Basic and Advanced Regulatory Control: System Design and Application

Third Edition

By Harold L. Wade
Dedication

To Mary, my wife of many years, who provides love, support, encouragement, and criticism when needed.
About the Author ........................................................................................................... xiii

Preface to the Third Edition ............................................................................................. xv

Chapter 1  Introduction .................................................................................................. 1
  Symbols ......................................................................................................................... 4
  Exercises ....................................................................................................................... 5

Chapter 2  Mathematical Background, Diagrams, and Terminology. ....................... 7
  Mathematical Foundation ............................................................................................. 8
    Functions of a Variable .............................................................................................. 8
    Derivatives .................................................................................................................. 8
    Integrals ...................................................................................................................... 10
    Differential Equations .............................................................................................. 12
    Transfer Functions .................................................................................................... 15
    Frequency Response ................................................................................................. 21
  Diagrams and Terminology .......................................................................................... 22
  Direct- or Reverse-Acting Controllers .......................................................................... 31
  Exercises ....................................................................................................................... 33
  References .................................................................................................................... 37

Chapter 3  Process and Control Loop Characteristics ................................................. 39
  Steady-State Characteristics ....................................................................................... 40
  Control Valves .............................................................................................................. 46
    Valve Stem Stiction and Stick-Slip ............................................................................ 50
  Process Dynamic Characteristics ................................................................................ 52
    Types of Dynamic Response ..................................................................................... 56
  Control Loop Characteristics ....................................................................................... 67
<table>
<thead>
<tr>
<th>Chapter 11 Feedforward Control</th>
<th>311</th>
</tr>
</thead>
<tbody>
<tr>
<td>Designing Feedforward Control Systems</td>
<td>315</td>
</tr>
<tr>
<td>Additive Feedback</td>
<td>318</td>
</tr>
<tr>
<td>Multiplicative Feedback</td>
<td>321</td>
</tr>
<tr>
<td>Feedback Adjustment of the Feedforward Controller’s Reference Value</td>
<td>325</td>
</tr>
<tr>
<td>Compensation for Process Dynamics</td>
<td>327</td>
</tr>
<tr>
<td>Determining A(s) and B(s)</td>
<td>329</td>
</tr>
<tr>
<td>Fine-Tuning the Feedforward Controller</td>
<td>333</td>
</tr>
<tr>
<td>Further Considerations of the Feedback Controller</td>
<td>340</td>
</tr>
<tr>
<td>Feedforward: In Perspective</td>
<td>342</td>
</tr>
<tr>
<td>Feedforward Control Using FOUNDATION Fieldbus</td>
<td>346</td>
</tr>
<tr>
<td>Exercises</td>
<td>349</td>
</tr>
<tr>
<td>References</td>
<td>352</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Chapter 12 Override (Selector) Control</th>
<th>353</th>
</tr>
</thead>
<tbody>
<tr>
<td>Override Control</td>
<td>353</td>
</tr>
<tr>
<td>Real-World Applications of Override (Selector) Control</td>
<td>363</td>
</tr>
<tr>
<td>Transition from Start-Up to Normal Operations</td>
<td>363</td>
</tr>
<tr>
<td>Operating Near the Limit of a Process Utility</td>
<td>363</td>
</tr>
<tr>
<td>Prevention of “Tower Flooding” in a Distillation Tower</td>
<td>364</td>
</tr>
<tr>
<td>Pipeline Industry, Compressor Control</td>
<td>366</td>
</tr>
<tr>
<td>Other Methods of Implementation</td>
<td>368</td>
</tr>
<tr>
<td>“Pass-Through” Method</td>
<td>368</td>
</tr>
<tr>
<td>Forced Manual for the Nonselected Controller</td>
<td>372</td>
</tr>
<tr>
<td>Velocity (Incremental) Mode Control Algorithms</td>
<td>372</td>
</tr>
<tr>
<td>Pseudo-Velocity Method</td>
<td>374</td>
</tr>
<tr>
<td>Selection Based on Error</td>
<td>375</td>
</tr>
<tr>
<td>Override Control Using FOUNDATION Fieldbus</td>
<td>376</td>
</tr>
<tr>
<td>Exercises</td>
<td>378</td>
</tr>
<tr>
<td>References</td>
<td>381</td>
</tr>
</tbody>
</table>

<table>
<thead>
<tr>
<th>Chapter 13 Control for Interacting Process Loops</th>
<th>383</th>
</tr>
</thead>
<tbody>
<tr>
<td>Variable Pairing</td>
<td>384</td>
</tr>
<tr>
<td>Decoupling</td>
<td>396</td>
</tr>
<tr>
<td>Forward Decoupling</td>
<td>398</td>
</tr>
<tr>
<td>Inverted Decoupling</td>
<td>402</td>
</tr>
<tr>
<td>Partial Decoupling</td>
<td>408</td>
</tr>
<tr>
<td>Decoupling Application Examples</td>
<td>409</td>
</tr>
<tr>
<td>Petroleum Refinery Heater</td>
<td>409</td>
</tr>
<tr>
<td>Spray Water Temperature and Flow Control</td>
<td>411</td>
</tr>
<tr>
<td>Exercises</td>
<td>415</td>
</tr>
<tr>
<td>References</td>
<td>417</td>
</tr>
</tbody>
</table>
Chapter 14 Dead-Time Compensation and Model-Based Control ........... 419
  Smith Predictor Control ........................................... 421
  Algorithm Synthesis ............................................. 423
  Internal Model Control ........................................... 428
  Nonlinear, Model-Based Control ................................ 434
  PMBC ............................................................... 435
    Provisions for Dead Time ..................................... 441
  Exercises ........................................................ 446
  References ........................................................ 447

Chapter 15 Multivariable Model Predictive Control .................... 449
  Real-World Problems ............................................ 449
  History ............................................................ 451
  Unconstrained MPC for SISO Processes .......................... 452
    Process Model Identification .................................. 453
    Prediction ....................................................... 456
    Calculating Control Moves .................................... 458
  Incorporating Feedback ......................................... 461
  Tuning ............................................................. 462
    Summary Diagram .............................................. 465
  Unconstrained MPC for MIMO Processes .......................... 465
  Constrained MPC ................................................ 469
  Variations in MPC Vendor Offerings ............................. 471
  MPC in Perspective ............................................. 473
  Exercises ........................................................ 475
  References ........................................................ 476

Chapter 16 Other Control Techniques ................................. 479
  Split-Range Control ............................................. 479
  Cross-Limiting Control ......................................... 482
    Configuration Details ........................................ 485
    Provisions for Changes in Required Fuel-to-Air Ratio .... 487
  Floating Control ................................................. 489
    Floating Pressure Control for Distillation Columns ....... 490
    Floating Pressure Control for Steam Systems ............... 492
    Hot or Chilled Water Supply Systems ......................... 494
    Cooling Tower Systems ....................................... 494
  Increasing Valve Rangeability ................................ 496
    Small and Large Valves Operating in Parallel .............. 496
    Small and Large Valves Operating in Sequence ............. 498
  Time Proportioning Control .................................... 501
  Exercises ........................................................ 503
  References ........................................................ 506
About the Author

Harold L. Wade, PhD, has had over 50 years of experience in designing, applying, and installing process control systems in such industries as petroleum refining, chemical processing, textiles, waste and water treatment facilities, and others. He has a BS in mechanical engineering and an MS and PhD in systems engineering. He has held technical positions with Honeywell, with Foxboro, and with his own consulting engineering firm. He is a Fellow of ISA, a Life Member of IEEE, and a licensed professional engineer in Texas. Dr. Wade was a 2002 inductee into Control magazine’s Process Automation Hall of Fame and was the 2008 recipient of the Donald P. Eckman Award presented by ISA.

For almost 20 years, Dr. Wade taught courses for ISA on process control systems design. He has also presented process control and control tuning seminars for many companies worldwide. Dr. Wade is the developer of the process control training program, PC-ControLAB.
Preface to the Third Edition

This book presents a practical approach to process control for the chemical, refining, pulp and paper, utilities, and similar industries. It is the result of many process control seminars presented in the United States and abroad. A typical participant in one of those seminars is an engineer, currently employed by a processing company, who may have had formal training in an undergraduate process control course, but who may not be able to fully relate the material from that course to his or her work experiences. This book aims to meet this need by explaining concepts in a practical way with a minimal amount of theoretical background. The book serves both the beginning and the experienced control systems engineer. For the beginning engineer, it initially presents very simple concepts. For the experienced engineer, it develops these initial concepts to provide deeper understanding and/or new insights into familiar concepts. The purpose is to provide everyone, beginner or experienced engineer, with something they can put to beneficial use in their plant.

Although this is intended to be a practical how-to book, readers should not infer that it is devoid of mathematical concepts. Where such concepts are utilized, it is their application to practical situations, rather than the theory behind the concepts, that is emphasized. A theme of the first edition was that wherever a choice had to be made between providing mathematical rigor or promoting intuitive understanding, preference was given to understandability. This theme has been carried forward into later editions, and this practicality distinguishes this book from many academic texts. End of chapter exercises have also been added to the present edition.
The book is organized in three parts. The first three chapters present background information, including a brief non-rigorous mathematical review, a discussion of symbols and terminology, and a description of general characteristics of processes and of selected types of control loops.

The second part—Chapters 4 through 7—addresses feedback control. The objective is to provide the reader with a thorough intuitive grasp of feedback control behavior and all its nuances. Examples of current commercially available products, DCSs and PLCs, have been updated. In Chapter 5, the brief mention made in the previous edition about “two degree of freedom controllers” (called set-point weighting by some manufacturers) has been expanded to give proper emphasis to this recent (within the digital era), very important, and often overlooked addition to PID capabilities. This modification is also mentioned in the following chapter, in relation to controller tuning. In Chapter 6, on feedback controller tuning, the discussion on improving as-found tuning (also called intelligent trial-and-error tuning) was expanded considerably in the second edition, and is further expanded in this edition, to consider overshoot ratio, as well as decay ratio when formulating controller-tuning objectives. This tuning technique has been proven in practical applications and has been well accepted in training classes in which it has been presented. An additional technique is presented in this edition for improving-as-found tuning; this technique considers process dynamics as approximated by a first-order-lag plus dead-time model. Also in this chapter, new material is included on tuning liquid-level control loops. The tuning of these loops, which have a completely different characteristic from most other process control loops, has, in general, received very little specific attention in process control literature.

The third portion of the book, Chapters 8 through 16, covers advanced regulatory control topics. Chapter 8 defines the penalty that must be paid if feedback control alone is used. This leads into a discussion of advanced regulatory control techniques that can significantly reduce this feedback penalty. Additional chapters provide a detailed discussion of the technology and application examples of cascade (Chapter 9), ratio (Chapter 10), feedforward (Chapter 11), override (Chapter 12), decoupling (Chapter 13), model-based (Chapter 14), and model-predictive control (Chapter 15). The chapter on feedforward control covers both additive and multiplicative forms of feedforward. The chapter on override (selector) control includes additional application examples for this technique, as well as an assessment of the performance of several alternative techniques. The chapter on the control of multiple-input multiple-output
(MIMO) processes (Chapter 13) contains additional coverage of inverted decoupling. This MIMO technique was introduced in the first edition; new material previously available only in technical journals is presented here.

The chapter on model-based control in the first edition has been divided into two chapters. Chapter 14, devoted primarily to dead-time compensation, covers Smith predictor control and internal model control. The other chapter, Chapter 15, contains an introductory discussion on model predictive control.

Chapter 16 covers topics on process control application that do not readily fit into any of the other chapters. In addition to cross-limiting control for fired heaters, which was covered in the first edition, these new topics include floating control, techniques for increasing valve rangeability, and time proportioning control.

Also contained in this section are illustrations of the use of FOUNDATION Fieldbus (FF), which permits the control strategy to be distributed directly into field devices. The network architecture, communication, and implementation aspects of FF are briefly summarized in Chapter 5. The chapters on modifications to feedback control, cascade, ratio, feedforward, and override (Chapters 5, 9, 10, 11, and 12) all conclude with an example in which that chapter’s strategy is implemented using FF function blocks.

I would like to express gratitude to the many students who, by asking probing questions, have enabled me to revise and sharpen my presentation and come up with examples that are more meaningful.

I would also like to express my thanks to Dr. R. Russell Rhinehart for many helpful comments and suggestions, to my longtime friends Greg Shinskey and Victor Wegelin for their mentoring and helpful suggestions, and to my friend Vu Van Phin in Viet Nam (with whom I have had the privilege of meeting only via email) for many enlightening technical exchanges.

Lastly, I have special thanks to my editors, Liegh Elrod and Scott Bogue, for their diligence, which has helped me to produce a better final product.
The term *process control* implies that there is a *process* for which there is a desired behavior and that there is some *controlling function* that acts to elicit that desired behavior. This broad concept can embrace everything from societal processes governed by some regulatory control authority to automated manufacturing processes. In practically all cases, however, a common thread is that some measure of the actual process behavior is compared with the desired process behavior. This feedback action then generates a control policy that acts to minimize or eliminate the difference between desired and actual behavior.

This book focuses on a particular segment of automated process control that is applied to chemical, refining, pulp and paper, power generation, and similar types of processes. Even within this limited scope of applications, the discussion is restricted primarily to processes that are operated continuously for long periods of time and within a narrow region of the operating variables. In other words, the book excludes such important operating modes as batch processing, start-ups, and grade changes. Many of the control techniques to be presented here, however, can be adapted to these other modes of operation.

The behavior of the processes in this book is often characterized by measured values of process variables, such as temperatures, flow rates, and pressures. The desired behavior is stated to be the set points of those process variables. Until fairly recent times, most applications of industrial process control used simple feedback controllers that regulated the flow rates, temperatures, pres-
sures and other process variables. These controllers required a form of adjustment called tuning to match their controlling action to the unique requirements of individual processes.

As long as most of the control systems were implemented with analog hardware, applications were limited to simple regulatory control. This was due to the cost of additional components and additional interconnections that more advanced control required, along with the burden of additional maintenance and the vulnerability to failure of many devices in the control loop. With the advent of digital control systems, however, more sophisticated loops became feasible. Advanced regulatory control techniques, such as ratio, cascade, and feedforward control, as well as additional forms such as constraint (selector) control and decoupling, could readily be implemented simply by configuring software function blocks.

With this additional capability, however, a need developed for a systematic approach toward using it. This is called control strategy design. In order to design a technically successful and economically viable control strategy, the control system engineer must be well grounded in the techniques of feedback control as well as the tools of advanced regulatory control. The requisite knowledge includes both how to implement and how to tune both the basic and the advanced regulatory control strategies. Even before that, however, the control system engineer must be adept at recognizing when to use (and conversely, when not to use) certain control strategies, as well as projecting the expected benefits.

Using advanced regulatory control provides multiple benefits. One of the most important is simply closer control of the process. It will become clear later in this book that with basic regulatory (i.e., feedback) control, there must be a deviation from a set point before a control action can occur. We will call this the feedback penalty. While the feedback penalty also applies to advanced regulatory control, the primary objective of advanced regulatory control is for the control action to be taken while incurring only a minimal feedback penalty. The reduction in feedback penalty may be stated in a variety of ways, such as a reduction of the maximum deviation from the set point, or as a reduction in the amount of off-spec product produced. This reduction in feedback penalty can provide several forms of economic benefit, such as improvement in product quality, energy savings, increased throughput, or longer equipment life.
Process control is but one part of an overall process control and information system hierarchy (see Figure 1-1). It extends downward to safety controls and other directly connected process devices and upward to encompass optimization and even higher levels of business management, such as scheduling, inventory, and asset management. Indeed, corporate profitability may be enhanced more significantly as a result of these higher-level activities than from improved process control *per se*. However, since each layer of the hierarchy depends on the proper functioning of the layers beneath it, one of the primary benefits of advanced regulatory control is that it enables the higher levels, optimization and enterprise management and control.

![Figure 1-1. Overall Process Control and Information System Hierarchy](image-url)
Symbols

Many mathematical symbols are used throughout this book. Some symbols are used only for the discussion of a particular topic; these symbols are therefore defined in that discussion and are not listed here. Some symbols may have multiple meanings that depend on the context; additional meanings are provided as needed for clarity in the appropriate sections of this book. Chapter 2 discusses the graphical symbols used in control system documentation. Chapter 15 uses a unique set of symbols that are defined at the beginning of that chapter.

The following symbols are used throughout this book:

\[ b = \text{bias value (manual reset) on proportional-only controller output} \]
\[ e = \text{error (deviation between the set point and process variable)}^{1} \]
\[ E = \text{when capitalized, refers to (Laplace) transform of error} \]
\[ K = \text{steady-state gain of first-order lag} \]
\[ K_C = \text{controller gain (noninteractive and interactive control algorithms)} \]
\[ K_D = \text{derivative gain (independent gains control algorithm)} \]
\[ K_I = \text{integral gain (independent gains control algorithm)} \]
\[ K_P = \text{proportional gain (independent gains control algorithm)} \]
\[ K_p = \text{process gain (change in process variable/change in controller output)} \]
\[ m = \text{manipulated variable, controller output} \]
\[ M = \text{when capitalized, refers to (Laplace) transform of manipulated variable} \]
\[ PB = \text{proportional band} \]
\[ PI = \text{control algorithm with proportional and integral modes} \]
\[ PID = \text{control algorithm with proportional, integral, and derivative modes} \]
\[ PV = \text{process variable (see also symbol } x) \]
\[ SP = \text{set point (see also symbol } x_{SP}) \]

---

1. The symbol \( e \) can also be used as the basis of natural logarithms, for example, when expressing the Laplace transform of dead time.
\[ T_D = \text{derivative time (noninteractive and interactive control algorithms)} \]
\[ T_I = \text{integral time (minutes/repeat) (noninteractive and interactive control algorithms)} \]
\[ x = \text{process variable (see also symbol } PV \text{)} \]
\[ x_{SP} = \text{set point (see also symbol } SP \text{)} \]
\[ u = \text{disturbance variable} \]
\[ \alpha = \text{derivative gain (when a derivative filter is used with a noninteractive or interactive control algorithm)} \]
\[ \theta = \text{dead time} \]
\[ \tau = \text{first-order lag time constant} \]

**Exercises**

1.1 An incongruous illustration of feedback control is that of driving an automobile by looking only in the rear-view mirror. You can see where you have been, but you cannot see what disturbances (hills, curves or bumps in the road, other vehicles, etc.) are coming. Nevertheless, an analogy can be made between this example and components of an ordinary feedback control loop. Complete the following table:

<table>
<thead>
<tr>
<th>Set point</th>
<th>Center of your lane</th>
</tr>
</thead>
<tbody>
<tr>
<td>Process variable</td>
<td>____________________</td>
</tr>
<tr>
<td>Error (or deviation)</td>
<td>____________________</td>
</tr>
<tr>
<td>Controller</td>
<td>____________________</td>
</tr>
<tr>
<td>Final control element</td>
<td>Steering mechanism</td>
</tr>
</tbody>
</table>

1.2 Another, more realistic, example of a feedback control loop is speed control. (Assume that what you are driving does not have automatic cruise control, which would be a true feedback loop.) Here, if you see on the speedometer that the vehicle is slowing down, you apply more pressure to the accelerator pedal; if you see that it is traveling faster than your desired speed, you reduce the pressure on the accelerator pedal. Complete the following table:
1.3 Another analogy between driving and process control is driving while looking forward. For example, if you see a hill ahead, you depress the accelerator pedal just as you approach the base of the hill, and you reduce pressure on the accelerator pedal as you approach the crest of the hill. If these actions are taken at the right times, and in the right amounts, there will be no deviation from the desired speed. This type of control is called: ____________________.

1.4 There are many important topics in an overall automation project. As important as they are, which of the following topics are not within the scope of this book?

- Enterprise integration
- Safety instrumented systems
- Sensor/analyzer selection and installation
- Alarm management
- Abnormal situation management
- Human-machine interface (HMI) design
- Cybersecurity
- Communications technology (for plant instrumentation and control systems)
In order to design, analyze, or commission a process control system, one must be familiar with the characteristics of the process itself. Although it is highly beneficial for the control engineer to have a good understanding of the physical and chemical phenomena that govern the process, his or her view of the process will usually differ from that of the process design engineer. The discussion in this chapter is meant to develop the thought processes of a control engineer. Although some of the following points may seem to overstate the case, they will enable us to highlight the differences in the ways the control engineer and process engineer think, and to call attention to the process details that the control engineer should thoroughly understand.

The process engineer is concerned with meeting production rate and quality specifications, which are often called the design conditions. The control engineer is concerned with operating an existing process outside of design conditions, often with reduced throughput, variations in feedstock, or other abnormal conditions.

The process engineer’s objective is often to minimize the initial cost (or the life-cycle cost) of the processing equipment. The control engineer’s objective is to make the most efficient use of the equipment that is already installed.

The process engineer considers those design parameters that can be specified as independent variables. Other parameter values that are derived from these are dependent variables. For example, the pressure of a saturated steam sys-
tem might be an independent variable that can be specified during the design process; the temperature then becomes a dependent variable. The control engineer considers as independent variables the control points (for instance, valve positions or flow rates) that can be manipulated to affect the process. The steam pressure then becomes a dependent variable that results from those valve positions or flow rates.

The process engineer is usually concerned with the steady-state conditions of the process. The control engineer must necessarily take into consideration both the steady state and the dynamic, or transient, behavior of the process.

The characteristics of each process will be different. Even so, from the control engineer’s viewpoint, certain characteristics are similar from process to process. It is these characteristics that will be emphasized here.

**Steady-State Characteristics**

When all inputs and external influences are held constant, most, but not all, processes come to a steady state. (Liquid level is different. Unless the inflow to and outflow from a liquid-level process are equal, the process will not come to a steady state, even though the inflow and outflow themselves are constant.) We will use the heat exchanger depicted in Figure 2-7 to illustrate the nature of the steady state. This is redrawn as Figure 3-1 here, but in this case the independent and dependent variables are identified. Let us assume that we have a liquid phase process stream that must be heated to a specified temperature. Let us also assume that we have a liquid phase heating medium, such as hot water or hot oil. (If this process unit is just a heat exchanger, with no chemical reaction taking place, then both the inlet and outlet temperatures of the process stream will be at a lower temperature than the inlet and outlet temperatures of the heating medium. We will designate the process side as the cold side, the flow rate as $F_c$, and the inlet and outlet temperatures as $T_{c-in}$ and $T_{c-out}$. Similarly, on the hot side, the flow rate of the heating medium will be designated as $F_h$, with its inlet temperature as $T_{h-in}$ and its outlet temperature as $T_{h-out}$.)

From a process control viewpoint, the independent variable is the valve position, or equivalently, the controller output. The dependent variable of interest is the process outlet temperature, $T_{c-out}$. Other dependent variables are the flow rate of the heating fluid, $F_h$, and the outlet temperature of the heating medium, $T_{h-out}$. In a typical operating plant, these may be monitored to detect abnormal operation, but from a control viewpoint they are inconsequential.
Other important variables to be considered are the \textit{disturbances} to the process. These are sometimes called \textit{load changes}. They can be considered as external, random influences on the process. It is the purpose of the control system to counteract the effects of these disturbances. Some of the disturbances that could affect the heat exchanger are:

- Changes in the process flow rate, $F_c$
- Changes in the process inlet temperature, $T_{c-in}$
- Changes in the source temperature of the heating medium, $T_{h-in}$
- Changes in the upstream or downstream pressure of the heating fluid (this would change the hot stream flow rate, $F_h$, even though the valve position did not change)
- Scaling of the heat exchanger tubes—thus affecting the heat transfer coefficient
- Environmental effects, if the heat exchanger is not perfectly insulated
For the purposes of illustration, we will disregard the latter three of these disturbances (i.e., we will assume that they are constant) and concern ourselves only with $F_c$, $T_{c-in}$, and $T_{h-in}$. For the time being, we will also consider that these three variables are also being held constant. In other words, the only independent variable is the valve position, which uniquely sets the heating medium flow rate. With this consideration, we state an important principle:

*If all external influences on a process are held constant, then each value of the control signal (independent variable) produces a specific and unique measurement value (dependent variable).* (This is somewhat of an idealized statement, since it ignores the very real probability of valve stick-slip and hysteresis, which is described in a subsection below.) There are rare cases, such as the discharge pressure of a centrifugal compressor versus suction flow or index of refraction versus composition, where this unique relationship may not be true.)

This one-to-one relationship can be depicted in graphical form, as shown in Figure 3-2. We call this relationship the *process graph*. Keep in mind that the process graph depicts the steady-state relationship between the controller output (valve position) and the measurement for a particular combination of the disturbance variables.

![Figure 3-2. Process Graph](image)
If any of the disturbance variables change in value, then we will have a new process graph. Figure 3-3 shows the process graphs for three combinations of disturbance variables. The process graph for the original values of $F_c$, $T_{c-in}$, and $T_{h-in}$ is shown by the dotted curved line. The upper line is the process graph for an increase in $T_{c-in}$. The lower line is the process graph for an increase in $F_c$.

![Figure 3-3. The Shifting of a Process Graph as a Result of Disturbances](image)

The process graph—the steady-state relationship between the controller output and measured (dependent) variables for a particular combination of disturbance variables—is an important concept for understanding control loop behavior. However, it is not something that we need to determine in actual practice. Indeed, it would be impractical to determine the infinite number of process graphs that would result from all combinations of the disturbance variables.

Nevertheless, we can deduce that if we wish to control the measurement to a particular value (set point), the process graph determines the required value of the controller output. If there are load changes on the process that cause the process graph to shift, we will need a new value for the controller output. It is
the duty of the controller to determine the precise point on the process graph that brings the measurement to the desired value, as shown in Figure 3-4.

Although we will normally not have a precise process graph available for even one combination of disturbance variables, there are certain attributes of the process graph that we must know. First, we must know whether the process graph slopes upward or downward. This is equivalent to saying that we must know whether the process is direct-acting or reverse-acting. An upward slope represents a direct-acting process (an increase in controller output causes an increase in measurement); a downward slope signifies a reverse-acting process. Recall from Chapter 2 that to avoid positive feedback, the controller must be of opposite action—reverse-acting for a direct-acting process and vice versa.

We must also know, either explicitly or implicitly, the slope of the process graph, at least in the vicinity of the most probable operating point. The slope can be defined as the change in measurement divided by the change in con-
troller output. This is called the process gain. Process gain, \( K_p \), is defined by the following equation:

\[
K_p = \frac{\text{Change in measurement}}{\text{Change in valve signal}}
\]

\[
= \frac{\Delta x}{\Delta m}
\]

(3-1)

The process gain often varies with the operating point. This is equivalent to stating that the process, and hence the process graph itself, is often nonlinear. Except for some rare misbehaved processes, however, the process graph is monotonic. That is, the direction of the slope (upward or downward) does not change, so there is a unique relationship between each value of the controller output and the measurement.

Process nonlinearities can be caused by a number of conditions, including physical or chemical factors inherent in the process itself. In one frequently encountered situation, the process variable responds linearly to changes in the ratio between two variables, such as the manipulated variable and a disturbance variable, whereas it would respond nonlinearly to each of the two variables considered separately. To illustrate this, suppose that a process heater can be modeled by a simple heat-balance relationship:

\[
F_p C_p (T_{out} - T_{in}) = F_g H_v E_{ff}
\]

(3-2)

where

\[
F_p = \text{heater feed rate (the disturbance variable)}
\]

\[
C_p = \text{specific heat of the processed material}
\]

\[
T_{out} = \text{outlet temperature (the process variable)}
\]

\[
T_{in} = \text{inlet temperature}
\]

\[
F_g = \text{fuel rate (the manipulated variable)}
\]

\[
H_v = \text{heating value of the fuel}
\]

\[
E_{ff} = \text{heater efficiency}
\]

Equation 3-2 can be rearranged to show the outlet temperature on the left-hand side of the equation and all other terms on the right-hand side:

\[
T_{out} = T_{in} + \frac{H_v E_{ff} F_g}{C_p F_p}
\]

(3-3)
This demonstrates that the outlet temperature responds more or less linearly to the fuel rate-to-heater feed rate ratio, \( F_g/F_p \). If a temperature controller directly manipulates the fuel rate, then the process gain seen by the controller is the sensitivity of the outlet temperature to changes in fuel rate. Specifically:

$$\frac{\Delta T_{out}}{\Delta F_g} = \frac{H_v E_{ff}}{C_p} \times \frac{1}{F_p}$$  \hspace{1cm} (3-4)

In other words, the process gain of the control loop is inversely proportional to the process flow rate. At a low process flow rate (such as during start-ups), the process gain will be high; at higher flow rates, the process gain will be lower.

If the ratio itself were the manipulated variable, rather than simply the fuel rate, then the process gain seen by the control loop would be the following:

$$\frac{\Delta T_{out}}{\Delta (F_g/F_p)} = \frac{H_v E_{ff}}{C_p}$$  \hspace{1cm} (3-5)

As long as the fuel heating value, heater efficiency, and specific heat of the process fluid remain fairly constant, then the control loop’s process gain will remain constant. This strategy will be used in relation to ratio control in Chapter 10 and in multiplicative feedforward control in Chapter 11.

**Control Valves**

Nonlinearity in a control loop may also be caused by the nonlinear characteristic of the valve. There is also the friction in the packing gland of the valve that will affect stem movement, and contribute to the dynamics of the control loop. Therefore, before continuing with dynamic characteristics, we will discuss control valves.

There are many types of control valves; these differ by the type of valve body, by the type of valve actuator, and by whether or not a valve positioner is used. The discussion below will focus on a common type of control valve: a globe valve with a sliding stem plug and seat (the closure mechanism), and a spring and diaphragm actuator, with and without a valve positioner as depicted in the elementary diagram, Figure 3-5.
The power of the PID controller is that it can be adjusted to provide the desired behavior on a wide variety of process applications through the judicious choice of one, two, or three parameter values, and with only modest knowledge about the process. Determining acceptable values for these parameters is called tuning the controller.

In the process industries, those who tune the controllers often face a number of adverse factors:

- The process dynamics are usually not well known.
- There are often nonlinearities in the process that cause the process response to change with operating conditions.
- There is often an unwanted signal component (called noise) on the measurement.
- The loop may be subject to random load changes.
- Frequently, the interaction between control loops makes it difficult to discern the tuning effects of a particular loop from the interactive response with other loops.
- If the final actuator is a valve, it may also contribute to the problem:
  - It may be the wrong size (often oversized).
It may contribute to the nonlinear response, as described in Chapter 3.

Due to valve stiction (described in Chapter 3), the process variable may be driven into a limit cycle, even if the set point is being held constant. Depending on the type of valve and positioner, if one is being used, it may or may not be possible to reduce this limit cycle by controller tuning.

In addition to these problems, the loop tuner often must work on an ongoing process, which allows for only minimal or no experimentation or testing. Given these adversities, it is a wonder that so many PID loops provide more or less satisfactory performance. On the other hand, it is probably true that the tuning could be improved for a significant number of all control loops.

Also contributing to the difficulty the tuner faces is the fact that there is no general agreement as to what constitutes good tuning. Therefore, before discussing the tuning procedures, we will discuss performance criteria, both informal and formal. The tuning procedures that will then be presented can be grouped into the following categories:

- Tuning for self-regulating processes
  - Trial-and-error tuning
  - Tuning from open-loop tests
  - Tuning from closed-loop tests
  - Improving “as-found” tuning

- Tuning for integrating processes
  - Idealized liquid-level control loops
  - Real world considerations for liquid-level control
  - Other approaches to liquid-level control tuning
  - Other integrating processes

Following this, we will present typical tuning values for some of the more common types of control loops, the chapter will close with general admonitions and recommendations for the controller tuner.
Performance Criteria

If an operator were asked their preference of the various responses to a set-point change shown in Figure 6-1, they would probably designate “a” as the first choice, “b” as acceptable, “c” next, then “e,” and “f” last. Response “d” would probably be unacceptable, due to the undershoot of the first peak. Unfortunately, however, preferred choice “a” provides a poor choice for rejection of a step disturbance at the process input, whereas “f” probably provides the best choice for disturbance rejection. One behavioral fact about any type of traditional PID controller is that if the loop is tuned to give a desirable response to a set-point change (say, both a low decay ratio and a low overshoot ratio), the response to a disturbance may be too sluggish. On the other hand, if the controller is tuned for a more aggressive response to minimize the response to a disturbance, the response to a set-point change may be overly aggressive.

![Figure 6-1. Examples of Various Forms of Response to a Set-Point Change](image-url)
For many control loops, the set point is rarely changed. The purpose of these loops is to minimize the effect of disturbances. Even so, because set-point changes are usually made more easily than load changes, in actual practice many loops are tuned for a suitable response to a set-point change. Then, the resulting response for a load upset is accepted even though it may not be the best. A preferable tuning strategy would be to tune the controller for the best response to a load change, then use one of the set point “softening” techniques or the two-degrees of freedom controller described in Chapter 5 to ameliorate the effect of occasional set-point changes. If actual load changes cannot be made, the effect can be simulated by placing the controller in Manual, changing the controller output, and then quickly returning the loop to Automatic. Another alternative is to initially tune the loop for a quarter-decay response to a set-point change, because tuning for that response provides a very acceptable disturbance response. You would then apply one of the set point softening techniques mentioned in Chapter 4 (derivative mode on measurement, proportional mode on measurement, set point ramping, or two-degree of freedom controller) to obtain an acceptable set-point response.

The performance criteria can be refined beyond the qualitative figures by quantifying the requirements for certain parameters. For example, values may be designated for maximum and minimum amounts of overshoot. A value may also be specified for the decay ratio as shown in Figure 6-1f. This figure illustrates the traditional definition of decay ratio as the ratio of the deviation from the set point at the second peak after a set-point change to the deviation at the first peak. This is also depicted in Figure 6-2a. Occasionally, the set-point response is such that this definition is not useful. A better definition of decay ratio is the ratio of the difference between the second peak and its succeeding valley to the difference between the first peak and its succeeding valley. This is depicted in Figure 6-2b. This definition, though more cumbersome, will work in all cases. Most of the time, however, the simpler and more widely used definition depicted by Figure 6-2a will suffice.

The decay ratio can also be defined for a disturbance or load upset. For a step change in load, the behavior depicted in Figure 6-2c is typical. Here, the decay ratio must be determined by the ratio of peak-to-valley differences. A load-

---

1. This second definition, although not exactly correct, is more defensible theoretically because the set-point response is the composite of a filtered exponential rise and a damped sinusoidal signal. If the rise time of the exponential is sufficiently fast, then the two definitions are essentially the same. This is why the first definition is valid in most circumstances.
upset response like that depicted in Figure 6-2d is somewhat unusual for most processes but is typical of the load-upset response of level control loops.

One well-known criterion for controller tuning is a decay ratio of one-fourth following a set-point change. This is also called quarter-wave decay, quarter-wave damping, and quarter-amplitude decay. This criterion states that if a loop is oscillating, each peak deviation should be only one-fourth of the previous peak deviation on the same side of the set point. This is equivalent to stating that on each half cycle, the amplitude of deviation should be decreased by approximately one-half, making the total decrease one-fourth for a full cycle.

The decay ratio, however, may not provide a useful performance specification. The overshoot ratio may be a preferable performance criterion. Even if a loop exhibits quarter-decay response to a set-point change, it probably will not be considered as being acceptably tuned if there is excessive overshoot. Most people would probably prefer the overshoot to be no more than 5–10% of the magnitude of the set-point change. A lower limit, such as 5%, may also be specified for the overshoot ratio; if however, there were no overshoot, the rise time to the set point may be excessive. These are soft limits, as will be seen when we discuss improving as-found tuning.
Other criteria that are sometimes used to measure control loop performance include *rise time*—the time between a set-point change and the first crossing of the set point; and *settling time*—the time, following a set-point change or disturbance, that it takes for the oscillation to become so small that the deviation does not exceed some specified amount. These will have little use in this book, due to their relative infrequent use in normal process control tuning activities.

The criteria mentioned so far have been primarily judgement-based. That is, there is no formal justification for them, other than they appear to be satisfactory. Formal criteria that can be used to evaluate loop tuning are based on minimizing the integral of some function of the error. The following four integral-error criteria can be considered:

Integral of the absolute value of the error (IAE):

\[
\text{IAE} = \int |e| dt \quad (6-1)
\]

Integral of the square of the error (ISE):

\[
\text{ISE} = \int e^2 dt \quad (6-2)
\]

Integral of time × absolute value of the error (ITAE):

\[
\text{ITAE} = \int t|e| dt \quad (6-3)
\]

Integral of time × square of the error (ITSE):

\[
\text{ITSE} = \int te^2 dt \quad (6-4)
\]

Note that the simple integral of the error is not a valid criterion because the integration of a positive error would be canceled out a half-cycle later by the integration of a negative error. The four criteria just listed avoid this, either by taking the absolute value of the error or by squaring the error.

Minimizing each of these integral-error criteria will produce a different response. For example, a plant that uses the ISE criterion pays an increasingly large penalty as the magnitude of the error increases. Therefore, for a given
loop, the ISE criteria will result in a smaller maximum deviation value than the IAE criterion, but it may cause the oscillation to persist longer.

The rationale for the last two criteria listed (ITAE and ITSE) is that the longer an error persists after a set-point or load change, the more heavily it should be penalized. Thus, the ITAE criterion will permit a greater initial deviation than an IAE criterion, but it will force the oscillation to die out sooner.

If there is noise on the process variable, any of these criteria will increase without bounds. To be valid as measures of performance, the same time span should be used for integration in any cases to be compared.

Integral-error criteria are also useful for academic, theoretical, and control simulation studies for providing insight into the tuning process. They may also be used in control loop audits. They are rarely used in actual control loop tuning, however. For a typical process, there can be quite a variation from the point of optimum PI tuning parameters (those that minimize the IAE) without a significant increase in IAE and, consequently, without a significant change in the observed response.

Figure 6-3 depicts graphically the typical amount of change in IAE as the PI tuning parameters are varied from their optimum values. The data for this figure was obtained by simulating a typical process. Because comparing actual numbers for two different examples is meaningless, normalized parameter values are used in Figure 6-3. Table 6-1 shows both actual and normalized parameter values at both the minimum IAE point and at point A. The point the figure and table illustrate is that there can be a considerable variation of tuning parameter values from the so-called optimum values without incurring a significant change in the IAE metric.
Figure 6-3. A Typical Relation between Normalized IAE and Normalized PI Tuning Parameters

Table 6-1. Actual and Normalized Values for IAE, $K_C$, and $T_I$ at Minimum IAE and Point A

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Minimum IAE</th>
<th>Point A</th>
</tr>
</thead>
<tbody>
<tr>
<td></td>
<td>Actual Value</td>
<td>Normalized Value</td>
</tr>
<tr>
<td>IAE</td>
<td>68.7</td>
<td>1.0</td>
</tr>
<tr>
<td>$K_C$</td>
<td>1.57</td>
<td>1.0</td>
</tr>
<tr>
<td>$T_I$</td>
<td>8.53</td>
<td>1.0</td>
</tr>
</tbody>
</table>
To summarize this section, what constitutes good tuning is often a subjective matter that can vary from application to application—as well as from person to person tuning the loop.

**Tuning for Self-Regulating Processes**

Because self-regulating and non-self-regulating processes have a different character, the tuning procedures that are applicable to one may not be applicable to the other. In this section, we will restrict our comments to self-regulating processes. The next section is devoted entirely to tuning controllers for non-self-regulating processes, primarily liquid-level control.

**Trial-and-Error Tuning**

Most loops are tuned by an experimental technique. Even when a formal technique, such as open-loop or closed-loop testing, is used to determine the initial tuning, a final bit of fine tuning may be in order. Trial-and-error tuning requires the user to observe the response of the loop to a previous event, either a set-point change or load change, and then decide what tuning parameter (or parameters) should be changed, in which direction, and by how much. Experience helps in interpreting the response. The user must also have a thorough understanding of the effect of changing each of the tuning parameters.

**Response to Various Tuning Parameter Combinations**

For a self-regulating process controlled by a PI controller (interactive or non-interactive), the tuning map on Figure 6-4 depicts the response to a set-point change for various combinations of proportional and integral tuning. Movement from top to bottom of the map represents an increase in control algorithm gain or a decrease in proportional band. Movement from left to right on the map represents an increase in integral action. The left-hand column of graphs represent proportional action only. Moving to the right represents adding integral action, starting with minimum integral action as determined by the largest possible value for minutes per repeat, or the smallest possible value for repeats per minute.  

---

2. Minutes per repeat and repeats per minute are the most widely used measures of integral action. In general, we will speak of minutes per repeat in this book. Some manufacturers, as well as the Fieldbus Foundation, use *seconds per repeat* for integral action and *seconds* for derivative time.


Index

ABB 179
absquare 130–131
actuator 68, 132, 154, 161, 498
adaptive control 259
adaptive gain 260
adaptive tuning 259–260, 264, 266
additive 309, 325, 421
disturbance 431, 433
feedback 318, 320, 325, 341
feedforward 334–335, 346–347
stream 296
advanced control 269
advanced regulatory control 2–3, 152, 252, 269–270, 273, 275, 396, 473–474
air-fuel ratio control 295, 299, 333
air-to-fuel ratio 296, 299–302, 324, 487, 515–516
air-to-open 32
algorithm 101
synthesis 423, 429
analog
control 144, 285, 297, 426, 482
controllers 70, 106, 117, 122, 126–128, 134, 136, 146
equipment 332
hardware device 316
input 144, 152, 284, 290
output 32, 144–145, 152, 285, 482, 502
system 112, 287
world 332
analog-to-digital (A/D) converter 144
annealing furnace 343
anti-reset windup 140–141
 protection 136
 techniques 139
apparent
control loop 423–424
dead time 59–60, 71, 128, 181–182, 185–186
time constant 182, 186
approximate process model 422
approximate time constant 262
ratio adjustment 299
reset 96, 107
set ratio control 300, 305, 307, 517
switch 25, 132, 156, 285
to manual 133–134, 146
auto-tuning 259
auxiliary variable 449, 453, 469–470
averaging liquid-level control 247
back pressure regulator 72
backward link 155
basic regulatory control 152, 156
basis weight 421
batch 139
chemical process 343
controller 139
digester 363
switch 138–140
bias 4, 88–90, 95, 152, 246, 297
blending 296, 434, 441–442
Bode plots 21
boiler-drum control 346
box car integration 142
Bristol 386
bumpless tuning 135, 261, 375

control calculus 7–8, 10–11
cascade-loop 114, 116, 121, 228–229, 284–285
cascade-local switch 285
catalytic cracking 451
characterization 130, 132, 247
chart recorder 8, 97, 99, 102
chemical reactor 28, 114, 249
closed-loop test 162, 169, 177, 188–192, 212, 264
closed-loop tuning 263
Cohen-Coon 178–179
coincidence point 434
combining feedback and feedforward 318, 325, 327, 342, 414
combustion control 300
commissioning 151, 282, 288, 451
composition analyzer 64, 316, 342–343
controller 283, 343, 373, 385
loop 106
compressor station 366, 368
configuration option 111, 113–114, 135, 153, 314
cross-limiting 144
control control algorithm 4–5, 26, 29, 75, 106–107, 111, 116,
372, 420, 422–424, 426, 429, 440, 474, 497
cross-limiting control graph graph 91
crude switch 343
damped frequency 231
damping characteristics 174–175
damping factor 222, 224, 231–234
data vector 453
DCS 151–152, 156, 265, 299, 404, 426, 438, 452, 482
dead zone 132
dead-band 51
dead-time compensation 21, 186, 199
dead-time-to-time-constant 186, 199
Decay Ratio 228
decoupler 396, 404–405, 408
dead time 2, 384–385, 396, 400, 403, 409, 411–412, 420, 427, 450, 473–475
dependent variable 16–17, 39–40, 42, 90
deriv. time 107
derivative 8–9, 11, 67, 388
action 74, 127, 173–174, 188, 249
contribution 103, 113
gain 4, 128, 176
correction 102
derivative on error 114, 118, 249
derivative on measurement 114, 116, 118, 121, 144, 180, 249
derivative spike 112–115, 128
derivative time 5, 101, 121, 123, 125, 128, 169, 176, 179, 190
design conditions 39
design ratio 303
detune 384
device description 153
differential equations 7, 12, 14–15, 84
digital control 88, 127
digital control system 2, 32, 331
digital-based control system 331
digital-based controller 70
digital-to-analog (D/A) converter 145
direct digital control 147
direct effect 29, 386, 389, 391, 395
direct-acting 31–32, 44, 88, 102, 113, 115, 142, 221, 301
discharge temperature 282
discrete algorithm 148
distillation 434, 491
distillation column 71–72, 343–344, 348, 383, 428, 490
control 342–343, 395
distillation tower 28, 74, 245, 247, 283, 296, 329, 364, 373, 409
distributed control systems 26, 141
drum level 65–66, 219
behavior 12, 15
characteristics 46, 52, 56, 67, 329, 395
response 12, 56
system 18, 231–232
effective controller gain 123, 130, 248
effective integral time 248
efficiency 45–46, 300, 316–317, 342, 357
electric motorized valve 132
electrical stepping motor 147
elements 56
energy efficiency 299
engineering units 88, 142, 144, 148–149, 154, 485–486, 515–520
equal percentage 499–500, 513
valve characteristic 498
valves 498, 507, 509
ERF 358–359, 365
error signal 8, 10, 98, 102, 130, 247, 375
error-squared 130
algorithm 131, 247
EXACT 264–265
exothermic reactor 54–55, 106, 248
external feedback 288
external reset 137
external reset feedback 137–138, 358, 365–368, 372
fail-closed 32–33, 287, 356, 480
incremental form 144, 146–147
independent gains 4, 126–127, 174, 176, 434
independent variable 8, 16–17, 39–40, 42, 90
indeterminate process gain 182, 184
indirect effect 387, 390, 395
industrial process control 1, 18–19, 22, 63–64, 84, 383, 451
inherent 45, 48–49, 75, 441
initialization 153, 155–156, 287–288, 304–305, 404
inner loop 278–283, 288
installed characteristic 49
installed characteristics 49–50, 496, 512–513
integral 8, 67
  - gain 4, 174, 176
    - of the error 142, 166
    - response 98, 112, 171–172
    - time 5, 70–71, 97, 121, 123, 125, 137, 171, 174–176, 179, 184–185, 190, 193, 201, 228, 233–235, 248, 280, 341, 369, 384
  - tuning 135–136, 169
integral-error criteria 166–167
integral-mode contribution 135, 139
integral-only
  - control 121, 491, 497
  - controller 121, 496
  - mode 121
integrating process 53–54, 65, 73–76, 162, 183, 187, 218, 246, 428
interact 383
interacting 66
  - controller 122, 124–125, 172
  - form 123, 125–126, 128, 174, 178
  - form of PID 174
interchangeability 151
internal model
  - control 187, 420, 428, 432, 451
  - controller 188
inverse response 65–66, 419, 428, 449
inverted 396
  - decoupler 406–407
  - decoupler configuration 411
  - decoupling 397, 401–404, 407
ISA symbols 25, 275, 295
ISE 166–167
ITAE 166–167
ITSE 166–167
lag time 61, 331, 336–338, 340
Laguerre polynomials 473
Lambda tuning 187–188, 199, 266, 424
Laplace notation 97, 101, 122, 128
Laplace transform 4, 15, 20–21, 106, 118, 221, 422
Laplace, Pierre 15
lead-lag ratio 331, 336–338, 340
level arrest time 226, 230
level controller 66, 72, 76, 219–220, 231, 233, 245, 343
level loop 106, 247
level-control
  - loop 75, 187, 219–220, 226
  - system 66, 346
linear controller 130
linear valve 49, 69–70, 509, 512
Lipták 52, 369, 494
liquid-level
  - load 29–30
  - load change 29
  - load changes 41
  - loop diagrams 26
  - Lopez 178, 199
magnetic flow meter 70
magnitude ratio 21
manual reset 4, 89, 96, 246
manual to automatic 10, 133–134, 288
manual/automatic switching 133, 146, 284, 287
manufactured characteristics 509
maximum rate of change 147, 227–228, 231
McAvoy 388, 407
measurement 28
microprocessor-based
  - control 90, 114, 126, 141, 144, 146, 269
  - controllers 106, 128
  - system 98, 111, 134, 144, 148, 188, 284, 287, 297
min./repeat 107, 197
minutes per repeat 97–99, 124, 169–170, 174–176, 179, 188, 197, 222
model predictive control 269–270, 396, 420, 427, 450, 474
model-based
control 187, 270, 420, 425, 434
counters 419
modes of control 85
monotonic 45
move suppression 462–463
multi-level cascade control 282–283
multiple-input 449, 467
multiple-input multiple-output processes 66, 76
multiple-level cascade 287
multiple-output 449, 467
multiplicative
feedback 321, 323–325, 327
feedforward 335
feedforward control 46, 325, 334, 347
feedforward controller 341
natural frequency 232–233, 240
negative feedback 31–32
negative overshoot 141
noninteracting
controller 124
PID 175, 180
noninteractive 4–5, 126–127, 169, 174, 179, 201
controller 122, 124–125, 172, 178
form 123, 125, 153, 180
form of PID 174
nonlinear 45, 68, 162, 245, 259, 474
control 247
nonlinearity 46, 69, 71, 299
nonlinearization 130–132
nonrealizable 401, 405
nonselected controller 357–359, 368–372, 374, 376
non-self-regulating processes 53, 169
normalized value 142, 148–149, 168, 304, 517–519
Nyquist diagram 21
offset 86, 107, 191, 247, 297
on/off 84, 102, 263, 280, 311, 501
on-demand tuning 260, 262–263, 266
one-quarter decay 193, 227
one-twentieth decay 227
open-loop test 178, 183, 185, 187, 201, 262–263
tuning from 162, 177
open-loop unstable 52, 55, 106
operating point 16, 44–45, 94–95, 136–139, 177, 188, 251, 259–261, 263, 329, 343, 363, 469–470, 474
optimization 3, 269, 450–451, 473, 496
outer loop 278–279, 283
outflow arrest time 226–227, 231
output bias 88–89, 92–94, 96, 99, 141
adjustment 94
overdamped 223
response 170
overshoot ratio 165, 199, 226, 233
P 85–86, 107, 115, 137, 175, 178–179, 190, 193, 197, 201, 212, 215, 226, 453, 460, 517, 519
P/Ti ratio 199, 201, 217
P&ID 22–24, 275, 295
pairing 384–386, 388, 391–392, 394–396, 412
paper machine 421, 428
partial decoupling 401, 405, 408
pass-through 368–370, 372, 374
performance criteria 162–164, 266
petrochemical industry 395
petroleum refinery 343, 451
heater 409
pH control 131, 434
phase lag 121
pipeline industry 366
piping and instrumentation diagram 22
plant 29
plastic extrusion 501, 503
position 99, 143
position-mode 318, 372
algorithm 261, 319
positive feedback 31, 44, 395
power generation industry 26, 30
pre-act 107
predicted value 104–105, 437, 450, 453, 456, 461, 463, 470
prediction horizon 453, 456, 458, 462, 465
Index 565

predictive functional control 434
predictive nature 105
predictor 332, 428
pressure control 54, 72, 74, 489–492, 494
loop 72, 74–75
pressure reducing station 72, 493
pressure regulator 72
primary 275
principal disturbance 315
process 1, 29
behavior 1, 66
characteristics 54, 66, 71, 189, 260, 335
control 1, 3, 7–8, 18–19, 22, 39–40, 63–64, 83–84, 100, 106, 126, 149, 152, 166, 249, 270, 363, 383, 385, 451, 489, 496, 503
control engineer 8, 14, 250, 451
controller 84, 96, 107, 363–364, 485, 487–488, 492, 494, 497
design engineer 39
dynamic behavior 76, 342
dynamic characteristics 52, 71, 327
dynamics 75, 128, 161, 316, 327–328, 333, 342, 395, 400
engineer 22, 39–40, 107
flow diagram 26
graph 42–45, 76, 90–92, 94–96
heater 45, 71, 299, 315, 324, 327, 331, 334–335, 354, 470, 482
input 455
response 19–20, 63, 161, 171, 176–178, 184, 262, 340, 399, 434
time constant 178, 182, 186, 262, 407, 422, 426
processing cycle 142
process-model-based control 434
product quality 2, 342, 411, 491
prop. band 107
proportional
contribution 115
control 86, 102–103, 174
controller 86–88, 105
gain 4, 174–176
kick 115
proportional mode 85–86, 98, 107, 115, 117–118, 121, 130, 217, 247
contribution 102, 117
on error 121
on measurement 112, 114, 153, 164
proportional on error 117
proportional on measurement 117
proportional response 98, 112, 115–116, 128, 180
proportional-band 90
proportional-only
control 246
controller 4, 89, 92, 100, 105, 246
proportional-on-measurement configuration option 121
proportional-plus-derivative control 103
proportional-plus-integral controller 100
pseudo dead time 63
pseudo time constant 63
pulp and paper industry 188, 363
pulse-response model 471–473
pure dead time 56, 59, 421
pure delay 331
PV tracking 133
quadratic programming 470
quarter-amplitude decay 165, 174, 191, 197, 222
quarter-wave damping 165
quarter-wave decay 165
quick-opening 48
quick-opening valve 69
ramp generator 102
ramped 116
ramping 98
rangeability 413, 479, 496, 498–499, 507
rate action 101, 107
rate of change 8, 15, 17, 86, 101–102, 104, 116, 147, 173, 227–228, 234, 246, 436, 469
rate of change of variable 449
realizability 400, 404, 406–408
realizable 400–401, 404, 406, 429
decoupling elements 401, 404
reboiler 245, 364, 373, 383, 492
reference trajectory 453, 462–463, 473
reliability 283, 342, 383
relative gain 387–388, 391–394, 408
analysis 386, 395–396
array 387–388, 391, 393–395
matrix 394
relay method 263
repeats per minute 98–99, 107, 169, 171, 174, 188
reset 96, 98, 107, 171–172, 174–175, 191, 250, 260, 341
action 192
preload 141
windup 136–137, 139, 151, 154
reverse action 32, 287, 356, 419
reverse-acting 31–32, 44, 88, 94, 142, 153, 301, 371, 480
reverse-acting controller 31–32, 89, 102, 104, 113, 115, 118
RGA 387, 395–396
rise time 164–166, 180, 182
robustness 407
root locus 21
Routh 84
runaway 54–55, 76, 248–249
SAMA 25–27, 275, 295
SAMA symbols 25–26, 275, 295
sample period 371, 456, 460–462
sampled data step-response model 453
sampling period 186
sampling time 136, 439
scheduled tuning 248, 260–261, 263, 375, 427
Scientific Apparatus Makers Association 25
secondary 275, 295
seconds per repeat 169
selected controller 138, 369, 371–372, 374, 376
selector 2, 137, 354–355, 363, 374, 376, 483, 485, 487, 494
control 353
device 138
switch 117, 353, 360, 373
self-tuning 259–260, 262, 265
sensor-transmitter 29
Shinskey 67, 74, 125, 139, 365, 385, 396, 490, 492, 500
shrink-and-swell 65, 245
side lag 61–63
signal generator 97
signals 8
simplified 396
sinusoidal disturbance 231, 234
SISO process 452–453, 461, 465
Smith predictor 420–423, 426–429, 434, 438
softwiring 144
split-range control 479
S-shaped curve 58–59, 61, 63, 177
stability 84, 171, 249, 266, 404–408, 434, 451
stabilizing direction 172
states 84
status bits 146, 155–156, 288
steady-state
characteristics 40, 76
error 171, 341
gain 4, 18, 329, 331, 386–388, 394–395
offset 74, 92–95, 100–101, 105, 107, 170, 246
steam generators 65, 482
stem position 48, 50–52, 68, 156, 413
step- and pulse-response 471–472
step response 59, 63, 177, 456
stick-slip 50–51
stiction 50
stoichiometric 299
superposition 337
sustained oscillation 189–191
tank holdup time 220, 224, 230
tank time constant 220
Taylor Instrument Co. 179
temperature control loop 22, 31, 70–71, 74, 139, 282, 411, 479
thermostat 84, 94, 280, 282, 311–312
throughput 2, 39, 71, 192, 252, 321, 327, 342
time 8
time delay 19, 59, 217, 332, 371, 426
time proportioning
control 479, 501–502
controllers 502–503
tower flooding 364, 373
transducer block 154, 156
transport lag 20, 56
trial-and-error tuning 162, 169, 173, 192, 218
tube temperature 354–357, 359–362, 469–470
controller 354–357, 362
tuning 2, 161
aids 265
log 197, 212, 215, 251
map 169–172
the decoupler 408
values 124, 162, 171, 178, 184, 186, 250, 261–262
two-degree-of-freedom controller 117
two-position control 84
ultimate effect 301, 387, 390–391, 395
ultimate gain 189, 191, 264
valve 28
actuator 46, 50–51, 71, 290
characteristics 48, 69, 76, 251, 283, 413, 507
position controller 364, 490, 492, 494, 496–497
positioner 32, 46, 52, 68, 150, 154, 156, 219, 251, 283, 480
rangeability 479, 496, 498, 507
variable pairing 384–385, 392
velocity (incremental) 372, 376
velocity mode 144
vortex meter 70
water supply systems, hot or chilled 494
wild flow 295–297, 299, 303
Wobbe index 316
Ziegler-Nichols 178–179, 184, 245, 426