in a fixed-bed catalyst regeneration. For many industrially important processes which have been in operation for a number of years, we frequently find that we do not fully understand even their steady state behavior, let alone the dynamics. This can be attributed mostly to the inherent complexity of chemical and catalytic reactions. Our limited ability to measure state variables also aggravates the situation. Only very limited quantities, such as temperature, pressure, and flow rate, are readily measurable. Implicit quantities, such as catalyst activity, gasoline octane number, and carbon-on-catalyst (as in fluid catalytic cracking) can also be measured but are either expensive or time-consuming. Besides, many unknown or uncharacterized disturbances persist. Some disturbances consist of slow, drifting changes in feedstock compositions or temperature changes in the environment, while others are high frequency, steplike fluctuations in pressure from compressors or blowers.

Perhaps the most common and serious disturbances are those related to the feedstock compositions, impurities (for example, sulfur, nitrogen, metals), and catalyst quality. Slight changes in the crude source or in the operation of crude distillation towers can influence significantly the operation of downstream units such as the reformer, hydrocracker, and catalytic cracker. The product distribution from a given reactor generally changes considerably when catalyst qualities vary as a result of changes in manufacturing process or poisoning. The feed to the product distillation column now becomes quite different from the design feed, thus requiring major adjustments in the column operation. Mechanical failures of auxiliary facilities, such

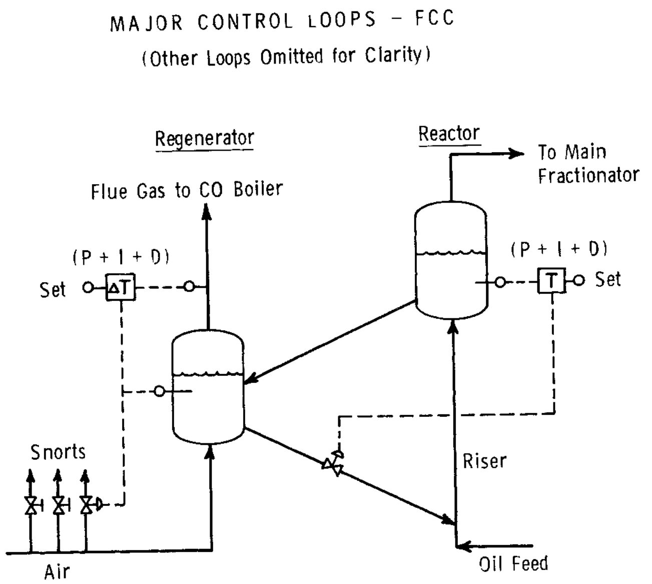

Fig. 1. Schematic of conventional control scheme.

as pumps, heat exchangers, or steam generators, require shifts in the heat removal rates around a distillation column. The quality of product streams may be detrimentally affected if these columns are not adequately controlled. The disturbance will then propagate to the next downstream unit.

Unfortunately, it is not easy to make reliable on-line measurements of many vital quantities. Chemical compositions of complex feedstocks, catalyst properties and activity, and product qualities must generally be determined in the laboratory. Thus, critical information about disturbances are usually not known at the time of occurrence. This renders the task of characterizing disturbance and understanding the process dynamics very difficult. Quantitative characterization of the dynamics for rational design of advanced control is clearly time-consuming.

Process Design Vs. Dynamics

Small changes in the process design could influence the process dynamics profoundly. Thus, control configurations must be considered in the design of the process itself. For distributed or multivariable processes, it is not obvious how to determine the best configuration of control loops. Which temperature should be measured and controlled by which valve? This freedom of choice may be regarded as an advantage in disguise to the control designer because it may provide an excellent opportunity to improve the closed loop dynamics by simply switching a few control loops around. This type of concept has been successfully applied to improve the control of multicomponent distillation columns (for example, Shinsky, 1971). Nevertheless, it is not easy for the designer to select the best control configuration among a number of alternatives without reliable dynamic simulation. The enumeration and comparison of all of the possible alternatives may eventually lead to the solution, but the combinatorial problem may be prohibitive for relatively complex reactor systems.

In the process design, the design engineer is concerned mostly with the steady state characteristics with little consideration about the dynamics. The control scheme is then derived mostly from his past experience. The difficulties in determining control configurations can be best illustrated with an example of a fluid catalytic cracker commonly used in petroleum refineries.

A brief description of the catalytic cracking plant (FCC) is in order. This process (Figure 1) has been widely employed in converting heavy hydrocarbons into more valuable lighter products such as gasoline. The catalytic section of the FCC system consists of two vessels: a reactor and a regenerator. Gas oil feed (boiling typically between 430° and 900°F) is introduced at the bottom of the riser, where it meets hot catalyst from the regenerator. The cracking reaction starts immediately, while the mixture of catalyst and gas oil passes through the riser into the reactor. During the encounter of catalyst with gas oil, carbonaceous material deposits on the catalyst surface, which consequently deactivates the catalyst. The product vapor is transferred to the fractionation section, and the spent catalyst flows down through the reactor standpipe into the regenerator, where the carbonaceous deposits on the catalyst surface are burned off. The regenerated catalyst passes through the regenerator standpipe into the riser. This completes the catalyst circulation circuit.

For the FCC plant (Figure 1), for instance, it is not obvious if the reactor temperature should be controlled by the catalyst flow rate, feed rate, feed temperature, or even by feed qualities. How about the control of regenerator temperature? Typically, there are more than a dozen thermocouples available in the regenerator. Which temperature or temperatures should be controlled? And one can think of many similar questions to be answered before any final control configuration is arrived at. In the most conventional control scheme of the FCC (Figure 1), the reactor temperature is controlled by the regenerated catalyst circulation rate and the air flow rate by the excess oxygen content in the regenerator. Of course, there are many other control loops in the FCC plant; for the sake of clarity, however, we will concentrate only on those affecting the reactor-regenerator heat balance. It is not clear how this current configuration was arrived at, but, as will be seen later, there is no assurance that this particular configuration is the best control scheme, although it has been in operation on many units for many years

Lack of Good Process Models

Rational design of control systems requires both steady state and dynamic information about the process. The more sophisticated the control system, the less the tolerance of inaccuracies in the model. A complete and perfect model is not only technically impossible but not even necessary. What is needed is a reasonably good model which accounts for major process variable effects and the dominant dynamics. This has previously been called the principle of optimum sloppiness (Prater, 1970).

For chemical reactors, essential parameters to be estimated include kinetic rate constants for major chemical reactions as well as poisoning and fouling. Prediction of catalyst activity and product selectivity of a commercial plant is still in the state of an art rather than a science. Modeling of commercial reactors and their fluid dynamics as related to process variable changes frequently requires a significant effort. For multicomponent distillation columns, for instance, extensive thermodynamic properties and considerable commercial data are required to characterize tower behavior. It is clear that the current state of the art in model building of petroleum and chemical processes still has considerable room for improvement (Weekman, 1974). Simplified models of complex processes are frequently inadequate in describing the true process characteristics, thus leading to erroneous control design.

For example, in Figure 2 the steady state temperature behavior of the FCC is simulated by using two different process models of recent publications and plotted against the feed rate. One shows a very flat monotonic response and a unique steady state (Lee and Kugelman, 1973), while the other shows multiple steady states for certain ranges (Iscol, 1970). The dynamic characteristics are also completely different. The contrast between these two models of one of the most widely studied industrial processes is too large to be attributed to any design differences. The unusual multiple steady states of Iscol's model can be attributed to the peculiarity of a temperature cubic term in the coke-make correlation. The most important point to be emphasized here is that the characterization of complex industrial processes exemplified by the FCC is very sensitive to the parameters and types of models used. Consequently, the choice of the best control scheme for the FCC process, with a particular type of model used, may not be representative at all.

Conflicting Control Objectives

As pointed out by Shinnar and his co-workers (Kestenbaum et al., 1976), one can think of several criteria, some of which are conflicting, that must be considered in the evaluation of process controllers. The ability to maintain a controlled variable at a desired set point is by far the most common objective. The response to changes in set points or loads should be fast and smooth; however, it is usually more important in the process industry to maintain the transient behavior within reasonable limits than to achieve the minimum time response with large excursions.

Asymptotic stability, satisfactory frequency response over wide ranges, insensitivity to parameters and models used, and no excessive control action are other requirements to be satisfied by process controllers. Notice that fast response and no excessive control action are contradictory. It is not easy to define quantitatively an objective function that satisfies all of these requirements. A compromise is unavoidable here, for which optimal control theory seems not well suited. Even for a very simple first-order process with time delay, Kestenbaum et al. (1976) found that the overall performance of the conventional PID controller is more satisfactory than the optimal controller.

DIFFICULTIES IN APPLICATION

From the above discussion it is clear that we fully agree with Foss (1973) and Shinnar (see Kestenbaum et al., 1976) in their assessment of the value of advanced control techniques to the process industry. We feel, however, that they missed some very vital elements which must be considered in any control project in the process industry.

Economic Incentives for Process Control

In general, the overall value of the project is dominated not by the controllers' action but by the ability to predict the best operating point in terms of process variables. Economic incentives for advanced control may be considered to come from three different categories. First, the major benefit results from moving the steady state operation to a better operating point. Second, improved control allows operation closer to any limiting constraint, thus resulting in additional benefits. Lastly, smaller excursion and faster response to set point changes and load disturbances move the operation to the desired point faster and with less variation in product yields. For many petroleum plants, the first of these three incentives dominates. Thus, where 3 to 5¢/barrel might be achieved by holding an FCC plant under tight control near a desired point or constraint, 30 to 50¢/barrel might be realized by moving the plant from some previous steady state point to the new optimal operating point. For example, our experience in computer control of commercial plants indicated that a major portion of the profit realized was attributed to activities such as steady state and dynamic optimization, modeling, improved data gathering, and improved understanding of the plant, while the remaining much smaller proportion came from improved regula-

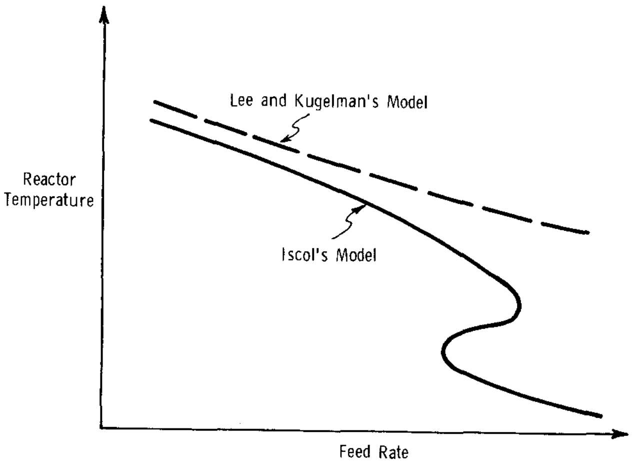

Fig. 2. Model sensitivity of steady state FCC behavior.

tory functions. This has been typical of the result experienced by many others throughout the petroleum industry (for example, Davis et al., 1974).

Of course, improved regulation by maintaining a controlled variable as close as possible to a set point sometimes can significantly increase profits as in the case of product quality control. Proper assignment of control loops, feedback, and feedforward controls have been proved useful for separation columns (Shinsky, 1974). Fisher and Seborg have studied control of a pilot plant evaporator (1973), in which they showed that advanced control methods significantly improve its performance and could be adapted to commercial plants.

Any number of factors can cause a plant optimal operating point to change; thus, variations in feedstock quality, catalyst activity, or even product value are all likely to occur. These shifts occur with time constants of from hours to days yet represent the major economic incentives for advanced control. With these time constants, time optimal control does not offer significant profit gains. The ability to track this movement of the optimum operating point requires the combination of a good process model and accurate identification of changes in feedstock composition. Identification techniques can make or break a project where feedstock changes dominate the overall economics.

Varying Constraints

Our experience in the petroleum industry also indicates that the optimal operating point commonly lies beyond the range of practical constraints. This probably occurs because of savings incorporated into the design due to capital cost considerations. Thus, a well-designed plant should operate at a constraint, or it is really overdesigned. It is also quite common to find that a particular constraint will change with feedstocks or prices. Thus, our advanced controller must not only be able to track constraints, but it must also change constraints to keep up with the optimum operating point. The real problem is not holding the plant close to some target but rather recognizing what the target is.

For both of the above considerations, it is clear that a good process model is a prerequisite for any control project. Without the capability of predicting the steady state operating point or constraints, the majority of the anticipated benefits may not be realized.

We illustrate these points further using the FCC plant as an example. The area of feasible operation is bounded by physical and mechanical limitations of the plant as shown in Figure 3. For instance, the highest reactor temperature allowable is limited by the metallurgy of the reactor, the upper limit on catalyst circulation rate by a

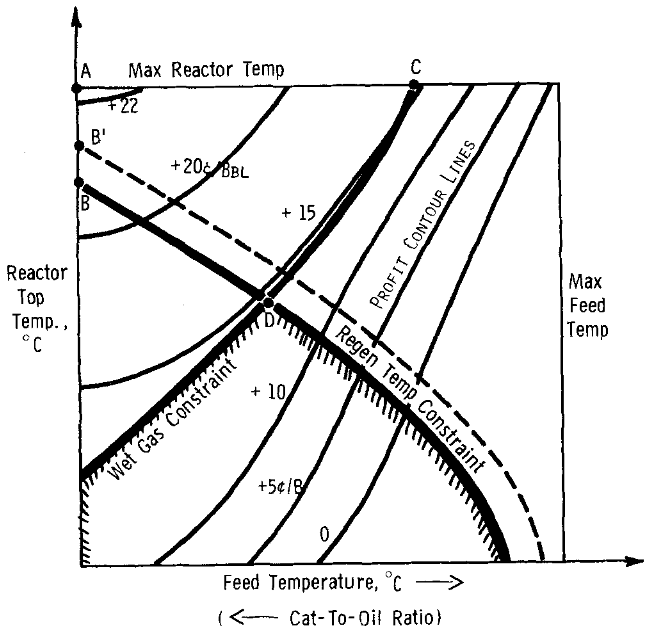

Fig. 3. An illustration of operating space of FCC plant.

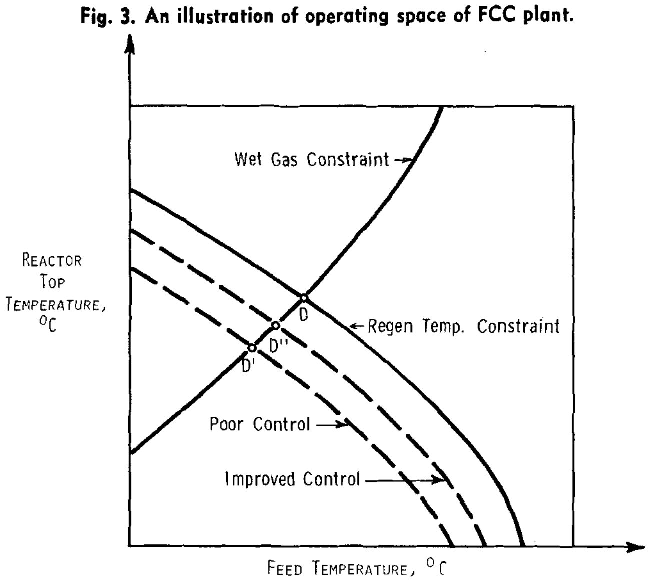

Fig. 5. Incentives for improved control of regeneration temperature of FCC.

mimimum pressure differential across the slide valve, and the lower limit by the capacity of the feed preheater for the case shown the wet gas compressor and maximum regenerator temperature constrain the operation. The area encircled by these constraints represents the feasible operating region. The contour lines of profit are superimposed over this area to show the relative profit margin at different points (Lee, 1972). Notice that profit increases strongly with the reactor temperature and decreases with the feed temperature. If there are no constraints, we would like to maintain the plant at or near the upper left corner of the map (point A). A difference of 10¢/Bbl on a typical FCC plant corresponds to approximately $5,000/day, or over $1.5 million annually. Our experience indicates that a good portion of the profit increase can be realized by steady state optimization in this case.

In reality, however, there are a number of other constraints to be considered, some of which may not be directly measurable and therefore have to be computed by using models. Typical constraints frequently encountered are those on the wet gas compressor, regenerator temperature, or air blower capacity. If the regenerator temperature becomes the limiting constraint, the best operating

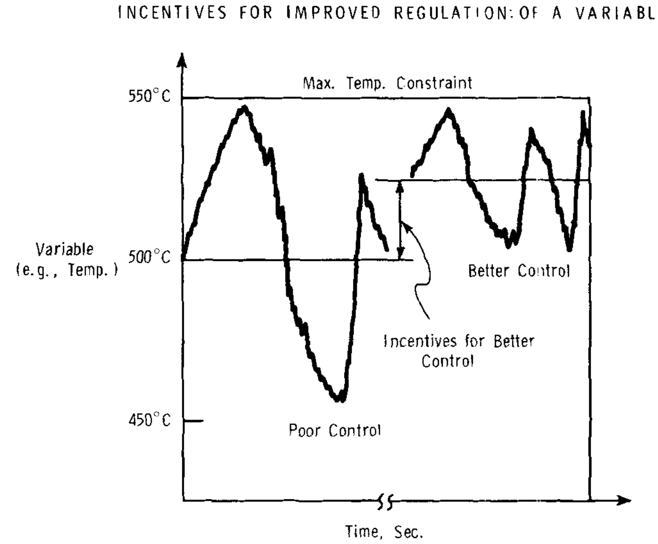

Fig. 4. Incentives for improved regulation of a variable.

point moves to a lower temperature at the maximum cat circulation (that is, maximum cat-to-oil ratio) (point B). Suppose that we are currently under the regenerator temperature limit (point B). As the feed composition gets lighter, the coke production rate decreases, resulting in a lower regenerator temperature. The regenerator temperature constraint will move up in Figure 3 as shown by the dotted line; the operation should be moved to point B′. On the other hand, if the coke production is reduced significantly (owing to an improvement in catalyst manufacturing for instance), the regenerator temperature constraint could be eliminated completely from the consideration. The next limiting constraint might be on the wet gas compressors, under which the best operation would move almost to the opposite side of the map (point C). Notice how the best operating point jumps around, depending upon the limiting constraint.

It is very common that the operation is limited by a few constraints simultaneously, such as at point D. Therefore, the major problem is to determine correctly the exact location of the intersection of the prevailing constraints and to try to maintain the plant at this point as closely as possible. As such, it is most crucial to be able to trace the shifts in constraints as process conditions change from outside influences. One can hardly overemphasize the need of reliable process models in this case.

Incentives for improving regulatory performance (that is, maintaining the controlled variable close to the set point) is illustrated in Figure 4. For instance, to maintain the controlled temperature in Figure 4 below the maximum temperature at least 95% of the time, a considerable safety margin is required when the control is poor; however, this margin can be reduced significantly as the regulation is improved. It means that on average the temperature can be pushed closer to the constraint without violating the safety requirement. For some dynamically active reactors or distillation columns, better regulation could result in a substantial improvement in profit; however, in general, for processes involving chemical reactions, it turns out to be less significant compared with what can be achieved from steady state optimization. For the FCC plant, for example, the average regenerator temperature can be moved closer to the constraint (from point D′ to D″) by improved regulation (Figure 5), but usually the magnitude of the improvement was found to be of the same magnitude as the plant noise or well within the inaccuracies of model. Therefore, it very often becomes a major problem to verify the improvement, even though the implementation has been completed successfully.

Difficulties in Implementation

Anyone who has tried to implement advanced control schemes to an existing process plant quickly learns that he has to overcome several hurdles before he can make an attempt to try them. Some of these hurdles are of human nature, others technical. First of all, we must realize that safety and continuity of operation dominate the thinking of the operating staff. The implementor will quickly be made aware that even one small shutdown of the commercial plant will make him and his scheme persona non grata forevermore at that installation. The more radical the new control scheme compared with the old, the more difficult it is to convince the operating staff of its virtue. A hundred safety reasons and other fears will be cited in direct proportion with the complexity of the new scheme. He quickly finds that he gets a more sympathetic ear if he implements his new advanced control strategy in small steps, each of which represents only a small departure from the past procedure.

Generally, the advanced control scheme to be implemented covers only a small fraction of the total operating aspects of the plant, and the operator immediately wishes to know how it integrates in with his other control actions. His concern here is legitimate, for refinery or chemical plant process units are tightly interconnected, and any changes in one of the units may greatly influence the operation of the downstream units. It is the responsibility of the designer of the advanced control scheme to make sure that it fits into the rest of the plant without conflict.

In the operation of complex chemical plants, operators are an integral part of the information flow structure and must not be excluded in the implementation of new control schemes (Rijnsdrop, 1967). The interaction between human and automatic control has been the subject of many investigators since the early sixties (for example, see Edwards and Lees, 1974). The plant operators have developed strategies which were never mentioned in the operating manual but which strongly contribute to the smooth operation of the plant. Since no control loop can anticipate all contingencies, the operator must, by necessity, be considered in the design of any advanced control system. In fact, this represents a significant departure from the control techniques employed in the aerospace industry. Here, for example, the control problems associated with the Mariner X spacecraft can be largely specified and most contingencies accounted for. This is far from true in any typical chemical plant, where disturbances in upstream equipment never cease to amaze you in their variety and frequency. The human operator, of course, has learned to live with these and has developed strategies to cope with them. The moral of the story is that any successful advanced control scheme must consider more than just those parts of the plant under its direct control and must account for the operator's normal reactions to outside disturbances.

Catalytic processes are notorious for changes in catalyst behavior, and adaptive loops are found quite commonly in such applications. The ad hoc development of an adaptive loop has probably saved more advanced control projects from failure than any other single technique. Adaptive loops, however, can lead to stability problems, especially when they are coupled with an overall control scheme which is tracking or jumping between the constraints. All of these problems are, of course, rarely addressed in the clean world of aerospace or in the academic halls.

Another common revelation to the implementor of an advanced control scheme is that if the process design engineer had had the foresight to provide a certain control valve or heat exchanger, the whole advanced control project wouldn't have been necessary. As mentioned earlier, in the design stage of a plant, capital cost savings become

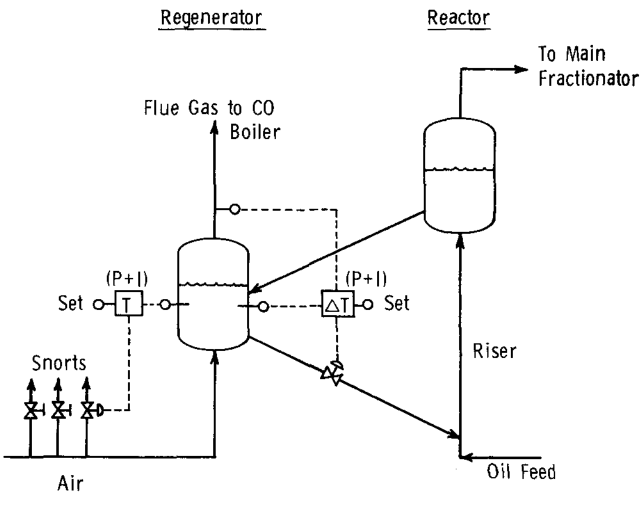

Fig. 6. Major control loops—FCC (other loops omitted for clarity).

the overriding rule, and valves, exchangers, intermediate tankage, and other equipments which vitally affect the dynamics of the operating plant are dropped out. Too many current plant designs are performed on a steady state basis with little consideration, if any, given to the interconnected dynamic performance of the overall plant. One practical cure, which we have heard by the grapevine, is that the problem would not occur if the process designers were required to operate the plant for the first year. However, since this seldom happens, dynamic simulations of the control and operation of the plant should be coupled with the steady state design to give the final plant a dynamic operation that requires less heroic advanced control.

These points will be best illustrated again with the FCC plant. Recall the conventional FCC control scheme, in which the reactor temperature is controlled by the catalyst circulation rate and the excess oxygen concentration by the air flow rate. Suppose we propose an entirely different control configuration developed by Kurihara (1967) to be implemented on a commercial unit. In this novel scheme, the oxygen concentration is controlled by the catalyst circulation rate and the regenerator temperature by the air flow rate (Figure 6). Notice that the control emphasis has been switched from the reactor to the regenerator, and that the reactor temperature is not even controlled directly in the proposed scheme. It is the most critical variable the operators watch and control closely, because it strongly affects product yields and the performance of the downstream distillation column. It will be difficult to convince the operators of this new scheme's merit so as to try it on a commercial plant, unless the operating staffs were experts on Pontryagin's maximum principle and state variable theories.

The Kurihara scheme was developed by using optimal control theory as follows. The process model developed was first simplified to allow the optimal control theory to be applied without undue complication. The optimum control policy $\underline{u}^*(t)$ was developed at a number of different initial points, and their trajectories were plotted in the state space $\underline{x}$. Then, a set of closed loop optimal control laws of the form $\underline{u}^* = \underline{h}(\underline{x}^*)$ were devised. These optimal laws were linearized around a steady state point, and insensitive terms were eliminated to arrive at the proposed control scheme. The objective function used was a gross profit rate plus two penalty functions to constrain the regenerator temperature and oxygen in the

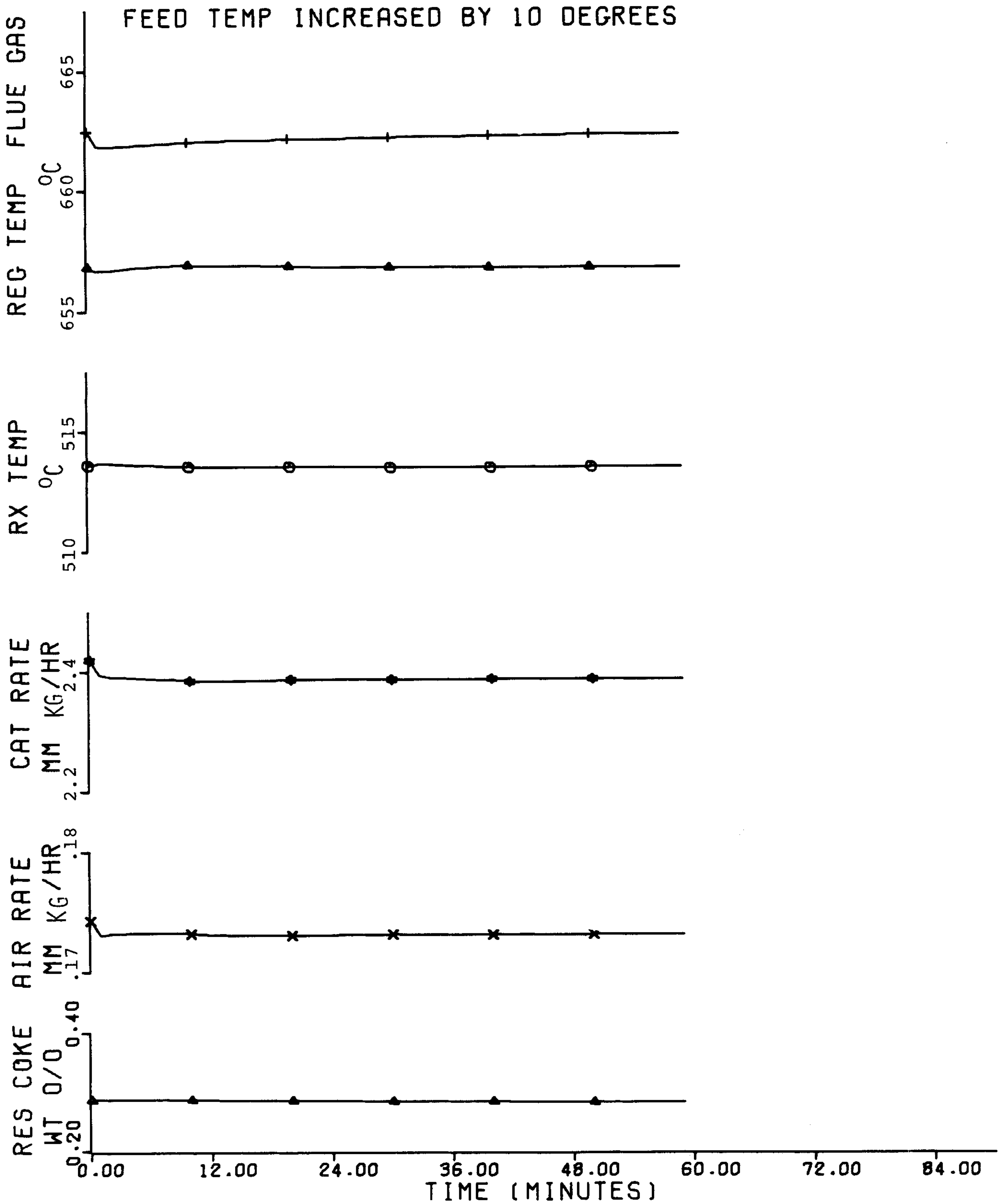

Fig. 7. Kurihara scheme.

flue gas. By dynamic simulation, he showed his scheme to be superior to the conventional scheme. However, its direct application to commercial plants in its original form seems very limited because of difficulties in implementation. The basic idea is quite sound and could find useful applications in the future if additional loops are added to give better reactor control.

It seems quite obvious, after the fact, that the optimal control laws are concentrated on the regenerator, for the FCC dynamics are dominated by the regenerator performance. The optimal control laws were also shown to reduce the positive feedback loops found in the conventional control scheme. The response of the optimal controller to a step change in the feed temperature (Figure 7)

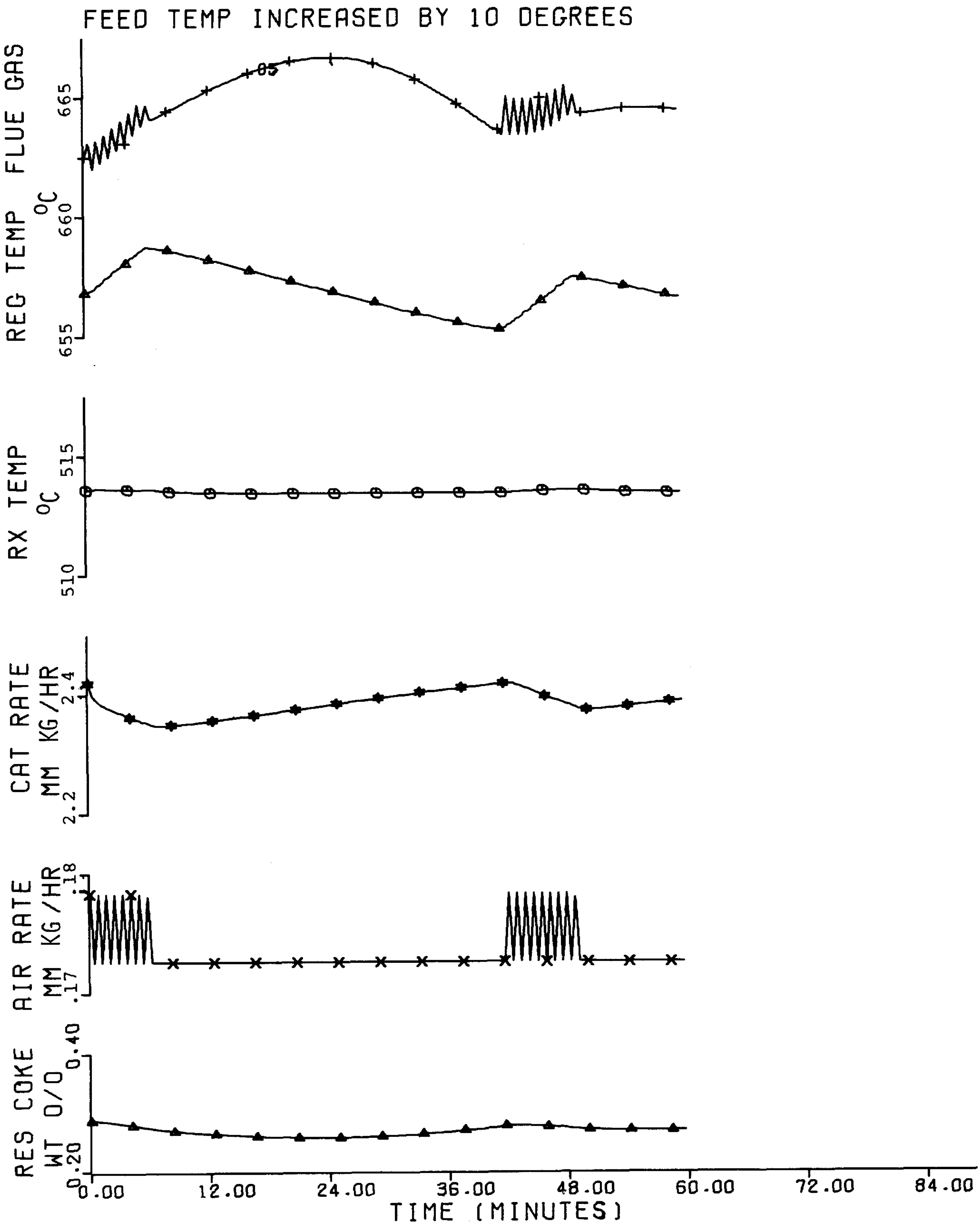

Fig. 8. Conventional scheme.

is considerably faster with less excursions compared with that of the conventional controller (Figure 8). For other typical disturbances, similar results were obtained.

There are several difficulties in the implementation of this advanced control scheme to replace an existing scheme. First of all, it is not clear how well the optimal controller will blend into the rest of the control in the plant. For instance, if it allows the reactor temperature, which is not controlled directly, to take wider swings, the downstream fractionator could be disturbed too much. This in turn will introduce further changes in the reactor and regenerator through the feedback of pressure from the compressor system of the distillation column.

Secondly, the operating staff is generally reluctant to

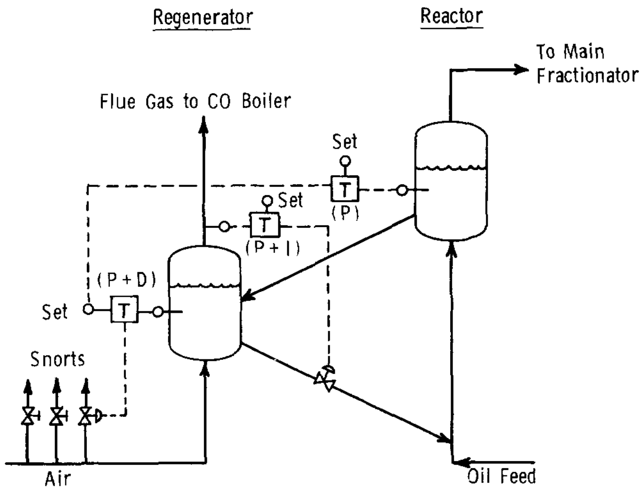

Fig. 9. Major control loops—FCC (Other loops omitted for clarity). Modified Control Scheme

try a new scheme unless the safety of plant operation is assured and the incentives are substantial. As mentioned before, the operator is an integral part of the control loop in the process industry, and the optimal control did not account for him. He has to modify his roles and strategies to accommodate the new control, which might take many months of trial and error.

One could devise many other competing configurations which might be slightly inferior to the optimal control in regulatory functions but might be more satisfactory to the operating staff. An example of such a modified scheme is illustrated in Figure 9 (U.S. Patent 3,753,893). Two modifications from Kurihara's scheme were incorporated in Figure 9. First, a cascade control was added to automatically adjust the regenerator set point as the reactor top temperature varied. This gave the operator the flexibility of changing the distribution of temperature excursions between the two vessels. Secondly, the flue gas temperature, instead of the temperature difference between the flue gas and the dense bed, was used to control the catalyst circulation rate. This was designed to eliminate the difficulties associated with the phase shift between the flue gas and dense bed temperatures. Dynamic simulations of this modified scheme were found as good as the basic Kurihara scheme or even better for certain types of disturbance. For instance, when a step change of a 6% decrease in the coking rate of feedstock was introduced, the basic Kurihara scheme (Figure 10) showed significantly larger excursion and took longer time to arrive at the new steady state than the modified scheme (Figure 11). The advantage of this modified scheme over Kurihara's is that it better accommodates the operator's concerns and passes less disturbances to downstream equipment.

Verification of Results

The problem of sparse and poor measurements frequently aggravates the chance of successful implementation because it is difficult to quantify the improvement accomplished. Sometimes one concludes after limited tests on a commercial plant that a proposed control is satisfactory by demonstrating its stability and the proper directionality of control efforts. Process models can be very valuable in defining the magnitude of the improvement.

Start-ups and Shutdowns

Commercial processes generally experience the most extreme operation during the transient period of start-up and shutdown. Many hours of preparation and lining out time are required for a commercial unit to be ready for operation. During this period and immediately thereafter, the unit may experience extreme variations in process variables. Currently, the majority of complex processes are placed on manual control during start-up and shutdown. This is because conventional controllers can neither operate satisfactorily for the extreme conditions nor account for various unforeseen contingencies. In addition, most process variable changes exceed the range of their controllers. Therefore, the operators are primarily responsible for establishing normal operation without undue safety hazard. Any new control scheme to be implemented must also be examined with respect to the control of start-up and shutdown procedures.

SOME OBSERVATIONS ON ADVANCED CONTROL THEORY

As practitioners, we may not be in the best position to critically review the inadequacies of the existing control theories and will leave the main burden to our academic colleagues. However, we will briefly express our sentiment about the usefulness of some of the existing control theory. As eloquently expressed by Foss (1973) and many others, the existing advanced theory of the process control assigns emphasis on the wrong problem. The ubiquitous minimum-time, minimum-fuel consumption problem in the aerospace industry is very rare in the process industry; nevertheless, this type of problem has been treated with a top priority in the process control literature. Recall that it is not our major problem to track the optimum trajectories in minimum time but to safely maintain the process near a desired point without undue disturbance.

Adaptive control techniques have been widely used in the machine tool manufacturing, aircraft landing, and aerospace fields. In the most popular, so-called model-reference approach, one compares the present position with the corresponding ideal position derived from the perfect model and tries to control the trajectories as perfectly as possible. Here, an ideal, perfect model of the system can be developed a priori, whereas it is generally an impossibility in the process industry. Modeling of chemical processes with most of their state variables unknown or not measurable requires more than what can be offered by Luenberger's observer theory. Precise knowledge of the process, noise-free measurements, values of the parameters, and methods for reducing order of the observer are all prerequisite for efficient and practical implementation.

The optimal control technique with a linear quadratic performance index has received considerable attention in the control literature and has been suggested as a solution to the problem of regulating an industrial plant in the vicinity of a steady state point (for example, Athans and Falb, 1966; Lapidus and Luus, 1967). Although it has certain theoretical properties of merit, as pointed out by Edgar et al. (1973), its application to the process industry has been limited thus far. In addition to the difficulties of phase shift, noise amplification, or conditional stability discussed widely (for example, Rosenbrock and McMorran, 1971), the fact that most of the state variables are not measurable makes the linear quadratic Gaussian approach (Athans, 1971) impractical for the majority of problems in the process industry. Interestingly, a quadratic performance synthesis technique has been applied to the FCC process, and the simulation of resulting control was shown, as expected, to be superior to the conventional single loop analogue control (Schuldt and Smith, 1971). Using an extremely simplified model and only several measurable state variables, they tested eight multivariable control schemes based on a priori feedback relationship. In spite of the improvement demonstrated by the dynamic

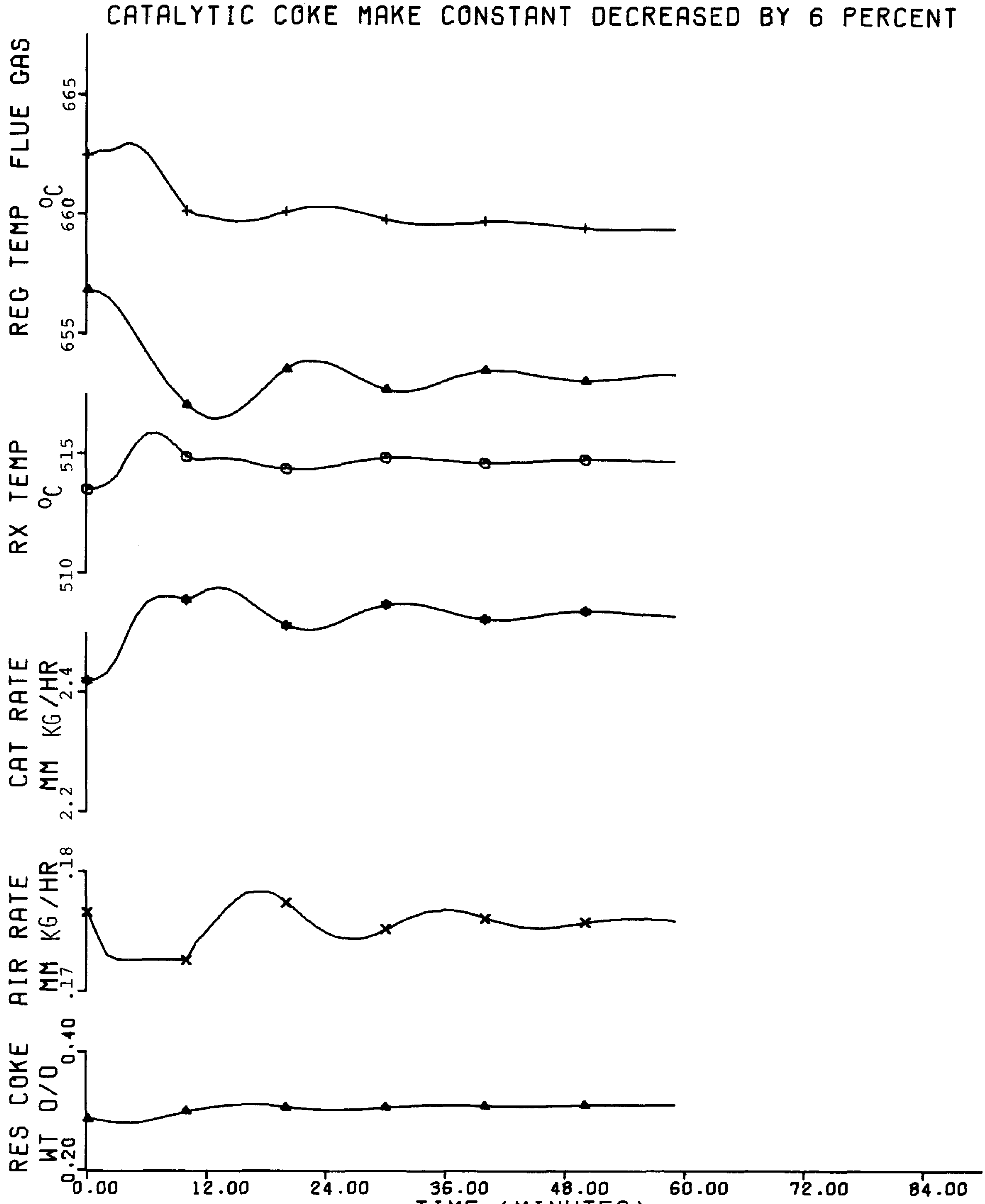

Fig. 10. Kurihara scheme.

simulation, this control scheme has not yet been widely used. Reasons for its limited application might be additional capital investment for a process computer to implement multivariable control, questions on the conditional stability when one of the measurements fails, uncertainties related to the unrealistically simplified models and objective function, and finally limited economic incentives and extensive effort required to implement it in the place of an existing PID control.

The stability problem associated with parameter adaptation for chemical processes has not been fully resolved. The magnitude of slow drifting disturbances as well as uncharacterized noises relative to that of control action is manyfold larger than those encountered in the aerospace

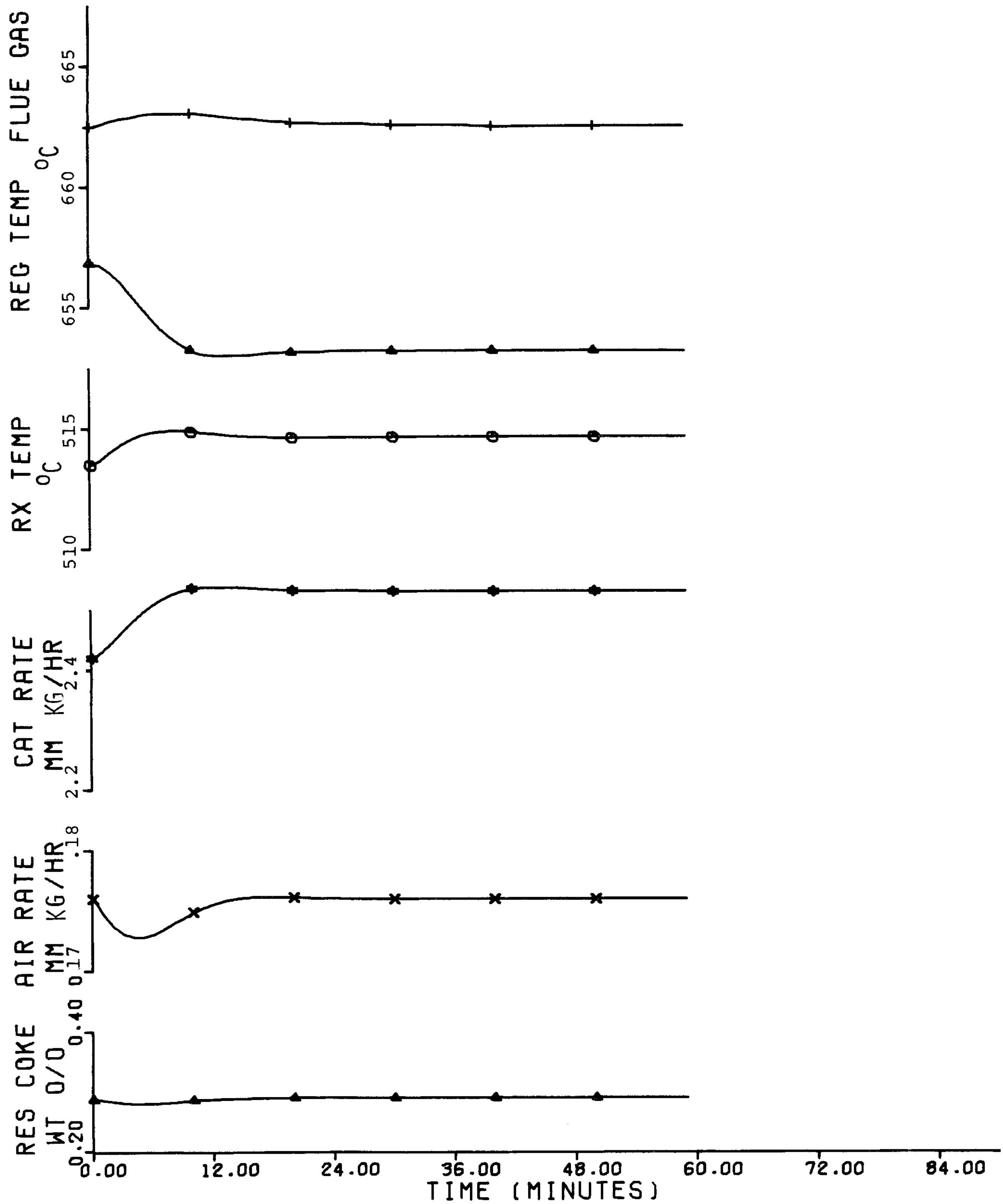

Fig. 11—New scheme catalytic coke make constant decreased by 6%.

industry. This gives rise to the instability problem when parameters of models or controllers are changed significantly in the adaptive procedure.

Practical methods for designing control structures for multidimensional interacting processes are badly needed. Theories on modal control and noninteracting control have been developed, but only in piecemeal fashion. However, attempts have been made to apply the technique of modal analysis (Rosenbrock, 1962) to determine the control system structure of multivariable systems (for example, Davison and Chadha, 1972; Davison and Goldberg, 1969; Gould, 1969, Chapter 7). In principle one can force the slowest eigenvalues to be large negative values by using the correct amount of negative feedback, thus achieving rapid and stable recovery from small disturbances. In practice the restrictions imposed by the order of the system, ability to measure only a very few states, number of control variables available, and accuracy of measurement, can cause serious deficiencies for large chemical plants. It is generally not the knowledge of the modal character of the process that determines the choice of measured and manipulated variables, but considerations of economics and physical limitations.

SOME NEW DIRECTIONS

A good theory must be useful to process control engineers and should be developed to accommodate the needs and skills of the potential users. It should not start from what skills the academicians happen to have or can borrow from other disciplines, but from the major needs of the end user. Indeed, the user and the theoretical developer should cooperate very closely: listening to each others problems, observing the process operation, and getting the user involved in the development effort. Without this, the theory will remain as an intellectual game and fail to fulfill its expectation. It was once said that, "Only Frank Lloyd Wright can design a house for a family without asking about the number of children, or the family budget."

First, the process design and control design have to be integrated so that the dynamics and control configuration could be considered in the process design stage. New techniques must acknowledge the problems of sparse and poor measurement, imprecise process knowledge, and computational difficulties in parameter estimation. Perhaps techniques such as the modal analysis of Rosenbrock could be specifically tailored to alleviate the critical problem arising from sparse measurement (Gould, 1969, Chapter 7; Rosenbrock, 1970).

A high priority should be assigned to new techniques designed to aid modeling of chemical processes. Drastic improvements in measurement techniques and estimation of parameters with unknown control structure are badly needed. For example, in the petroleum industry, on-line analytical techniques for determining chemical compositions of heavy fractions, catalyst activity, and coke deposit on catalyst will tremendously improve the accuracy of process models and the usefulness of advanced control. For additional areas which warrant increased research emphasis in modeling, refer to a recent review by Weekman (1974) on the state of the art of process modeling.

Methods for characterization of disturbance, to identify the process dynamics have to be improved. Filtering and prediction theories developed by Wiener-Kolmogrov or Kalman-Bucy can be inhibited by the requirement of extensive numerical computation and an accurate model. However, application of covariance functions have found to be successful in many instances enhancing the quality of model produced. A number of identification programs (software) as well as instruments (hardware) are currently available in the marketplace (Aiken, 1975). The recent advent of inexpensive microcomputers looks promising in providing better measurement and control in the future. Stability problems of adaptive control must be resolved. In the face of constant noise and uncharacterized disturbances, stability of the closed loop adaptive system is required over a wide range of frequencies to assure safe operation. It is also important to be able to smoothly change to manual operation and vice versa.

Finally, what we need is ease of implementation. There should be a unified, systematic procedure for resolving the difficulties encountered in the implementation of advanced control. A new control scheme has to be integrated smoothly with the operator as well as other controls of the plant. Perhaps, we might need a staged approach in which it is gradually implemented step by step.

CONCLUSION

We have commented on the status of the advanced control practice in the chemical process industry. Emphasis has been given to the difficulties frequently encountered in the implementation of advanced control schemes to chemical reaction processes. Major reasons for its limited success to date were identified as relatively small economic incentives for improving regulatory function compared with steady state optimization or design improvement, our limited ability to measure process variables to accurately characterize the process, difficulties in integrating into other controls of the plant and the operators, and, finally, inadequacies of the existing techniques originally designed for different problems in other fields. It is not our intention to disparage optimal control theory but to point out the practical limitations in the implementation of advanced control theory to the chemical process industry. We are not suggesting that our academic friends solve industrial problems for us. But we do suggest that they concentrate on developing new techniques which are badly needed to overcome the difficulties discussed heretofore. We believe that it is high time to reevaluate our present approach to process control and try to develop sets of tools specifically designed for the problems of the chemical process industry.

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Combined the authors have accumulated more than two decades of industrial experiences in the process control and reaction kinetic modeling of petroleum processes. They report that their efforts in the process control area have been successful and rewarding, but tempered with the occasional blunting of their swords in the implementation of advanced control strategies, the latter of which is emphasized intentionally in the paper to expose some of their anxieties regarding advanced control techniques. The present paper was originally prepared as a plenary lecture at the 1974 JACC.

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Experimental Investigation of Models for Fluidized Bed Catalytic Reactors

The countercurrent backmixing model of a fluidized bed reactor predicts axial concentration profiles quite different from those suggested by simple two-phase models. The models can also be distinguished in terms of the dependence of conversion on operating variables.

An experimental study of ozone decomposition in a reactor of 22.9 cm diameter has provided extensive data for comparison with backmixing and two-phase models which incorporate bubble size variation. The measured profiles show a minimum concentration within the bed at gas velocities above a critical value, as predicted only by the backmixing model. The effect of operating variables on the shape of the profiles is also well accounted for by this model. The backmixing model is further supported by good agreement between predicted and measured reactant conversion. In particular, the variation of conversion with rate constant and gas velocity is fitted more accurately by the backmixing model than by two-phase models.

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SCOPE

The countercurrent backmixing model, proposed by several investigators, attempts to account for gas mixing behavior in bubbling gas-fluidized beds of fine particles. The model predicts downflow of gas with solids in the particulate phase at fluidizing gas velocities above some critical value. The earlier and simpler two-phase models of Davidson and Harrison (1963) assume either that gas in the particulate phase is completely mixed or that gas flows upward in plug flow through the particulate phase.

Previous experimental studies have shown that the backmixing model provides a good description of mixing of a tracer gas injected just under the surface of a fluidized bed. Available data on fluidized bed reactor performance are not suitable for testing the validity of the