Critique of Chemical Process Control Theory

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When it is stated, as it has been in more than one recent publication, that there is a wide gap between the theory of process control and its application, one is left with the unmistakable impression that those who conceive the theory are in some sense leagues ahead of those who would use it. That the contrary is the case is the thesis of this essay. Indeed, the theory of chemical process control has some rugged terrain to traverse before it meets the needs of those who would apply it.

PROBLEMS OF CHEMICAL PROCESS CONTROL

The needs are intimately related to the problems, and the problems, as usual, wear a sometimes effective camouflage. Superficially the problem of chemical process control appears to involve the regulation of complex, often poorly understood physicochemical processes in the face of many unknown and uncharacterized disturbances. The word *regulation* has several interpretations when applied to industrial processes, but it usually implies the desire to hold constant certain states or perhaps a time-average of the states. Disturbances may consist of slow drifts such as those originating in diurnal temperature changes, of monadnocks illustrated by the explosive vaporization of a slug of water in the feed to a crude still, or of persistent random upsets such as the step-like fluctuation in pressure in utility headers. Sometimes disturbances result from a purposeful change in the level of operation, but with the exception of batch operations, these changes are infrequent. In addition to the continuous regulation tasks, there are the tasks of start-up and shut-down. In all of these activities, there is the dominating necessity for safe operation in the event of malfunction or failure of any part of the process or control system.

But there is much beneath the surface. One finds that the processes possess dynamic components and that in the design of control systems attention must be given to this characteristic. In one view the dynamic characteristics are seen to affect adversely the controllability of the process owing to lagging of process states to corrective commands. In another view the dynamic characteristics are welcomed as fortunate endowments that lend a degree of stability to processes and that sometimes can be exploited for useful purposes. And because the dynamics of a process are directly influenced by its design, the control system designer finds that his sphere of responsibility encompasses process design as well. Indeed, major contributions to effective control system performance often derive from perceptive and clever modifications of the process itself.

But perception is difficult to acquire in this field because the dynamic behavior of chemical processes is not simple. There are many variables whose dynamic behavior is of consequence. The behavior of any one variable is influenced by many others through the innumerable physical and chemical interactions found in these processes. The interactions in a modern methanol process, for instance, are complex enough to tax one's ability to untangle the relationships even at steady conditions (Shah and Stillman, 1970). The inert level in the synthesis loop may be regulated by manipulation of the purge rate, but since the purge constitutes a large fraction of the reformer fuel flow, one finds that such an adjustment influences the temperature and methane content of the reformer process effluent stream. Steam production in the waste heat boilers is thus upset and so too are the compressors driven by that steam. The compressors are also upset by the change in recycle rate in the synthesis loop. The resultant flow changes influence temperature rises in the synthesis reactor, leading perhaps to higher temperatures at which the side reactions that produce ether and methane contribute significantly to the total heat liberation and to reduction in methanol production. Further, the upsets in the reformer lead to concentration disturbances in the synthesis loop makeup gas, which also disturbs the methanol production rate.

Now add to these static interactions the influence of a 20-minute thermal lag in the reformer, the composition and thermal lags in the carbon dioxide absorber-stripper system, the treacherous feed-effluent heat exchange in a synthesis reactor that can exhibit wrong-way (nonminimum phase) temperature effects, and the long-lived composition transient in the synthesis loop. Even were the process well understood, the dynamic cause-effect relationships would be difficult to untangle.

But such processes are not completely understood. The coke level on the reformer catalyst is likely unknown, different from furnace to furnace, and drifting; the firing distribution of the burners in the reformer changes in an undefined way with fuel flow changes; and the carbon dioxide separation efficiency is a poorly understood function of changing flows and absorbent concentrations. Uncertainties in other processes are, for example, the course and rate of the solid reactions in cement kilns, the reactions and their dependence on the hydrodynamics in multiphase hydrocracking reactors, and the dependence of conversion on the flow regime in fluid bed reactors. And it is well recognized that no amount of detailed study will ever replace all uncertainties with certainties. Rather, it is for the control system designer to recognize the significant uncertainties and to conceive controls that function effectively nonetheless.

While it is the presence of coupling among many variables that is primarily responsible for the near inscrutable complexity of dynamic processes, the nature of the coupling as well often plays a significant role. By nature and by design the coupling among most variables in chemical processes is nonlinear: temperature-reaction rate, feed enthalpy-vapor/liquid split, temperature-equilibrium conversion, boilup rate-overhead concentration, and on and on. Such effects often need direct attention; indeed, the stability of exothermic chemical reactors may be assured only through consideration of the nonlinear dependence of the chemical reaction rate

on temperature and concentration. Thus, the nonlinearities and the intricate multiple interactions combine to form a formidable web of interrelations, the perception and understanding of which are often instrumental in solving process control problems.

One also finds that practical control systems require measurement of process variables but that measurements can be made of only a very small fraction of the variables. The measurement of some variables is practically impossible, of others very costly. All measurements are subject to various random and systematic errors and misinterpretation through confounding of unaccounted effects. Oftentimes the measurements are long delayed owing to the complexity of the chemical or physical analyses. The lack of measurements, the errors, and the delays all adversely affect the quality of control. Measurement delays, for example, adversely influence the stability of controlled processes.

Yet the problems of control go still deeper. One is eventually forced to make both qualitative and quantitative decisions about the controlled system. One of the most important decisions concerns control objectives. Regulation of certain process variables at desired levels is by far the most common objective in continuous chemical process systems. All measured variables need not be regulated, however. In some instances regulation is made of functions of the state variables, as in the case of the control of internal reflux and feed enthalpy of distillation columns. Clearly, the choice of the process variables to be controlled is a crucial one. Once decided, does one demand invariance of the chosen controlled variables or something more realistic such as a minimum weighted deviation of the variables? That is, exactly what is meant by regulation, and how is one to distinguish good regulation from poor?

A definite, quantitative measure is required. In fact, several measures are frequently needed; certainly in systems as complex as industrial chemical processes, one cannot expect a single criterion to suffice. It can be expected that some of the criteria will be affected by control action in conflicting ways and that the best control results from a compromise. By choice of the functional form for each element of the performance vector, one ultimately defines what is meant by quality of control. These can include a profit index, a peak excursion of variables, an integral squared error, a state- and control-weighted quadratic index, a settling time, time itself, and many others. Obviously, each of these criteria reflects different measures of control performance, and one of the designer's tasks is to decide which is most suitable.

It is by no means obvious how to achieve the best control. In addition to the possibility of making modifications to the process, there is the crucial step of conception (or invention) of the control system configuration. Which variables should be measured, which inputs should be manipulated, and what links should be made between these two sets? This problem is considered by many to be the most important problem encountered by designers of chemical process control systems. It is certainly the most prevalent. With only a handful of measurements and manipulatable inputs at his disposal, the designer must somehow discover the control links that effectively and economically meet his control objectives within the constraints of physical realizability, safety, and stability. Such a problem is obviously extremely difficult. It has always been tackled in an almost wholly qualitative manner and with considerable reliance on the successful control configurations of the past.

The information needed to make decisions about control system design consists of both static and dynamic characteristics of the process. While the design of many control system has been based only on the static characteristics, it is known that such procedures are in general insufficient. Dynamic characteristics, of course, are necessary for rational design. However, the quantitative characterization of chemical process dynamics can be so difficult and time-consuming that considerable judgment is required concerning the extent and form of the information to be developed. A complete process description is impractical and in the end, unnecessary; only the dominant dynamic elements need be included in process models. Forms of the models range from sets of nonlinear differential equations to empirically or experimentally derived transfer functions. The forms of the models may not be selected arbitrarily; they are determined in part by the control objectives and the type of control analysis to be pursued. In short, process modeling is a substantial and crucial task, and by no means routine.

The problems of control system operation, while related to those of design, differ in flavor. Drifting process conditions adversely influence control system performance; there is thus the obvious need to adapt control system parameters to values appropriate for the current conditions. Some efficient method of parameter estimation is needed to accomplish this. For the nonlinear multiparameter processes and the multiloop control systems of the process industries, practicable adaptive systems are difficult to conceive and difficult to operate reliably. The operation of control systems of modern design also requires estimates of the process states used for control. This requires a process model, perhaps different from that used for design calculations, and a means of rapid solution of the model equations.

The large dimensionality of the processes, their strongly interactive nature, their poorly known characteristics, the ability to make only very limited measurements, and the undetermined control system structure place the process control problem in a unique position in the spectrum of all control problems. The uniqueness is further marked by the potential to make process modifications for the purpose of achieving control objectives. These characteristics not only place the process control problem in a unique position, but also greatly complicate the well-recognized tasks of process modeling and the estimation of states and parameters.

AVAILABLE THEORIES AND THEIR INADEQUACIES

With such a wide spectrum of problems to be addressed, it is no surprise that there is no single all-embracing theory to cope with them all. Various elements of the theory, however, purport to handle problems of limited scope, and it is the utility of these elements that is examined here.

The process control systems of a decade ago and still most today typically consist of more or less isolated single loops and sometimes the two nested loops of cascade systems. These single-input-single-output systems are handily analyzed and designed by the linear design technique that evolved from servo-mechanism theory. The rationale of the theory and the methods of application to chemical processes are clearly and succinctly presented by Gould (1969). The use of a linear theory on inherently nonlinear processes is not a restriction of significant consequence in the great majority of cases. Linearized models of processes are found quite adequate in their representation of the important poles, zeros, and delays at the nominal conditions of operation. Such dynamic process

parameters are sufficient to determine the design parameters of a controller (usually of specified type) and its anticipated performance. High attenuation of disturbances and rapid recovery from upsets exemplify perfomance criteria sought in the design, and these may be achieved by maximizing the gain-bandwidth product. The concepts are elementary and the calculations simple; indeed, an experienced designer can determine controller parameters within acceptable accuracy in a few minutes by hand calculation.

There is little doubt that this classical theory is useful for single-loop systems. Some criticize it for not yielding an optimal control and for its inability to handle nonlinearities. These criticisms fade to insignificance when one discovers its almost total inability to guide control system design for interacting multivariable processes. The control of both top and bottom compositions of a distillation column is an oft-cited example of control difficulties in interacting chemical processes. While the classical theory is capable of analyzing the difficulty, it has nothing to suggest as a remedy. The seat of its impotence is its reliance on the designer to specify the control configuration.

In view of the inadequacy of single-loop methods in the treatment of dynamically interacting processes, one might be tempted to turn to the theory of noninteracting control (for example, Gould 1969, p. 101). In this technique, a design is sought such that a change in set point of one variable influences that variable only. This is attempted by choice of a control matrix such that its product with the process transfer fuction matrix is diagonal. With the diagonalization achieved, the various control loops are noninteracting, and single-loop theory may then be used to design each loop separately. A control system with noninteracting properties, if they can be achieved, may be desirable for aircraft but not for chemical processes. There is no compelling reason to impose such servo performance on chemical processes when it is the diminution of disturbances that is important. Nor does it follow that process systems should be made noninteracting simply because there is difficulty in designing multivariable control systems by the single-loop methods. Indeed, the point has been made by many that interaction among variables should instead be exploited to achieve control. In attempting to design noninteracting controls, one finds that the diagonalization restricts the form of the compensation that can be applied to the single loops. Further, when the determinant of the process transfer function matrix has right-half-plane zeros, as is frequently the case in multivariable chemical processes, the technique yields poor or unstable performance (MacFarlane, 1970). Such difficulties have been encountered in recent attempts to apply the method to process control problems. With these very severe shortcomings evident, noninteracting control is not worthy of consideration for chemical processes.

The exploitation of interactions inherent in multivariable physical processes "takes judgment, brains, and maturity" and a good theory, but that last is elusive. The theory of modal control introduced by Rosenbrock a decade ago (1962), proposes that the rate of response of the natural modes (the eigenvector-eigenvalue pairs of the state matrix) be increased to an acceptable level by positioning the mapping of the input control vector in the subspace spanned by the eigenvectors of the controlled modes. When this can be accomplished and when the activation of the modes can be measured, the eigenvalues (or time constants) of the closed-loop process can be adjusted independently through a proportional feedback control matrix. In principle, very rapid and stable recovery can be achieved. Modal control is thus a technique that exploits the interactions among variables to achieve control objectives. Unfortunately, and not unexpectedly, such ideal circumstances do not obtain in chemical process systems. The ability to measure only a very few of the states results in confounded estimates of modal activations, and the handful of process inputs in general cannot be aligned with process modes. Both of these facts of life can cause serious deficiencies in control systems designed by the modal techniques. Nevertheless, attempts have been made to apply the idea to the control of diffusive distributed processes (Gould, 1966) and to a distillation column, a nuclear reactor, a boiler, and a 41-variable chemical plant model (Davison, 1967, 1969, 1972). In these investigations knowledge of the modal character of the process was used to suggest the choice of measured and manipulated variables, that is, a control configuration. While there remain potentially useful developments of modal control to assist in the determination of control system configuration, there are at the same time several disadvantages in this idea (MacFarlane, 1970). The method gives attention only to the poles; the closed-loop zeros, which also influence system behavior, are left unattended with the possibility that some will appear in the right-half-plane. Further, the method at present does not address the regulation of specific output variables. Both of these are serious detractions.

The theory of optimal control would seem to promise, by its very name, a best compromise among all the conflicting requirements of process control systems. But it doesn't. Indeed, as Rosenbrock and McMorran (1971) wryly remark, "the word optimal carries with it the suggestion that the system has desirable properties in general, but of course this need not be the case." There are indeed many undesirable properties and unworkable features of the theory as it has been developed for chemical processes.

Optimal control, defined in various ways, has been attempted of stirred tank reactors, tubular processes, distillation columns, extraction columns, absorption columns, and the second-order-plus-delay process. Objectives in these attempts have been to determine (open-loop) manipulatable input histories that minimize the time required to bring the process to a new state or the (closed-loop) feedback and feedforward gains that minimize a quadratic functional of process states and control inputs (or the expected value of the function). Other objectives such as maximization of reaction yields have also been used.

Upper and lower bounds on the control input are imposed in the time optimal case with the result that the optimal inputs may reside at one or the other bound for all or part of the time (so called bang-bang control). The switch from one bound to the other is determined by switching curves that are calculated from the necessary conditions for minimum time, a calculation involving the integration of the state and adjoint differential equations with split boundary conditions. The calculations are difficult and lengthy. The results of such calculations consist simply of a program for the optimal input (and the resulting process states and the minimum time as well). In all of this, of course, one assumes perfect knowledge of the process model and parameters and that there are no process disturbances. In view of the unlikely attainment of these assumptions and the infrequency of purposeful state changes, startups, or shutdowns in continuous processes, it is difficult to appreciate the utility of such a theory for the control of chemical processes. The effort that chemical engineers have recently poured into the development of algorithms for the computation of this type of optimal control is also difficult to understand.

Perhaps the usefulness of such a theory can be argued in some special cases, but in the light of the circumstances encountered in the process industries, its quantitative value is slim for the processes and situations so far treated in the chemical engineering literature.

Optimal feedback control theory, viewed with respect to the applications to chemical processes so far treated, is no less immune from criticism. The objective function, consisting usually of a weighted sum of squares of state excursions and control effort is arbitrary and often partly fictitious; the resulting optimal control determined in such circumstances is therefore arbitrary optimal. Chemical engineers, in the scores of papers they have published on this topic, have demanded with few exceptions that all the state variables be measured and be measured without error. That is perplexing indeed. As a result, the very considerable effort on optimal feedback process control has yielded no conceptual foundations for the synthesis of control configurations with sparse measurements, guidance so desperately needed in the process industries.

When all states are not available, one might consider their determination by use of a Luenberger observer, but such a device introduces considerable phase advance (McFarlane, 1970; Rosenbrock and McMorran, 1971) and ultimately confronts the designer with noise amplification in practical cases. Further, the order of the observer will be large for large processes and hence difficult, costly, and impractical to implement. Measurement noise can be taken into account in a direct way through the use of the Kalman-Bucy state estimator, but the order of the estimator is often prohibitive, being equal to the process order. Moreover, the effectiveness of such observers and state estimators depends on the precision of one's knowledge of the process model and its parameters and one's ability to specify covariance matrices for the noise. These latter are difficult to determine and are often fictitious. But this does not mean that the concept of state estimation is without merit. There is indeed considerable merit in the idea, and the linear-quadratic-gaussian stochastic design approach (Athans, 1971) that combines state estimation with optimal deterministic control can give useful guidance in process control system design. But no definitive development and use of such ideas for the solution of the special problems of process control have yet appeared. The matter of unknown parameters has been recognized and there have been several attempts to identify such parameters through the estimation of a state vector augmented by the parameters, but this approach seems inappropriate for the process industries where the bandwidth for parameter variation is usually considerably narrower than that for the states. Even more disappointing is the almost total neglect of the identification of parameters in processes under feedback control. Despite the recent efforts in state estimation, there is still no generally applicable method for deriving practicable, low-order estimators that account for imprecise knowledge of the process and disturbance characteristics.

In addition to these detractions, there exists the possibility of system instability when some gains of the optimal controller are reduced to zero (Rosenbrock and McMorran, 1971). Such a consideration is of vital importance in the process industries where malfunction of measurement devices, transducers, actuators, and human operators is not unknown.

The conclusion is inescapable that chemical engineers have been working on the wrong problems; hardly a dent has been made in the significant fundamental problems of chemical process control. Somehow, the incisive questions have not been asked, or if asked, they have fallen on unhearing or uninterested ears. Instead, there has been a puzzling infatuation with the nearly direct translation of methods developed elsewhere to control problems that have been represented as those encountered in chemical processes. It has been argued here that the problems, needs, and objectives of process control differ so significantly from those in other fields that these translations have little practical value. Even the processes that have been selected to illustrate and to test the translations are largely unconvincing in appropriateness. The stirred-tank reactor is usually taken only as a two-variable process, and the second-order-plus-delay process has only a single input and a single output. In the testing of methods for distributed systems, the plug-flow, single-variable heat exchanger has been the overwhelming favorite. It is certainly necessary that proposed methods perform favorably for such simple processes, but such processes by no means offer a sufficient test. The two-variable reactor does not reveal, for example, the deterioration in control that accrues when only a small fraction of the state variables of a large process is measured. The heat exchanger lacks the coupling among variables and the dynamic elements that give rise to nonminimum phase behavior and to disturbance propagation velocities different from that of the fluid. Nor do applications of control methods to these one- and two-dimensional processes reveal clearly the large computational effort that would be encountered in processes whose size and complexity are representative of industrial processes. The gap is wide indeed but, lamentably, inverted.

UNWRITTEN THEORIES

But despair not; there are signs of life. A few jaunty explorers, somehow loosened from the hypnosis of the popular sterile exercises, have been scouting new routes through the foothills. They have dared to take on the problem of control system structure, have seen that low-order models are necessary, and have recognized that a practical method of state and parameter estimation is one of the keys to progress. There is also an unmistakable trend to reexamine and extend the classical methods, as so often occurs in science when inadequacies of the popular methods become apparent. Recent publications (for example, Rosenbrock, 1970, 1971) reflect this trend.

Perhaps the central issue to be resolved by the new theories of chemical process control is the determination of control system structure. Practicable solutions to this problem are not directly forthcoming from the current methods despite the views of some that the Kalman filter has completely solved the configuration problem for linear processes. The problem is tougher than that and not so easily cracked; it will require attack from several fronts as Rosenbrock assuredly knew when he tried to interest chemical engineers in the challenge a decade ago (1962). By definition, the problem involves the consideration of processes having many interacting variables, perhaps two or three score in an initial count. The methanol process has at least that many primitive variables, and it is a formidable task to sift from among these those that should be measured and manipulated and to determine the control connections among them. Should one measure the concentration of carbon monoxide? If so, where? Can one measure it with sufficient accuracy? Or perhaps it is the ratio CO_2/CO that should be measured. Or how about the ratio CO/H_2? Is the manipulation of the carbon dioxide feed to the reformer effective for the regulation of the CO/H_2 ratio in the synthesis loop? Such are the questions that need answers, and it is the burden of the new theories to invent ways both of asking and answering

the questions in an efficient and organized manner. The prospects of tackling problems of such high dimensionality will likely deter those who in the past have allowed themselves to be turned back by the curse of dimensionality. But dimensionality is a curse only for the analyst; for the inventor and engineer, it is a blessing. Indeed, the foundations of the chemical industry rest on the intricate and ingenious exploitation of interactions among many physicochemical effects. Be that heritage not forsaken.

The problem of control system configuration is pressing, and there is some danger that Rosenbrock's query of a decade ago will remain unanswered, as is the trumpet in Charles Ives' composition "The Unanswered Question." As in that piece, there are a few voices from the hinterlands struggling for an answer but receiving no assistance from the sciences who, in their preoccupation with accepted matters, have no ears for such ill-posed questions. Voices such as Gould (1966; 1969, Ch. 7) and Davison (1967, 1969, 1972) reveal attempts to couch the problem in terms of the modal characteristics of the process, as originally suggested by Rosenbrock. Synthesis of the configuration through use of the static input-output characteristics of the process has been explored by Bristol (1966) and Weber and Brosilow (1972), the latter including also the effects of measurement errors. However, an approach that does not include the process dynamics will not suffice. Niederlinski (1971) proposes that control links be selected from a multivariable generalization of the single loop gain-bandwidth criterion. None of these can be considered to be anything more than exploratory forays. But the response of the distant woodwinds, however imperfect it may be, should not be dismissed; there are hints of theme in these first searching notes.

An acceptable, broadly applicable solution to the control structure problem cannot be achieved by the dreaming up of a number of candidate configurations for a given process and then testing them. Such studies, a number of which have appeared in the recent literature, lead to few generalizations that have applicability to other conditions and other processes. Rather, the method must have its basis in a broadly applicable representation of the process dynamics and control objectives. It must acknowledge and address quantitatively problems of sparse and poor measurements and imprecisely known process characteristics. The task is to make decisions about control structure, and it is here where optimization can be most intelligently used. Such optimization problems are, however, a new breed. The standard formulation of optimal control will seldom suffice; it is decisions that are needed, not programmed trajectories.

The representation of the process dynamics alone is a major task. Despite the considerable number of investigations of chemical process dynamics in recent years, there is still no practicable method for formulating low-order models of large multivariable processes, other than the hand wrought construction long of service to process control engineers. Several techniques for the reduction of model order have been proposed, ranging from mode-based and singular perturbation procedures for lumped systems to moment matching and weighted residual methods for distributed processes. While each of these has certain merits, there remains the practical difficulty, particularly for large lumped systems, of formulating a sound, consistent model in the first place. It seems ridiculous to expend the effort to represent phenomena exhibiting weak interaction or rapid response knowing that these effects essentially will be dropped in the reduction to a lower-order model. In the body of knowledge of process dynamics, there seems to be missing an appreciation and an articulation of the dynamic structure of large interacting processes which, if known, would assist in the efficient representation of dynamic behavior in a form suitable for control analyses. The concepts of observability and controllability are involved here, but there is much more. In chemical processes, the matter of controllability and observability is usually not black and white. The ubiquitous interactions through chemical reactions and phase equilibrium relationships result in couplings among flows, temperatures, pressures, and compositions that produce a continuous spectrum of controllability and observability criteria of significance to the chemical engineer. But these have seldom been elucidated. Nor is it likely that such properties and dynamic structure will be elucidated through direct simulation alone. Simulation without analysis has little to offer to the codification of knowledge of dynamical processes or to a broadly applicable solution of the control configuration problem.

Associated with process modeling is the problem of state and parameter estimation. Estimates of measured and unmeasured states needed for control purposes require some sort of process model. But with the control structure unknown at the outset, the design of a state estimator becomes a much more difficult problem than in the past. The identity of the measured states will be unknown, and as it is certain that practicable estimators will not attempt to estimate all the states, the identity of those to be estimated will be unknown also. A state estimator of small dimension implies the use of a model also of small dimension, and the designer is thus confronted again with either the formulation of a low-order model or the reduction of one of large order. State estimators to be practicable must acknowledge and adapt to changing process parameters, and that requires the estimation of parameters. Both linear and nonlinear parameter estimation methods have been proposed, but nearly all those capable of on-line implementation are restricted to processes with a single output and just a few inputs. Further, the parameters determined by these methods are usually not basic physicochemical parameters, but rather those resulting from the numerical transformation of the model into one canonical form or another. Such model forms can easily have parameters numbering far in excess of the basic physicochemical parameters. While it is control system performance that counts (not parameter values per se), the large number of parameters in the transformed models can lead to computational tasks incapable of on-line implementation. Finally, the adaptive strategy for the state estimator as well as the controller needs to be stated. Little has been done by chemical engineers to elucidate the structure of adaptive systems capable of handling the continuously disturbed process with only a few measureable states. In the adaptive process there is considerable danger of instability, with the possible result that some adaptive functions cannot be carried out effectively. The chemical engineer's considerable expertise in stability analyses should serve him well here, but he must be forewarned that he can be no longer simply an observer of unstable phenomena; rather, he must synthesize stable control processes. Somehow that point of view seems unappreciated of late with the result that others (for example, Rosenbrock, 1971) have had to tackle the problem for him.

The objective functions by which control system structure is judged and selected must be carefully formulated. The choice is crucial to the applicability of the theory and the tractability of the analysis. The scalar indices, used almost exclusively in the past, will no longer suffice for most multivariable processes; vector valued indices

will be needed. What attributes of the controlled system these indices should reflect is not clear. Here it is imperative to reflect on the lessons of the past, for they have shown repeatedly that the primitive and superficial statements of control objectives must be transformed into criteria that have fundamental meaning within the ultimate framework of the theory. For example, the gain-bandwidth product and the relative orientation of eigenvectors are both far removed from simple statements of minimal process excursions, but both bear centrally and directly on accomplishing the latter. There is more than a suspicion that the work of genius is needed here, for without it the control configuration problem will likely remain in a primitive, hazily stated, and wholly unmanageable form.

Indeed, the same may be said for all aspects of the chemical process control problem. If not genius, then perception and the courage to tackle problems of an unfamiliar and unfriendly character. The insidious trend of the past decade to seek mere translations of the control techniques arising in other fields has left the chemical engineering profession destitute of incisive investigation and substantial resolution of its own unique problems. Instead only the elementary ideas should have been *borne across* the chasm separating spacecraft control from chemical process control and the seeds allowed to germinate in the virgin but unexplored valleys of the latter. That has not happened, but it must happen before practitioners can reasonably be expected to use the results of chemical process control theory. And it must be made to happen by those with experience in process engineering; there are few others who can perceive the problems and goals clearly and realistically. In fact, the chemical engineer is viewed by others to be in an extremely enviable position owing both to the wealth of control problems in his domain and his knowledge of processes. It would not be realistic to say, however, that he will be able to solve his problems singlehandedly; they are much too difficult. But if he can recognize those problems and respond with an imaginative attack and an inventive and pioneering spirit, there shall be some hope of narrowing the gap. The gap is present indeed, but contrary to the views of many, it is the theoretician who must close it.

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THE AUTHOR

The author is a pragmatic New Englander who has been trying to teach idealistic Californians about process control and dynamics for the last decade. He writes that his handful of publications on these topics, while not all of a theoretical genre, is in no way excluded from the set of gap-makers criticized in this essay.

A New Method of Parameter Estimation in Linear Systems

A time domain method is given for estimating the matrix or related parameters in linear systems with constant coefficients and real eigenvalues. The method consists of a one-dimensional search for the local minima of a scalar function $\mu(\lambda)$, which provide the eigenvalues of the system matrix and the matrix itself when observable. Applications are given to the determination of a transfer function and the estimation of the rate matrix of a monomolecular reaction system. Questions of accuracy, number, and type of measurements required are discussed.

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SCOPE

Process identification and parameter estimation constitute an important part of theoretical and applied work in the areas of process modeling and control, chemical kinetics, and transport phenomena. The present paper is concerned with parameter estimation in linear systems of ordinary differential equations with constant coefficients