THIRD EDITION

# TROUBLESHOOTING PROCESS PLANT CONTROL

## A PRACTICAL GUIDE TO AVOIDING AND CORRECTING MISTAKES

NORMAN P. LIEBERMAN

WILEY

# *Troubleshooting Process Plant Control*

# Troubleshooting Process Plant Control

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- *Troubleshooting Refinery Operations*—PennWell Publications
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Copies of the first three texts are best ordered from the publishers, but may be ordered through us. E-mail (norm@lieberman-eng.com). *Troubleshooting Refinery Operations* (1980) has been incorporated into *Troubleshooting Process Operations*.

# *Troubleshooting Process Plant Control*

## *A Practical Guide to Avoiding and Correcting Mistakes*

Third Edition

**Norman P. Lieberman**
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WILEY

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# *Dedication to Third Edition*

Time goes on. And life goes on. The seasons progress from fall, to winter, into spring, same with me. I was young, then progressed to middle age. Became old. And then I peaked at age 82, and became younger again!

How did this happen?

My wife and inspiration, Liz, explained it to me, “Norm, you had a chance to retire ten years ago. Now it’s too late.”

I guess then I’ll have to go on to the end, Liz and I together. Old process engineers never die; they just fade away.

March, 2024

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# *About the Author*

**Norman P. Lieberman** has been troubleshooting refinery and chemical plant process equipment since 1964. He began work at American Oil as a process engineer at their Indiana Refinery. Lieberman designed and operated sulfur plants, alkylation units, LPG fractionators, cokers, distillation towers, and vacuum systems for Amoco until 1980. He was next employed at the Good Hope Refinery in New Orleans where he worked on their polymerization unit, MTBE plant, crude unit, naphtha reformer, and hydrodesulfurization facilities.

In 1985, Lieberman worked for Good Hope, troubleshooting gas field compression, dehydration, and treating problems for their Loredo, TX natural gas production facilities. In 1988, he became a consultant for petrochemical and refinery process equipment problems.

Approximately 24,000 engineers and operators have attended his 1030 in-house troubleshooting seminars and webinars. Lieberman has a degree in Chemical Engineering, 1964, from Cooper Union. He lives in New Orleans with his wife, Liz, also a chemical engineer.

Norm runs every day for three miles and plans to do so until the end.

# *Preface to the Third Edition*

I have been practicing process engineering for 59 years. Mainly, as a refinery field troubleshooter for distillation operations, vacuum systems, fired heaters, waste water strippers, compressors, and pumps. The majority of the malfunctions I discover are not due to faulty equipment design, mechanical failure, or operator error. The big problem is with process control and measurement of process variables.

I always explain during my process equipment troubleshooting seminars, which I've instructed since 1983, that the process control engineer is the most important person in the plant. I was sure of that in 1983 and am equally sure as I write these words in 2024.

The problem that the refining, petrochemical, and chemical fertilizer industry has is that the university course of study for process control engineers is worse than bad. It's irrelevant! In 1979 at Northwestern, and in 1983 at LSU, I found this out personally, having been ejected from both institutions after 1 day as an instructor. My conception of the training required to be an effective process control engineer, being at odds with that of both universities.

Process control has little to do with math, or computers, or Laplace transforms. It's about understanding the following:

- How instruments work.
- How variables of temperature, pressure, level, flows, and composition are measured in the field.
- How controls interact with process equipment.
- How unit operators interact with controls.
- The tendency of instrumentation to be trapped in a "positive feedback loop."
- How process equipment itself works from a chemical engineering perspective.
- How to make field measurements.

I wasn't particularly knowledgeable about process control until 1974, even though I had been employed by Amoco Oil for 10 years. But in 1974, I worked as an operator for 4 months, during a strike in Texas City. Then again in 1980, there was an even longer strike, part of which I worked as the panel operator on a sulfur recovery and amine unit. Afterward, I was fairly competent to tackle a variety of process control issues.

Based on my subsequent 44 years of experiences, I have developed the following advice for young process control engineers:

"The price we pay for success is the willingness to risk failure." Michael Jordan, Chicago Bulls.

You can email me with questions at norm@lieberman-eng.com

# Introduction

Troubleshooting process plant control first requires an understanding of a wide variety of malfunctions that may develop in measuring variables such as:

- Levels
- Temperatures
- Pressures
- Flows
- Compositions

Certainly, if you cannot measure a level or pressure, you can't expect to control it.

Second, the console operator or process engineer must understand how the control valves and the signals to open and close the valves actually work. For instance, did you know that the valve position shown in the control center does not at all represent the actual valve position? It represents what the control valve position is supposed to be.

To troubleshoot process control problems, the operator, or engineer, has to understand the relationship of the controls to the individual process. This means, you will have to get to know the unit and how it works. This is the most difficult part of the job of understanding process plant control.

Many apparent control problems are, in reality, process problems. But, on the other hand, after 59 years of troubleshooting refinery process problems, I am quite sure that the most common sort of malfunctions I have encountered are related to the inability to measure a level, temperature, or flow correctly—and also to have a control valve respond in the manner needed to achieve the desired operational change.

## AUTO VERSUS MANUAL

When you see that the console operator is running a control loop on "manual" rather than in "auto," that is an indication that something is wrong with the field measurement of the variable, or with the control logic. Is it a metering problem, a sensor that is fouled, or a variable that is over-ranged? Perhaps, the variable is caught up in a "positive feedback loop"? Control loops are supposed to run in auto, and you should not accept loops that run in manual, as representing an acceptable mode of operation. Sooner or later, such broken (i.e., manual) control loops will slip out of an acceptable operating range.

## GAIN, RESET, AND ADVANCED COMPUTER CONTROL

This text does not deal with the time aspects of optimizing the relationship between variables. Typically, the console operator is far less concerned about how fast operating parameter returns to its set point, than if a particular variable is moving in the right direction, so that he can safely bring his products back on spec.

Advanced computer control is largely irrelevant to my work in field troubleshooting refinery and petrochemical plant process control problems. I cannot conceive as to why process control engineering is so often taught in universities as if it is a form of higher mathematics. Even more detrimental to unit operations is the perception by plant management and staff engineers that advanced computer controls are actually being utilized on the operating units, when in reality the console operators are struggling to run critical control loops on auto, without getting caught up in a dangerous positive feedback loop. I never understood anything about Laplace transforms in school, and I am certainly too old to start learning now.

The real problems with hydrocarbon processing that I first encountered in 1965 are 90% unchanged in my work as a chemical engineer in 2024.

# 1

# Learning from Experience

An old Jewish philosopher once said, "Ask me any question, and if I know the answer, I will answer it. And, if I don't know the answer, I'll answer it anyway." Me too. I think I know the answer to all control questions. The only problem is, a lot of my answers are wrong.

I've learned to differentiate between wrong and right answers by trial and error. If the panel board operator persistently prefers to run a new control loop that I've designed in manual, if he switches out of auto whenever the flow becomes erratic, then I've designed a control strategy that's wrong. So, that's how I've learned to discriminate between a control loop that works and a control strategy best forgotten.

Here's something else I've learned. Direct from Dr. Shinsky, the world's expert on process control:

- "Lieberman, if it won't work in manual, it won't work in auto."
- "Most control problems are really process problems."

I've no formal training in process control and instrumentation. All I know is what Dr. Shinsky told me. And 59 years of experience in process plants has taught me that's all I need to know.

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## LEARNING FROM PLANT OPERATORS

My first assignment as a Process Engineer was on No. 12 Pipe Still in Whiting, Indiana. This was a crude distillation unit. My objective was to maximize production of gas oil, as shown in Figure 1-1. The gas oil had a product spec of not more than 500 ppm asphaltenes. The lab required half a day to report sample results. However, every hour or two, the outside operator brought in a bottle of gas oil for the panel board operator. The panel operator would adjust the wash oil flow, based on the color of the gas oil.

While plant supervision monitored the lab asphaltene sample results, plant operators ignored this analysis. They adjusted the wash oil rate to obtain a clean-looking product. The operators consistently produced a gas oil product with 50–200 ppm asphaltenes. They were using too much wash oil, and the more the wash oil used, the lower the gas oil production.

I mixed a few drops of crude tower bottoms in the gas oil to obtain a bottle of 500 ppm asphaltene material. I then instructed the panel board operators as follows:

- If the sample from the field is darker than my standard bottle, increase the wash oil valve position by 5%.
- If the sample of gas oil from the field is lighter than my standard, decrease the wash oil valve position by 3%.
- Repeat the above every 30 minutes.

The color of gas oil from a crude distillation unit correlates nicely with asphaltene content. The gas oil, when free of entrained asphaltenes, is pale yellow. So, it seems that my procedure should have worked. But it didn't. The operators persisted in drawing the sample every 1–2 hours, not every 30 minutes like I had instructed.

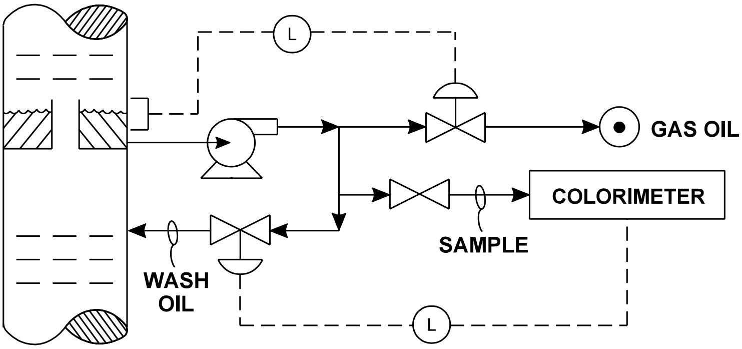

*Figure 1-1* Adjusting wash oil based on gas oil color

So, I purchased an online colorimeter. The online colorimeter checked whether the gas oil color was above or below my set point. With an interval of 10 minutes, it would move the wash oil valve position by 1%. This never achieved the desired color, but the gas oil product was mixed in a tank. The main result was that gas oil production was maximized, consistent with the 500 ppm asphaltene specification.

One might say that all I did was automate what the operators were already doing manually and that all I accomplished was marginally improving an existing control strategy by automating the strategy. But, in 1965, I was very proud of my accomplishments. I had proved, as Dr. Shinsky said, "If it does work on manual, we can automate it."

## LEARNING FROM FIELD OBSERVATIONS

Fifty-three years ago, I redesigned the polypropylene plant in El Dorado, Arkansas. I had never paid much attention to control valves. I had never really observed how they operate. But I had my opportunity to do so when the polypropylene plant was restarted.

The problem was that the purchased propylene feed valve was too large for normal service. I had designed this flow for a maximum of 1600 BSD, but the current flow was only 100 BSD. Control valve response is quite nonlinear. Nonlinear means that if the valve is open by 5%, you might get 20% of the flow. If you open the valve from 80 to 100%, the flow goes up by an additional 2%. Nonlinear response also means that you cannot precisely control a flow if the valve is mostly closed. With the flow only 20% of the design flow, the purchased propylene feed was erratic. This resulted in erratic reactor temperature and erratic viscosity of the polypropylene product.

The plant start-up had proceeded slowly. It was past midnight. The evening was hot, humid, and very dark. I went out to look at the propylene feed control valves. Most of the flow was coming from the refinery's own propylene supply. This valve was half open. But the purchased propylene feed valve was barely open. The valve position indicator, as best I could see with my flashlight, was bumping up and down against the "C" (closed) on the valve stem indicator.

The purchased propylene charge pump had a spillback line, as shown in Figure 1-2. I opened the spillback valve. The pump discharge pressure dropped, and the propylene feed valve opened to 30%. The control valve was now operating in its linear range.

Now, when I design a control valve to handle a large reduction in flow, I include an automated spillback valve from pump discharge to suction. The spillback controls the pump discharge pressure to keep the Flow Recorder Control (FRC) valve between 20 and 80% open. Whenever I sketch this control loop, I recall that dark night in El Dorado. I also recall the value of learning even the most basic control principles by personal field observations.

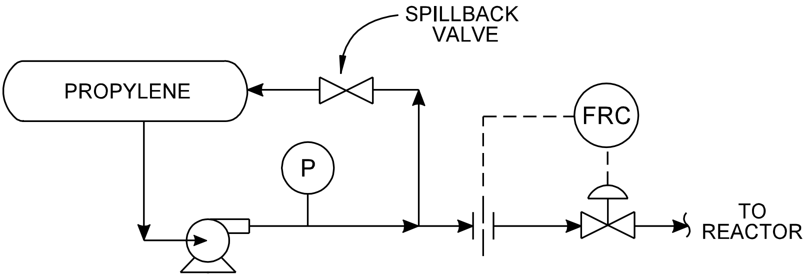

***Figure 1-2*** *Opening spillback to keep the FRC valve in its linear operating range*

## LEARNING FROM MISTAKES

Adolf Hitler did not always learn from his mistakes. For example, he once ordered a submarine to attack the Esso Lago Refinery in Aruba. The sub surfaced in the island's harbor and fired at the refinery. But the crew neglected to remove the sea cap on the gun's muzzle. The gun exploded and killed the crew.

I too had my problems in this refinery. The refinery flare was often very large and always erratic. The gas being burned in the flare was plant fuel. The plant fuel was primarily cracked gas from the delayed coker, supplemented (as shown in Fig. 1-3) by vaporized LPG. So much fuel gas was lost by flaring that 90% of Aruba's LPG production had to be diverted to fuel, via a propane vaporizer, to maintain refinery fuel gas pressure.

I analyzed the problem based on the dynamics of the system. I modeled the refinery's fuel consumption versus cracked gas production as a function of time. The key problem, based on my computer system dynamic analysis, was the cyclic production of cracked gas from the delayed coker complex. My report to Mr. English, the General Director of the Aruba Refinery, concluded:

1. The LPG vaporizer was responding too slowly to changes in cracked gas production from the delayed coker.
2. The natural log of the system time constants of the coker and vaporizer was out of synchronization.
3. A feed-forward, advanced computer control based on real-time dynamics would have to be developed to bring the coker vaporizer systems into dynamic real-time equilibrium.
4. A team of outside consultants, experts in this technology, should be contracted to provide this computer technology.

Six months passed. The complex, feed-forward computer system was integrated into the LPG makeup and flaring controls shown in Figure 1-3.

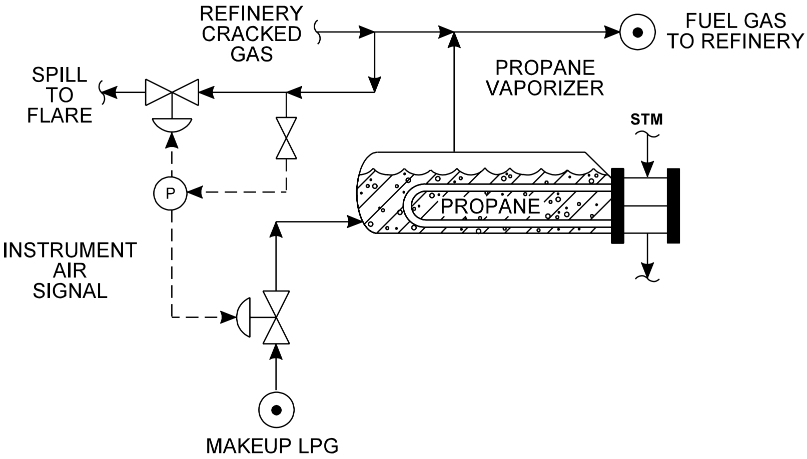

***Figure 1-3*** *Unintentional flaring caused by malfunction of the LPG makeup control valve is an example of split - range pressure control*

Adolf Hitler would have been more sympathetic than Mr. English. The refinery's flaring continued just as before. Now what?

Distressed, discouraged, and dismayed, I went out to look at the vaporizer. I looked at the vaporizer for many hours. After a while, I noticed that the fuel gas system pressure was dropping. This happened every 3 hours and was caused by the cyclic operation of the delayed coker. This was normal.

The falling fuel gas pressure caused the instrument air signal to the LPG makeup valve to increase. This was an "Air-to-Open" valve (see Chapter 13), and more air pressure was needed to open the propane flow control valve. This was normal.

But, the valve position itself did not move. The valve was stuck in a closed position. This was not normal.

You will understand that the operator in the control room was seeing the LPG propane makeup valve opening as the fuel gas pressure dropped. But the panel board operator was not really seeing the valve position; he was only seeing the instrument air signal to the valve.

Suddenly, the valve jerked open. The propane whistled through the valve. The local level indication in the vaporizer surged up, as did the fuel gas pressure. The flare valve opened to relieve the excess plant fuel gas pressure and remained open until the vaporizer liquid level sank back down, which took well over an hour. This all reminded me of the sticky side door to my garage in New Orleans.

I sprayed the control valve stem with WD-40, stroked the valve up and down with air pressure a dozen times, and cleaned the stem until it glistened. The next time the delayed coker cycled, the flow of LPG slowly increased to catch the falling fuel gas pressure, but without overshooting the pressure set point and initiating flaring.

My mistake had been that I had assumed that the field instrumentation and control valves were working properly. I did not take into account the probability of a control valve malfunction. But, at least, I had learned from my mistake, which is more than you could say for Adolf Hitler.

## LEARNING FROM THEORY

Northwestern University has an excellent postgraduate chemical engineering program. I know this because I was ejected from their faculty. I had been hired to present a course to their graduate engineers majoring in process control. My lecture began:

> "Ladies and gentlemen, the thing you need to know about control theory is that if you try to get some place too fast, it's hard to stop. Let's look at Figure 1-4. In particular, let's talk about tuning the reflux drum level control valve.
>
> Do I want to keep the level in the drum close to 50%, or doesn't it matter? As long as the level doesn't get high enough to entrain light naphtha into fuel gas, that's okay. What is not okay is to have an erratic flow feeding the light naphtha debutanizer tower.
>
> On the other hand, if the overhead product was flowing into a large feed surge drum, than precise level control of the reflux drum is acceptable.
>
> In order for the instrument technician to tune the level control valve, you have to show him what you want. To do this, put the level valve on manual. Next, manipulate the light naphtha flow to permit the level swings in the reflux drum you are willing to tolerate. But you will find that there is a problem. If you try to get back to the 50% level set point quickly, you will badly overshoot your level target.

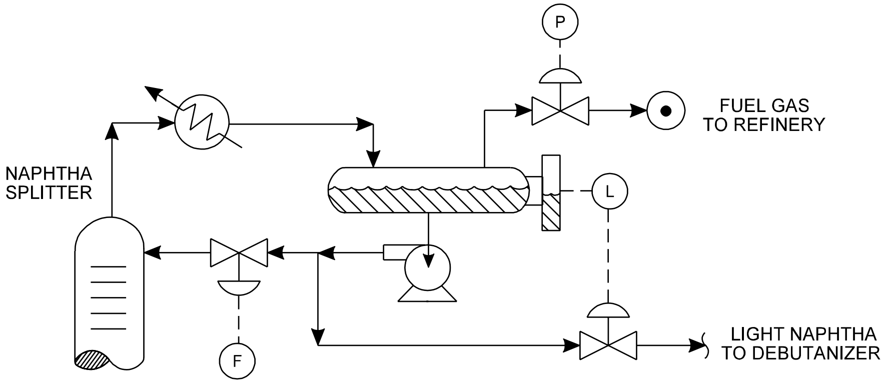

***Figure 1-4*** *Tuning a level control valve depends on what is downstream*

> If you return slowly to the set point, it's easy to reestablish the 50% level target. However, the level will be off the target for a long time.
>
> In conclusion, ladies and gentlemen, tuning a control loop is a compromise between the speed at which we wish to return to the set point and our tolerance to overshooting the target. To establish the correct tuning criteria, the control loop is best run on manual for a few hours by the Process Control Engineer. Thank you. Class adjourned for today."

My students unfortunately adjourned to Dean Gold's office. Dean Gold lectured me about the student's complaints.

> "Mr. Lieberman, did you think you were teaching a junior high school science class or a postgraduate course in process control?"

And I said, "Oh! Is there a difference?"

So that's how I came to be ejected from the faculty of Northwestern University after my first day of teaching.

## LEARNING FROM RELATIONSHIPS

My ex-girlfriend used to tell me, "Norm, the reason we get along so well is that I give you a lot of positive feedback." From this, I developed the impression that positive feedback is good, which is true in a relationship with your girlfriend. But when involved in a relationship with a control loop, we want negative feedback. Control logic fails when in the positive feedback mode of control. For example:

- **Distillation**—As process engineers and operators, we have the expectation that reflux improves fractionation, which is true, up to a point. That point where more reflux hurts fractionation instead of helps is called the "incipient flood point." Beyond this point, the distillation tower is operating in a positive feedback mode of process control. That means the tray flooding reduces tray fractionation efficiency. More reflux and more reboiler heat simply make the flooding worse.
- **Fired Heaters**—Increasing furnace fuel should increase the heater outlet temperature. But if the heat release is limited by combustion air, then increasing the fuel gas will reduce the heater outlet temperature. But as the heater outlet temperature drops, the automatic control calls for more fuel gas, which does not burn. As the heater outlet temperature continues to fall, because combustion is limited by air, the outlet temperature drops further. The heater automatic temperature control loop is now in the positive feedback mode of control. As long as this control loop is on auto, the problem will feed upon itself.

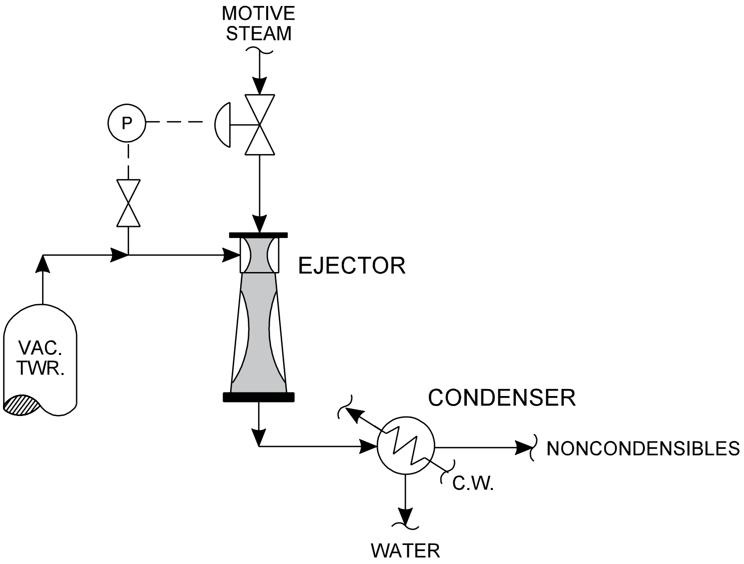

***Figure 1-5*** *Too much steam flow causes a loss in vacuum*

- **Vacuum Ejector**—Some refineries control vacuum tower pressure by controlling the motive steam flow to the steam ejector. As the steam pressure and flow to the ejector increases, the ejector pulls a better vacuum, as shown in Figure 1-5, but as the steam flow increases, so does that load on the downstream condenser. As the condenser becomes overloaded, the ejector discharge pressure rises. At some point, the increased discharge pressure adversely affects the ejector's suction pressure. A further increase in motive steam will make the vacuum worse, instead of better. As the vacuum gets worse, the control loop calls for more steam. Having now entered the positive feedback mode of control, the problem feeds upon itself.

Many control loops are subject to slipping into a positive feedback loop. The only way out of this trap is to switch the controls to manual and slowly climb back out of the trap. Once you guess (but there is no way to know for sure) that you are in the safe, negative feedback mode of control, you can then safely switch back to automatic control.

## NORMAL PURPOSE OF CONTROL LOOPS

Typically, a control loop is tuned to achieve two objectives:

1. To return a variable to its set point as fast as possible.
2. To avoid overshooting the set point.

If a heater outlet set point is at 700°F, and it is currently running at 680°F, the firing rate should increase. However, if the firing rate increases too fast, the heater outlet may jump past the set point to 720°F.

Tuning a control loop is meant to balance the instrument, "gain and reset," to balance these two contradictory objectives.

The balance between gain and reset (i.e., instrument tuning) is not the main object of this text. Only rarely have I seen a panel board operator complain about this problem.

Another purpose of control is to optimize process variables. This is an advanced control that attempts to optimize certain variables. This is also not the sort of problem that the panel operator would be concerned about. An example of advanced control would be to optimize the ratios of several pumparounds, versus the top reflux rate, for a refinery crude distillation tower. For the units I work on, such advanced computer control is rarely used, or has been simplified, so that it is not much different than ordinary closed-loop control.

## MANUAL VERSUS AUTO

In reality, the main complaint about control loops that are communicated to me by operators is that the controller will not work in the automatic mode of control and that the operators are forced to run the control loop in manual. This greatly increases and complicates their work.

To a large extent, this text examines why control loops are forced to run in the manual mode. A few of the reasons are the following:

1. The control loop is trapped in a "positive feedback loop." This is often a dangerous situation.
2. There is no direct relationship between the variable being controlled and the response of the control valve. This is typically a design error.
3. The facility that measures the process parameter in the field is not working correctly. This represents the majority of control problems that I have seen.
4. The bypass valve is open around the control valve.
5. The control valve is running too far closed because it is oversized or badly eroded.
6. The control valve is running too far open because it is too small, or its port size is too small, or an isolation gate valve in the system is partly closed.
7. The control valve's "Hand Jack" has been left engaged. Thus, the control valve cannot be manipulated from the computer console or panel. The hand jack is a mechanical device, used to manually move the control valve in the field.
8. The control valve is stuck in a fixed position.

9. The air signal connection to the diaphragm that moves the control valve has come loose.
10. The diaphragm is leaking, so the sufficient instrument air pressure cannot be applied to the control valve mechanism to force it to move.
11. The air signal is connected on the wrong side of the diaphragm.

## PROCESS CONTROL NOMENCLATURE

The reader who is new to process plant vocabulary may wish to briefly skip to the glossary at the end of this book. I have assembled a list of "Process Control Nomenclature Used in Petroleum Refineries and Petrochemical Plants." As in any other industry, your coworkers will have developed a vocabulary of their own and will assume you understand the terms they employ. To an extent, in the following chapters of this text, I have also made a similar assumption.

A brief review of these terms may make it easier for you to communicate with some of your coworkers.

# 2

# Process Control Parameter Measurement

I mentioned in Chapter 1 that I was ejected from the faculty of Northwestern University after teaching a single class. This was not the end of my academic career. I was also an instructor at Louisiana State University. Dr. Dillard Smythe had hired me on a trial basis to conduct a process control course for undergraduate chemical engineers. My course was excellent, but judge for yourself.

“Ladies and gentlemen. Welcome to Process Control 101. The course is divided into two segments:

- **Segment One**—Measuring Process Control Parameters
- **Segment Two**—Designing Control Loops for Process Parameters”

We must measure the parameter before we can control the parameter. That’s why we will study measurement first.

The Nazi army was able to initially defeat the allied armies in World War II because of the superior use of tanks. It wasn’t that the German tanks were better than the Allied tanks. It was that the Germans had excellent FM radios in their tanks. The data supplied from forward units enabled senior commanders to coordinate the Panzer attack. That is, the limiting factor for any control strategy is the quality of the data. “Garbage in; garbage out.”

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I plan to discuss measurement techniques and problems for the following process parameters:

- Liquid levels
- Temperature
- Pressure
- Differential pressure
- Flow

My experience is limited to continuous processes but excludes solids and high-viscosity fluids. So let's limit our study accordingly. My experience in the petrochemical and refining industry has taught me that most control problems are a consequence of the improper parameter measurement, most especially levels.

## HOW ARE LIQUID LEVELS MEASURED?

Most liquid level measurements are made by a level-troll. The level-troll is served by two pressure transducers. A pressure transducer is a mechanical device that converts a pressure in an electronic signal. Car engines have a transducer to measure the engine oil pressure.

Figure 2-1 shows the arrangement of the pressure transducers, one at the top and another at the bottom of the level-troll. The level-troll is a pipe a few feet long. The difference in the electrical output between the dual pressure transducers is proportional to the difference in pressure between the top and bottom of the level-troll. This delta *P* is caused by the head of liquid in the level-troll. The electrical output generated by this pressure difference is called the "milliamp output of the level indicator." The level indication is really a

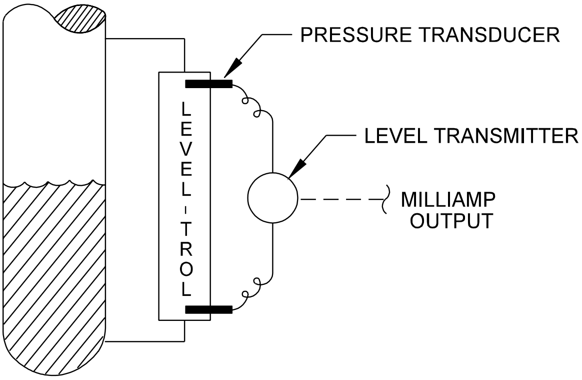

***Figure 2-1*** *Measuring levels by sensing liquid head pressure*

measure of the head pressure in the level-troll pipe. Head pressure, DP, is calculated as DP = (Height) • (Density).

The level-troll cannot discriminate between height (i.e., liquid level) and density (i.e., specific gravity). The Process Control Engineer has to specify the liquid's density or specific gravity (SG). Let's say the specified SG = 0.80 and the calculated level from the delta *P* output from the level-troll is 45%. This 45% level is displayed on the panel in the control room.

The 45% level multiplied by the specific gravity of 0.80 SG results in a delta *P* of 36 units of differential pressure:

$$(45\%) \bullet (0.80) = 36 \text{ units of delta } P$$

But now, a new situation has developed. The feed to the vessel has become lighter, the bottom's product has become hotter. Or the liquid in the vessel is aerated. For some reason, the specific gravity has dropped from 0.80 to 0.60 SG.

Assume that the delta *P* output from the level-troll is constant at 36 units of differential pressure. Thus, the indicated level in the control room is still 45%. But the real level has increased to 60%. That is, the 60% level multiplied by the specific gravity of 0.60 SG results in a delta *P* of 36 units of differential pressure:

$$(60\%) \bullet (0.60) = 36 \text{ units of delta } P$$

Thus, the level in the vessel has gone up by one-third, but the panel level indication has not changed. A reduction in fluid density will therefore result in an automatic increase in the level in a vessel as long as the level control loop is in automatic. This precise problem has resulted in explosions and fires and death and disaster throughout the process industry.

One way that we deal with this problem in refineries is with radiation level detection, which is expensive, complex, and potentially dangerous because of problems with handling radioactive materials. We could also mathematically correct the indicated level for changes in density by a closed-loop computer control. But this can only be done if we know the new fluid density. In cases where the density has dropped because of aeration, which is a common problem, we do not know the aerated density in the bottom of the vessel.

So, in conclusion, what is the answer? The answer is—there is no answer! But it is certainly something for the process control engineer to worry about. Fifteen people were killed at the BP Refinery in Texas City in 2005 because no one understood this problem in a naphtha fractionation tower, which erupted gasoline from its relief valve.

I call the S.G. used to calibrate the level indicator, "The Calibration Specific Gravity."

## HOW ARE TEMPERATURES MEASURED?

I always keep a spare thermocouple at home. I need it in case the pilot light fails on my water heater. It works like this:

- The pilot light flame heats the end of the thermocouple.
- There are two wires of different metallurgy, twisted together to form a “junction.”
- When the junction is heated, some of the energy of the flame generates a direct electrical current flow of a few volts.
- This voltage is sufficient to open a solenoid valve, permitting gas flow to the pilot light burner.

If the thermocouple malfunctions, you can keep the solenoid valve open with a 9-volt battery. But, perhaps, this is not one of my better ideas.

One would think that, except for the thermocouple burning out, temperature indication is reliable and may be used with confidence by the Process Control Engineer. After all, the thermocouple wires are protected by the thermowell. This is a thick pipe made of high-grade stainless steel, sealed at the process end. Unfortunately, such temperature indication has a whole range of problems.

Deposits on the surface of the thermowell will insulate the junction of the thermocouple wires. The external portion of the thermowell assembly radiates some heat. The heat loss from the thermowell is normally of no consequence. But if a portion of the thermowell inside the process vessel is fouled, the entire TI assembly will cool. I have observed temperature readings inside vessels operating at 800°F suppressed by 20–30°F because of coke formation around the thermowell. To verify this problem, temporarily wrap insulation around the external portion of the thermowell assembly. If the TI reading increases by 5–10°F, the thermowell is fouled and reliable temperature measurements cannot be determined.

I was working for Exxon on a vacuum tower problem recently. The tower feed temperature was 760°F. Eight feet above the feed nozzle, in the flash zone, the temperature of the rising vapors was only 680°F. What happened to the 80 °F? The answer was “Nothing.” Above the flash zone thermowell, there was the gas oil product draw-off pan. The pan has a drain hole, so the cool liquid accidentally fell onto the thermowell. I checked the vessel’s external skin temperature around the entire flash zone. It was all quite uniform and consistent with the 760°F feed temperature. Any single temperature indication in a large diameter vessel may not mean too much. The Process Control Engineer should specify several TI points at the same elevation. This was the practice for the 10-feet-diameter hydrocracker reactors designed for the Amoco facilities in Texas City.