Feedback Controllers for the Process Industries

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Preface

Much confusion currently exists about the performance of feedback controllers relative to one another and relative to what is even possible to expect. For example, model-predictive controllers are being sold for regulation where a proportional-integral-derivative (PID) controller could do better, and PID controllers are being viewed as outdated or unworthy of use in any modern control system. The reason for much of this confusion is the multifaceted nature of process control. Some, familiar with one or two facets, may feel that they have a universal solution to all process-control problems, when, in reality, their experience or perspective may not fit many or possibly most field applications. Academics are particularly susceptible to the narrow view in that most of their time is spent in a clean mathematical environment and a misapplication has no direct impact. Some even refuse to admit the existence of deadtime, although it can be observed wherever materials are transported. This book is an attempt to describe what, in fact, works and what does not in a plant situation, but based on very solid theory.

To be sure, process control sprung from humble beginnings, when mechanical levers and pneumatic bellows did most of the controlling and tinkerers such as John Ziegler and Nate Nichols worked out tuning rules using circular-chart records. As well known as the Ziegler-Nichols tuning rules seem to be, few realize that they apply strictly to step-load changes at the controller output on lag-dominant processes using the Taylor Fulscope controller. Try them on another process or with another controller or expect them to hold up for a load pulse or step at the process output or a set-point change, and you may be disappointed. Yet the fact that they have stood the test of 50 years of use bears witness to their fundamental worth and underlying truth. That they fail to meet your particular expectations in a given application does not mean that they are wrong or worthless. And while empirically derived, they do have a firm theoretical basis (which their authors may have known, but I only discovered years later).

Process-control theory has developed along several parallel paths. The frequency-response method used with electronic devices during World War II was applied to controllers and fluid processes. While this technology shed light on controllers, since they were mechanical and electronic devices, it was not very useful when applied to the processes being controlled. They were too slow and nonlinear to yield much information from frequency testing. While this method was then rarely used in the process industries, it continued to be taught in universities. Time-domain analysis is more applicable to the processes themselves—and easier to learn as well.

Later came optimal control theory and state-space analysis. Although applicable to aircraft and space vehicles, these approaches did not suit process control particularly well and were not adopted by industry. Yet they were taught extensively in universities, with the result that graduating students had to learn process control over again upon entering a plant environment.

The most recent trend in schools is toward model predictive and internal model control. Being based on a dynamic model of the process being controlled, they seem to have the requisite characteristics to succeed in a plant environment. If the process contains deadtime and nonlinearities, the model will include them. Control theory seems at last to be making a penetration into the plant environment.

Yet one unresolved issue remains—tuning. Proponents of model-based control hoped to have avoided tuning, and with good reason. Ziegler-Nichols rules were complicated enough for most practitioners, even in their limited scope. And the observation that most PID controllers have their derivative term set to zero indicates that many degrees of freedom only serve to confuse most people. The models being used in Smith predictors have 3 parameters to set in addition to the controller, and matrix-type controllers have as many as 30 parameters. How can all these be tuned?

In fact, they are not intended to be tuned at all. The model is matched to the process as closely as its complexity and testing accuracy will permit, and then it is held there until performance degradation indicates that another test is required. The operator is then given one adjustment over controller response—essentially the time constant of a filter that determines how fast the controller moves its output. As will be demonstrated, this is not enough to produce acceptable load-response performance on lag-dominant processes.

A look at the history of "one-knob" tuning does not impress. Taylor Instrument Companies produced a one-knob Bi-Act controller, and The Foxboro Company released its Model 59 controller in the 1950s; both were proportional plus integral controllers intended for flow control, and both were short-lived. Their knobs were uncalibrated or had no relevance to the process characteristics, and the controllers were too inflexible.

However, the history of model-predictive control is also barren. Otto J. M. Smith disclosed the Smith predictor in 1957,[1] but it was not even mentioned in Liptak's *Instrument Engineer's Handbook*[2] of 1970 or 1985. To a certain extent, this is understandable—the predictor requires a deadtime simulator, which was not commonly available before the advent of digital controllers. In addition, however, the need to set three model parameters and two controller modes certainly had to discourage users. Furthermore, its performance improvement exacted another price—reduced robustness. It was principally the work of E. B. Dahlin,[3] applying his model-predictive controller to paper machines, that brought attention to the method. His application benefited from two considerations:

1. Model-predictive control is most effective on deadtime-dominant processes (such as paper machines).
2. Stability requires model and process deadtimes to be matched (achievable by measuring machine speed).

The first condition provides higher performance than available with PID control, and the second provides the robustness that model-predictive controllers generally lack.

While not arguing with this success, the recent widespread application of model-predictive controllers to lag-dominant processes such as distillation columns is ill-advised. As will be demonstrated, performance declines exponentially with the ratio of lag-time constant to deadtime unless the model is intentionally *mismatched* against the process to maximize performance. This amounts to *tuning*, however. Because this is inconsistent with presently accepted practice and involves a combination of skill and empirical work, it is not promoted or even considered by proponents of model-based control.

My investigation into high-performance control exposed its dangerous cutting edge: As controller performance approaches 100 percent (of best possible), its robustness approaches zero. In other words, high-performance control teeters on the brink of instability. The high-performance controller is difficult to tune, demanding accuracy in its settings, and is extremely sensitive to parametric variations in the process being controlled. The price of performance comes high. This alone is enough to explain the staying power of the PI controller—it is extremely robust, although of relatively low performance.

Unfortunately, low performance does not guarantee robustness. A model-based controller matched to a lag-dominant process can have both poor performance and poor robustness at the same time. And if a filter or slow sampling or gain reduction is used to improve robustness, performance suffers even more.

The emphasis is on controller performance in this book, by placing it up front, in Part I. Chapter 1 examines economic measures of performance by describing the role of process control in plant economics. Chapter 2 then defines the theoretical limits of feedback-controller performance to set realistic goals for both the controller and the process. This theoretical limit is then the benchmark for performance evaluation of all types of controllers in the chapters that follow.

Part II introduces linear controllers, beginning with PID and its component parts in Chapter 3. This is followed by a presentation of sampling, a necessary evil in digital controllers. Chapter 5 then concentrates on several types of model-based controllers, pointing out their similarities, performance advantages, and limitations. An outgrowth of this study is the hybrid PIDτ_d controller, which combines high-performance with tunability.

Part III concentrates on controller tuning, first manually in Chapter 6, where the principles of dynamic modeling and performance optimization are developed. The procedures are then automated in self-tuning controllers in Chapter 7. Part IV investigates nonlinear elements, first by presenting various nonlinear controllers in Chapter 8 and finally by examining the nonlinear operating regions of constrained linear controllers in Chapter 9.

In my previous books, dynamic analysis has been confined to the time domain, because this is familiar to practitioners and easily assimilated by novices. However, certain particular aspects of controllers are easier to examine and compare by using frequency-response analysis and transfer functions. While the use of these methods will be minimized, they do assume on the part of the reader a grasp of process-control fundamentals and operational calculus. The control theory presented here is rigorous without being complex and is demonstrated by numerous simulations.

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Part

1

Performance Objectives

Chapter

1

Performance Criteria

The purpose of a controller is to keep a controlled variable at its desired value in the presence of disturbances from various sources and to cause it to follow changes in said desired value as closely as possible. The former—that of maintaining constancy in the presence of disturbances—is called *regulation,* while the latter—that of following changes in the desired value—is termed *servo response.* In mechanical systems such as machines and vehicles, servo response is the primary consideration, disturbances being relatively minor and intermittent. In control of fluid processes, however, regulation is the more important and more difficult function, in that unmeasured disturbances are frequent and severe; changes in the desired value (set point) tend to be common only in batch as opposed to continuous processes and in secondary, or "slave," loops. These distinctions are expanded and examined in more detail as individual applications are presented.

This introductory chapter is intended to establish the relationship between the ability of a controller to approach the preceding goals and the economic penalties for failing to do so. If a controller serves no economic function, then it has no justification in today's workplace. Be assured that safety and environmental protection fall under the economic umbrella, because accidents and pollution exact economic penalties. The issue here is that controllers and their support are costly, and the protection that they provide must justify their expense. Hence return on investment is always the primary consideration in industrial systems, and the controller that provides the earliest economic return represents the best investment. This establishes the need for economic measures of controller performance.

There are several areas where controllers can contribute to the economic performance of the plant being controlled. Each area has its own individual characteristics and needs which the controller must serve. Each is touched on below with respect to its own sources of economic penalties and the role of the controller in mitigating those penalties.

Limits of Safe Operation

Safety is the primary consideration in the operation of any system, be it an appliance, a vehicle, or a process plant. If the system cannot be operated safely, then it will not fulfill its primary productive function dependably or for long enough to repay its investment. The costs of accidents can be excessive—loss of life cannot even be evaluated satisfactorily, and the cost of damage to the environment keeps changing as we learn more about it. These factors therefore cannot really be entered into the economic equation. Safety simply must be built into the operation to minimize the likelihood of an accident through all the foreseen avenues. Control systems can contribute to safe operation, and they should. But they should not be the sole contributors. The plant must first be designed to fail safely, because fail it will, eventually. And the controls must be backed with a completely separate system of interlocks.

Independence of controls and safety interlocks

There should be several layers of protection built into any inherently hazardous operation. For example, a boiler will have several safety valves that lift at a pressure well below the stress limits of the vessel itself. Additionally, there should be a high-pressure interlock that will trip the combustion system before the safety valves lift. Third, there will be a steam-pressure controller that manipulates the firing rate and is intended to keep pressure well below either of the other limits. In this way, the interlocks and safety valves would be exercised only if the pressure-control system failed, which, although unlikely, could happen as a result of a severe upset, operator error, component failure, or some combination thereof.

It is mandatory that the different levels of safety systems have no common mode of failure. For example, the pressure controller and the high-pressure interlock should not use the same pressure transmitter or even the same type of pressure-measuring device. Their power sources should be separate, their signal wiring separately routed, etc., considering all events that might compromise both systems, such as a fire. And of course, both should fail safely, i.e., shut off the source of energy to the process, in the event of loss of either signal or power.

Redundant instrumentation can provide additional protection and should be used for devices whose reliability is lower than others in the rest of the system. Automatic selection should be provided between redundant pairs on a fail-safe basis so that the most likely failure will shut the unit down even if there is no accident. Since this event can be quite costly, a third redundant channel may be justified to protect against a single failure in either direction without causing a false shutdown. This is common practice in controlling the pressure inside balance-draft furnaces; three pressure transmitters send their signals to a median selector which discards the highest and lowest of the three signals. Three comparators are also required to identify errant transmitters. Two-out-of-three logic is also used commonly in protection and control of nuclear power plants.

Safety protection can be excessive to the point where the plant cannot be started or is subject to frequent "nuisance" shutdowns. This encourages operators to find ways around the interlocks, which may expose them to real dangers. Assuming that the controls and interlocks have been designed to be operable, the issue at hand is to avoid shutdowns caused by failure of the controls to keep critical variables from reaching the settings of the safety interlocks.

Cost of automatic shutdown

Loss of production is not the only cost of a shutdown. The shutdown operation itself will waste energy and material stored in the process, which must be removed. And the subsequent startup will require a similar amount of energy and material to be added to reach operating levels again. There also will be a period of time after startup before product quality will be acceptable, further extending production loss and requiring the recycling of off-specification product.

There are other hidden costs as well. Startup always stresses equipment and operators more than continuous production and is a time when most accidents occur.

Another factor is the interconnection that may exist between the tripped unit and others which may depend on it or may share the load with it. A tripped boiler may cause the shutdown of processes using its steam. If several boilers are supplying steam in parallel, production may continue only if others are able to pick up the load lost by the tripped unit. Having enough capacity on-line to continue at the same production rate is probably not economical. However, tripping one of several parallel units also stresses those remaining on-line. If their controls are unable to cope with the sudden load increase posed by the tripped unit, one or more of them may trip as well, which, if no automatic load shedding is in place, could bring down the entire plant. This was the cause of the Northeast power blackout of November of 1965—a component failure in one power station caused an overload which tripped one station after another until the entire