2006 Engineering Design Seminar

Instrumentation

Daniel O’Leary

UOP LLC

uop
A Honeywell Company

EDS 2006/Inst-1

Introduction

- **Instructor**
  - **Name** Daniel J. O'Leary
  - **Background/experience**
- **Session Objectives**
  - **Obtain basic understanding of concepts and fundamentals applied to instrumentation**
  - **Review from simple primary elements to complex control systems existing in Industry**
  - **Convey UOP's process control philosophy**

UOP A Honeywell Company EDS 2006/Inst-2

The session objectives start with a review of the basic concepts of the feedback control loop. The feedback control loop is sectioned into individual components with a discussion on the theory of the more common elements.

Several factors affect control loop performance. Among these are transmitter performance, control valve performance, and the speed of response of today's Digital Control Systems (DCS). Each of these topics will be covered in detail.

Upon completion of the concepts of the feedback control loop and its individual components, these concepts will be applied to some of the more common control systems and applications.

One major point to keep in mind is that typically more than one solution may be available to solve a particular control application. The contents of this material will convey UOP's process control philosophy as well as some of the background behind the philosophy.

Session Material

- **Presentation Material**
  - **General Instrumentation**
- **Text Material**
  - **Instrumentation**
  - **Glossary of Terms**

UOP A Honeywell Company

EDS 2006/Inst-3

As a handout each student will receive a copy of the power point presentation. User notes have been provided for the majority of the slides. Additional notes should be taken by the student when specific information is presented as examples not covered in the original presentation.

Each student also receives text material and a glossary of the more common process control jargon. The text material provides another level of detail above and beyond the power point presentation. If a specific topic sparks your interest, this text material should be reviewed for additional details.

Session Overview

- Terminology
- P&ID Representation
- Feedback Control Loop
- Individual Components of Feedback Loop
- DCS System Requirements
- Process Control Applications

UOP
A Honeywell Company

EDS 2006/Inst-4

We will review UOP's standard nomenclature for instrument representation on the P&IDs. UOP follows for the most part ISA S-5 symbology, but this nomenclature can easily be modified include additional items not covered by ISA.

We will review the FEEDBACK control loop and identify the various components of the feedback control loop. We will also spend some time looking at the details of each of the components along with a review of the most commonly-used components.

In today's modern world the distributed control system has proven to be a monumental advancement in process control; but due to the rapid improvements in system components we will limit our discussion to a few basic system requirements. UOP process units do not require DCS systems to meet guarantees and UOP does not mandate the use of them. However taking advantage of their use, can only improve the overall operation and optimize performance.

Lastly we will apply the knowledge learned about the feedback control loop and apply it to various process control applications. Starting with the simple feedback control loop, we will investigate some common process load disturbances and build on various improvements in the basic control loop to counterbalance these disturbances.

Terminology and P&ID Representation

- **Comprehensive Dictionary of Measurement and Control – Second Edition, Instrument Society of America**
  - **Glossary of Terms**
  - **P&ID Instrument Signals**

uop A Honeywell Company EDS 2006/Inst-5

The Glossary of Terms contains definitions of some of the more common control terms as applied to the Process Industry. For a more comprehensive listing of process control terminology and definitions, refer to:

<u>Comprehensive Dictionary of Measurement and Control Second Edition 1991</u> Instrumentation, Systems, and Automation Society (formerly called: Instrument Society of America) 67 Alexander Drive PO Box 12277 Research Park North Carolina 27709 USA www.isa.org

The instrument symbols provided are UOP's typical P&ID representation and are common industry practice. However these symbols may vary from client to client; and a basic understanding of these symbols is required in order to interpret the meaning of any instrumentation/control loop represented on the P&ID's.

Terminology

- **What is an Instrument?**
  - **A *device* used directly or indirectly *to measure* and/or *control* a *variable***
- **What is Process Control?**
  - **The *regulation or manipulation* of variables influencing the *conduct of a process* in such a way as to obtain a product of *desired quality and quantity* in an *efficient manner***

UOP A Honeywell Company

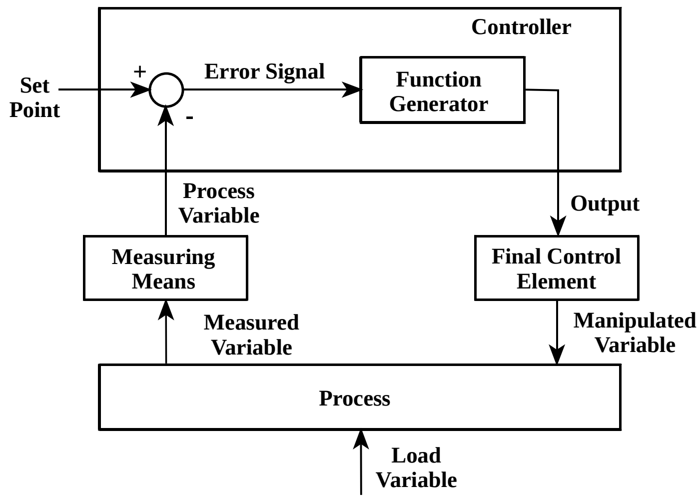
EDS 2006/Inst-6

The Process Industry is governed by economics and economics is a major player in determining whether or not it is an “efficient manner”.

As an example, separation of the isomer para-xylene (P-xylene) from its counterparts (meta, ortho, and ethyl-benzene) by way of distillation is economically not justified. The boiling points of the individual isomers are: O-xylene (292 °F), M-xylene (282 °F), P-xylene (281 °F), and ethyl-benzene (277 °F).

Two competing technologies, other than distillation, exist today for the efficient separation of P-xylene. These are selective adsorption (UOP’s Sorbex Process), and crystallization technology (freezing point of P-xylene is 56 °F, next closest isomer is O-xylene with a freezing point of -13 °F).

Terminology
(Feedback Control Loop)



A Honeywell Company

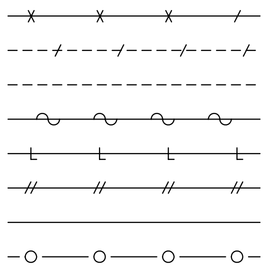
EDS 2006/Inst-7

This block diagram outlines the individual components that make-up the feedback control loop. The glossary of terms should be reviewed for definitions of the various components.

Examples of the process are flow, pressure, level, temperature, etc. In the process industry, the primary work horse for automatic control is the three mode PID feedback controller.

P is proportional action, most commonly known today as gain. I is integral action and D is derivative action. A more detailed discussion of each type of action, when to use or not use various combinations, etc will follow when we investigate the controller component in detail.

P&ID Instrument Symbols
(Instrument Line Symbols)



CAPILLARY TUBING (FILLED SYSTEM)
ELECTRIC BINARY SIGNAL (ON-OFF)
ELECTRIC SIGNAL
ELECTROMAGNETIC OR SONIC SIGNAL
HYDRAULIC SIGNAL
PNEUMATIC SIGNAL
PROCESS CONNECTION OR MECHANICAL LINK
SOFTWARE LINK

IST-LGND-001A

A Honeywell Company

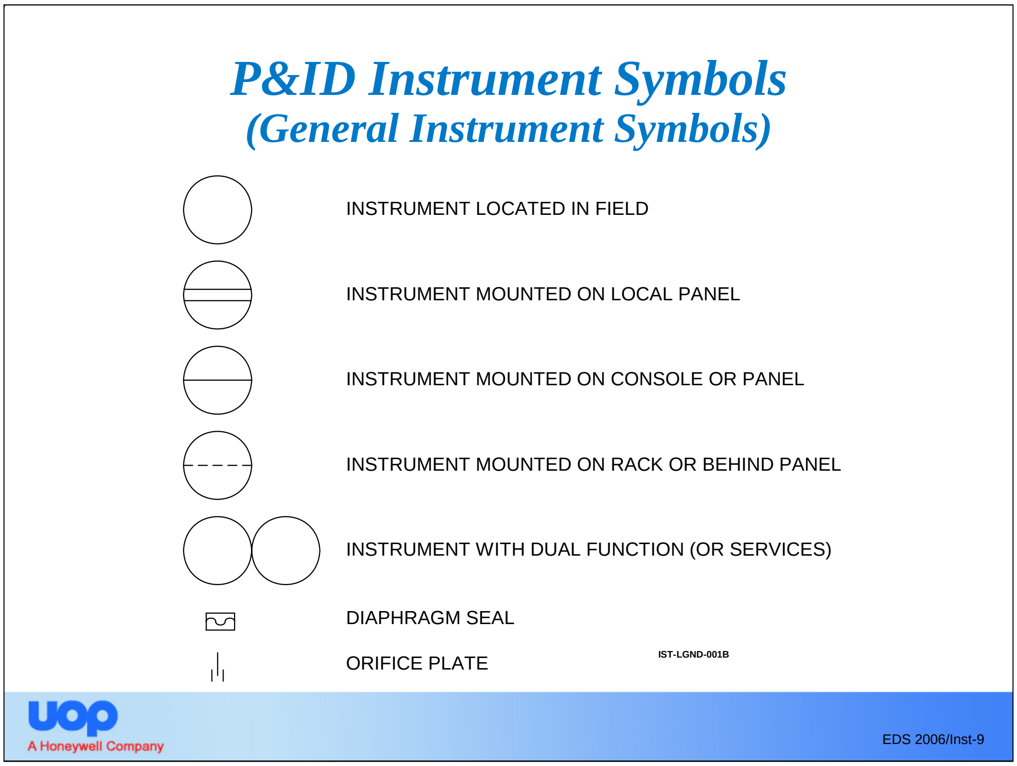
EDS 2006/Inst-8

In today's environment the more common symbols used are the electrical binary signal, electrical signal, pneumatic signal, and software link.

The electrical binary signal represents a discrete input or output signal and generally is a 24 Vdc, 48 Vdc, 110 Vac, or 220 Vac signal.

The most common input/output electrical signal is the 4-20 ma signal. However with the advent of digital signals, a variety of proprietary signals exist today. A brief discussion follows this section.

The pneumatic signal is typically reserved for the final control element and in most cases is the control signal to the control valve. The industry standard is 3 - 15 psig (0.2 - 1.0 kg/cm2(g)).

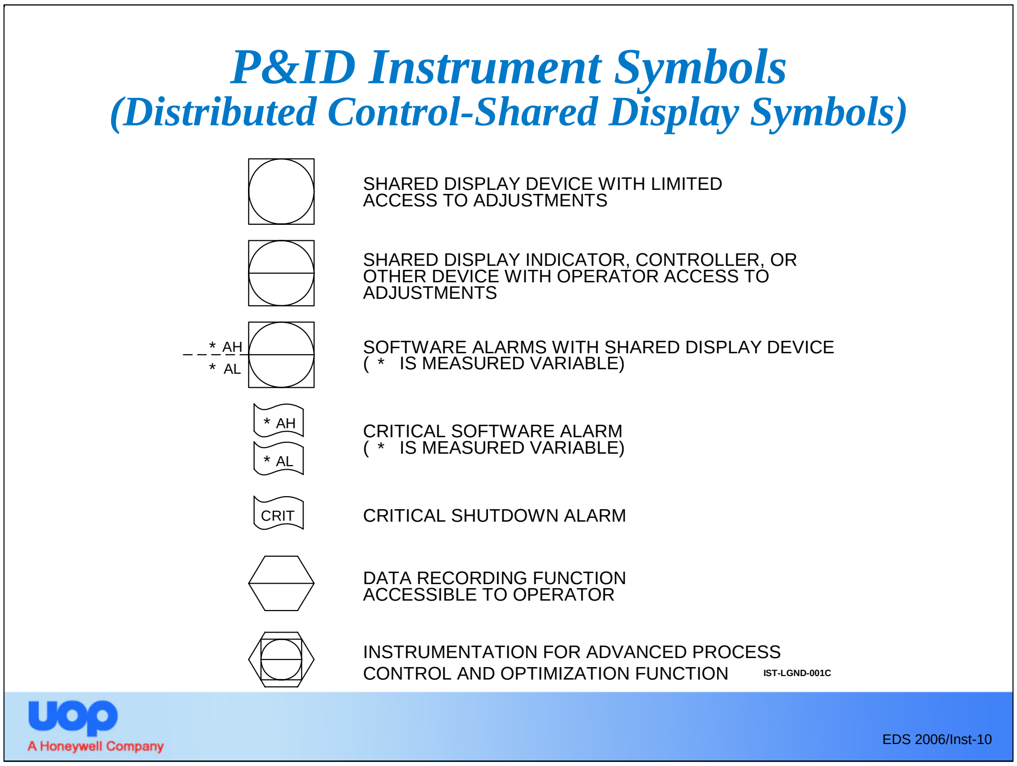


The circle represents a tangible object, i. e. a physical piece of hardware. The various lines (double line, single line, dashed line, etc) are used to distinguish the location of the device.

All in-line devices (primary flow elements, control valves, etc.) are examples of instruments located in the field. Other examples of instruments located in the field are temperature and pressure gauges.

Packaged units, such as compressors, instrument air dryers, centrifuges, etc., are often supplied with local control panels in which many of the instruments may be installed in the local panel supplied with the packaged unit.

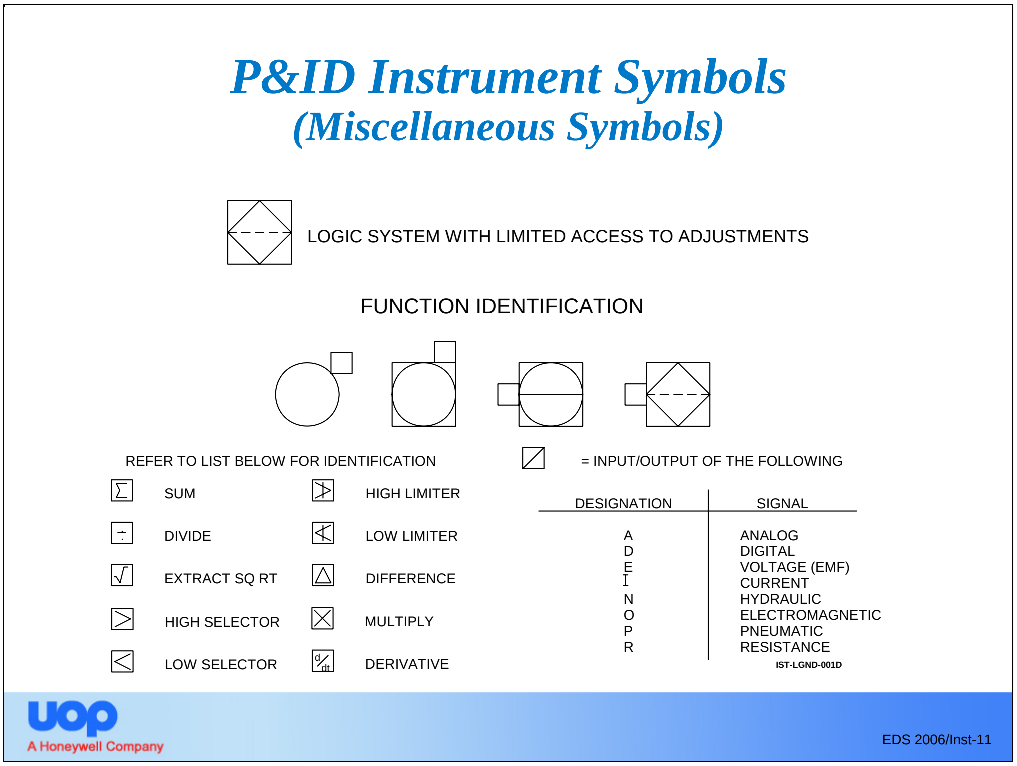
Board-mounted instruments are limited today because of the distributed control systems (DCS). Examples are hard-wired emergency shutdown switches and hard-wired annunciator alarms.



The majority of new installations today are equipped with DCS's. The flexibility of the DCS allows for easy configuration of many control functions including controllers, indicators, alarming functions, math functions, etc.

Essentially all main panel instruments have been replaced with a DCS configurable counterpart. One major advantage of the DCS is that the configuration can be easily modified compared to recalibration of its hardware counterpart.

As an example a current switch, which was used frequently for detecting say low process pressure, would have to be recalibrated by a qualified technician if the alarm point was changed. In the DCS changing the alarm point takes only seconds to implement.



A variety of mathematical functions exist in today's DCS's. Among these functions are summers, multipliers, signal selectors, signal limiters, etc.

Field-mounted signal converters, such as current to pneumatic (I/P) transducers, convert the 4 -20 ma electrical signal to a 3 -15 psig pneumatic signal compatible with the operation of the final control element.

Safety, interlock, and sequential logic systems have a variety of analog and discrete I/O modules. Serial communications are often provided between the logic systems and the DCS's used for basic process control.

P&ID Instrument Symbols
(Functional Identification of Instruments)

	FIRST LETTER	SUBSEQUENT LETTERS		
	MEASURED VARIABLE	MODIFIER	READOUT	OUTPUT
A	ANALYSIS		ALARM	
B	BURNER FLAME			
C		COMPENSATED		CONTROL
D		DIFFERENTIAL		
E			PRIMARY ELEMENT	
F	FLOW	RATIO (FRACTION)		
G			GLASS, GAUGE	
H	HAND (MANUAL)			
I	CURRENT		INDICATE	
J	POWER	SCAN		
K		TIME RATE OF CHANGE	CONTROL STATION	
L	LEVEL		LIGHT	
M				

IST-LGND-001E

A Honeywell Company

EDS 2006/Inst-12

The following tables outline the functional identification of the basic instrumentation loops. UOP will include similar tables on the legend P&ID.

The loop identification and tag number is contingent upon the number of letters, the letter sequence and the quantity of loops. The table is divided into two primary sections: 1) first letter and 2) subsequent letters. The typical flow loop consists of a Flow Element, Flow Transmitter, Flow Indicating Controller, and Flow Valve and would be shown on the P&ID with the following tag numbers, respectively: FE-001, FT-001, FIC-001, and FV-001. The numeric, per UOP general practice, is to number all loops sequentially (001,002, etc), while the alpha-characters are designated by type (flow, pressure, temperature, etc).

P&ID Instrument Symbols
(Functional Identification of Instruments cont'd)

	FIRST LETTER		SUBSEQUENT LETTERS	
	MEASURED VARIABLE	MODIFIER	READOUT	OUTPUT
N				
O			ORIFICE	
P	PRESSURE, VACUUM		POINT (TEST CONN)	
Q	QUANTITY	INTEGRATE, TOTALIZE		
R			RECORD	
S	SPEED, FREQUENCY	SAFETY		SWITCH
T	TEMPERATURE			TRANSMITTER
U	MULTIVARIABLE			
V	VIBRATION			VALVE
W	WEIGHT		WELL	
X		SKIN		
Y				RELAY, COMPUTE
Z	POSITION			

IST-LGND-001F

A Honeywell Company

EDS 2006/Inst-13

The tables can be modified to include additional items as needed. Therefore each set of P&ID may have unique legends.

Feedback Control Loop

- **Signal Transmission And Transmitters**
  - **Analog – Pneumatic/Electronic**
  - **Electronic Analog/Digital**
  - **Transmitter Performance**
  - **Fieldbus**
- **Loop Components**
  - **Process**
  - **Measuring Means**
    - **Temperature, Flow, Pressure, Level, Analysis**

uop
A Honeywell Company

EDS 2006/Inst-14

Over the past 1/2 century, technology has advanced from pneumatic to microprocessor-based digital transmitters. Digital transmitters offer various advantages over its analog counterpart, but the speed of response has become an issue with respect to control performance and transmitter performance can vary from vendor to vendor.

Analog – Pneumatic vs. Electronic

- **Pneumatic**
  - **Dry air used as transmission medium**
  - **Standard range 3-15 psig (0.2-1.0 kg/cm^2(g)) (corresponding to 0-100% of signal)**
  - **Transmission response is typically slow**
- **Electronic**
  - **Low power level system (0-24 volt DC)**
  - **Standard range 4-20 milli-amperes (mA)**
  - **Transmission response is instantaneous**

uop A Honeywell Company

EDS 2006/Inst-15

Fast responding pneumatic loops were generally limited to local field-mounted controllers or at best short distances to the control room.

Electronic transmitters proved to be exceptionally better than the pneumatic transmitters. Transmission response was greatly enhanced with the electronic analog instruments, the overall loop performance was improved, and transmission distances were not the limiting factor for the typical process installations.

Electronic Analog vs. Smart Digital

- **Electronic Analog**
  - **A signal representing a variable that is continuously being measured/transmitted**
  - **Output is a continuous 4-20 ma signal**
- **Digital**
  - **A signal representing a variable that is sampled**
  - **Sampled values are a set of discrete values**
  - **Output is a continuous 4-20 ma signal** held **at the last sampled value**

A Honeywell Company

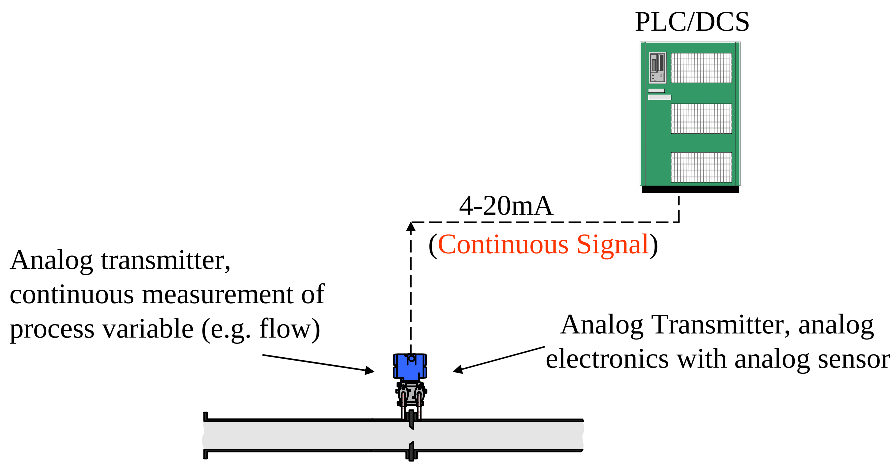
EDS 2006/Inst-16

An analog signal parallels the process variable being measured. The process is continuously being measured and the transmitter signal is analogous to the process variable.

A digital signal is a discrete sampling of the process variable updated periodically.

Analog Signal

- **A signal representing a variable that is continuously being measured and/or being transmitted.**



uop
A Honeywell Company

EDS 2006/Inst-17

To understand the difference between analog and digital instruments, an understanding of the difference between an analog signal and a digital signal is required.

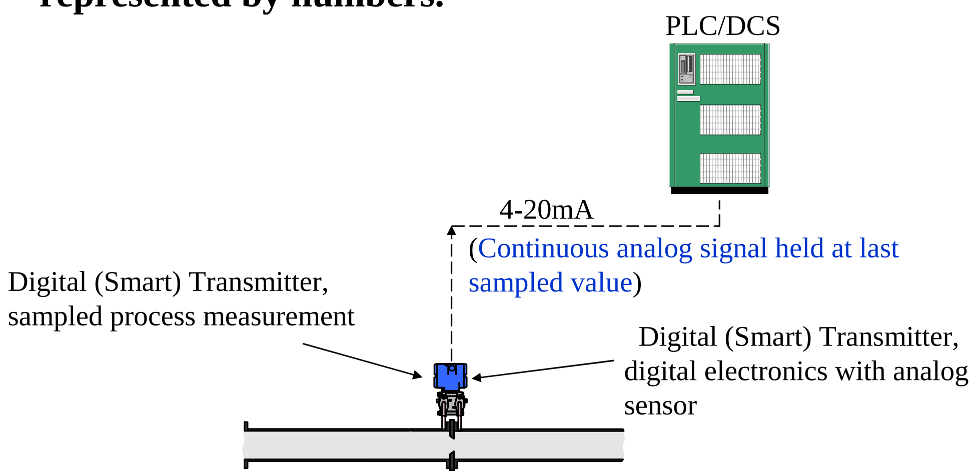
The definition of an analog signal is as follows: “A signal representing a variable that is continuously measured and/or being transmitted”.

Are digital smart transmitters better than analog transmitters for response and accuracy?

Many digital transmitters on the market, although more accurate than analog transmitters, respond poorly compared to analog transmitters..

Digital Signal

- **A signal representing a variable that is sampled. These sampled values are a set of discrete values that are represented by numbers.**



A Honeywell Company

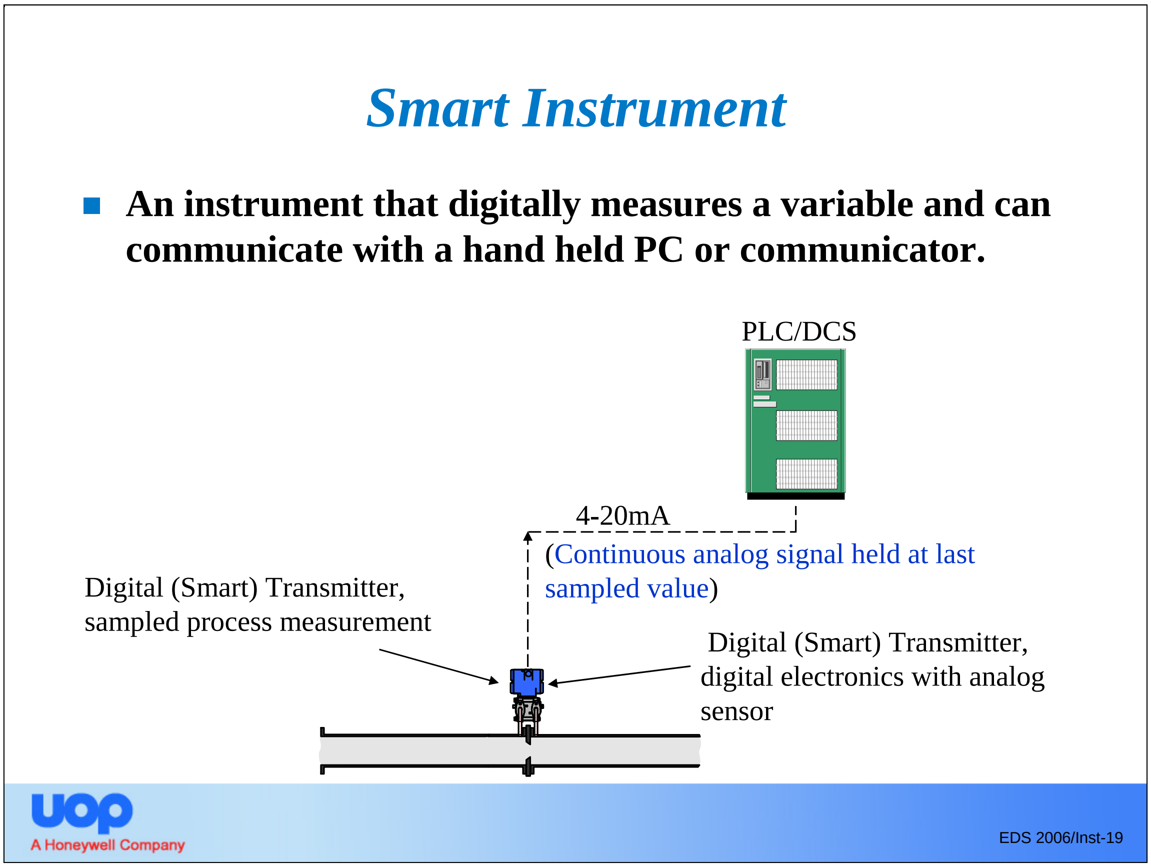
EDS 2006/Inst-18

The definition of a digital signal is as follows: “A signal representing a variable that is sampled. These sampled values are a set of discrete values that are represented by numbers.”.

Even though the 4 - 20 ma signal is an analog signal , the update of the 4 - 20 ma signal is dependent on the digital sampling rate of the measured variable in the digital transmitter.

This difference between an analog signal and a digital signal defines the differences in design between analog and digital instruments.

Many digital transmitters on the market today update the measured variable very slowly and thus respond worse than analog transmitters. The slow updating of the process variable adds dead-time to the transmitter performance. This dead-time degrades the overall transmitter performance.

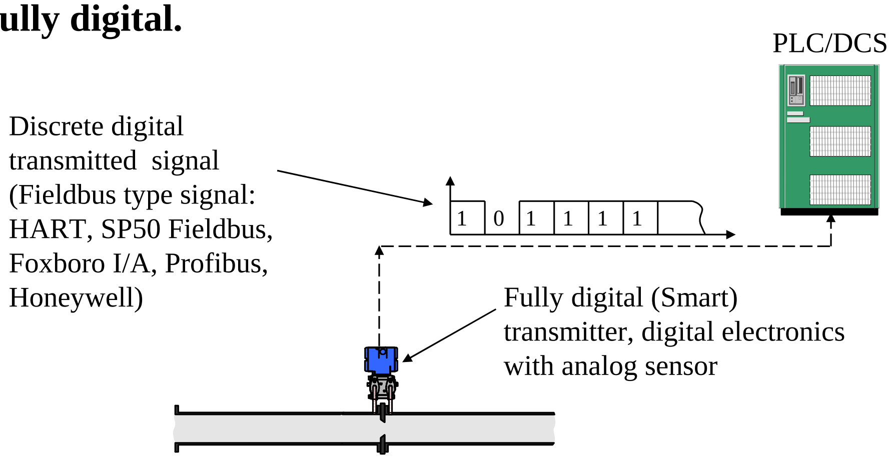


One of the advantages of the digital transmitter is that the design incorporates a microprocessor, i. e. digital electronics.

Therefore a smart digital transmitter by definition is: "An instrument that has a microprocessor, that can communicate with a hand-held communicator or PC, and whose output is a 4 -20 ma signal".

Fully Digital Smart Instrument

- **An instrument that digitally measures a variable, can communicate with a hand held communicator or PC and who's transmitted output of the process variable is fully digital.**



A Honeywell Company

EDS 2006/Inst-20

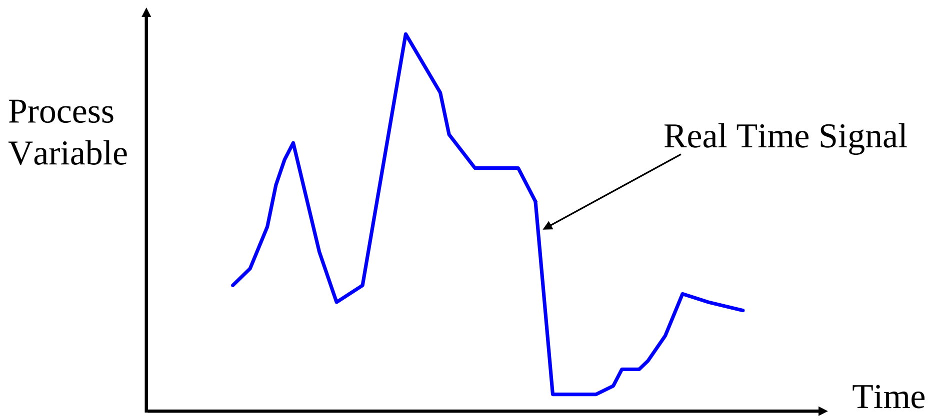
A fully digital smart instrument is one that is fully digital including it's output. Digital instruments with digital outputs for communication are of the fieldbus type.

There are some digital instruments on the market that are both <u>smart</u> and <u>fully digital smart</u> because the output can be selected to be 4 -20 ma or fully digital.

As with the digital (smart) transmitters updating of the process variable adds dead-time to the transmitter performance, the digital communication of the fully digital smart transmitters adds additional dead-time to the transmitter and performance. Therefore the digital communication of the measured variable needs to be as fast as possible to limit the effects of the digital dead-time.

Real Time Versus Deterministic Time

- **Real time in process control:**
  - **Continuous measurement or transmission of a process variable or control signal.**



UOP
A Honeywell Company

EDS 2006/Inst-21

Real time control or measurement refers to any instrument, control device, or system that measures or controls continuously.

Virtually all digital measurements, control devices, or systems do not operate in real time because the sampling rate (deterministic interval) is too large.

Real Time Versus Deterministic Time

- **Digital Deterministic Time (D_t)**
  - **Number of times per second (rate) that process variable is measured or signal is updated**
- **Rate value is dependent on the vendor**
  - **Considered as real time because the rate value (from 1 to 20 times per sec) is a constant value?**
- **D_t is real time when a digital system measures /outputs a signal an infinite number of times per second**

UOP
A Honeywell Company

EDS 2006/Inst-22

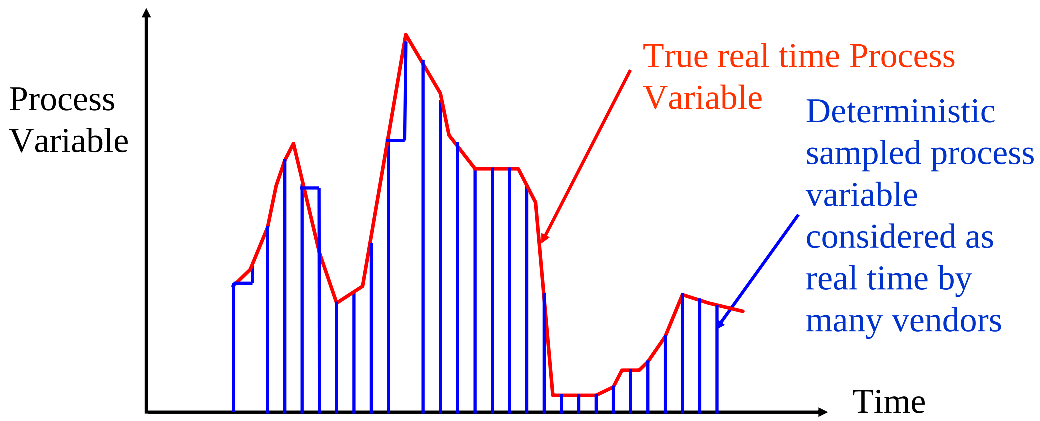
Any control system or instrument, equipped with a microprocessor, has some time delay because of the digital sampling rate. Combined with the fact that microprocessor also cycles through its own diagnostic program, the transmitter is not capable of transmitting the process variable continuously as does the analog counterpart.

The digital deterministic interval should be such that the instrument or control system is able to measure, display, and control the fastest process variations.

Misused Real Time Definition

- **Would this be considered as real time?**

In this example rate is five times per second



A Honeywell Company

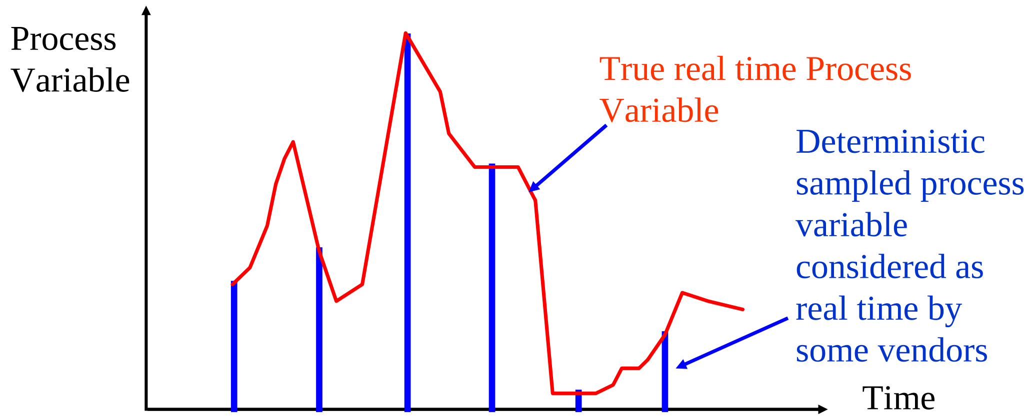
EDS 2006/Inst-23

Many vendors consider this example as real time and would consider the sampling rate of five times per second as too fast and not required. For fast loops such as flow, five times per second is a bare minimum rate.

Misused Real Time Definition

- **Would this be considered real time?**

In this example rate is once per second



UOP
A Honeywell Company

EDS 2006/Inst-24

Many vendors would consider a sampling rate of once per second as real time.

A sampling rate of once per second is often considered as a "standard rate". Vendors would further argue that a faster rate is not required for any loop fast or slow. For many control loops, once a second is a long time to <u>not be controlling</u>. This is incorrect and misleading.

For control loops in fast responding processes, such as flow loops or liquid pressure loops, degradation of loop performance is often attributed to the slow sampling rate of the associated control system or instruments.

Correct Sampling Rate

- **The sampling rate should be fast enough to see the fastest changes in the process**
  - **Fast control loops, such as flow, pressure or compressor anti-surge, require a sampling rate of at least 10 times a second**
- **High rates alone will not improve loop performance on a fast control loop**
- **Controllers with fast execution rates & fast responding control valves are also required**

UOP A Honeywell Company

EDS 2006/Inst-25

Any field device should be able to measure and transmit the fastest changes in the process.

On fast loops the following combination among the devices should exist:

a very responsive field device sensor;

the field device is sampling the sensor at a high enough rate (10 to 20 times per second) to “see” the process changes;

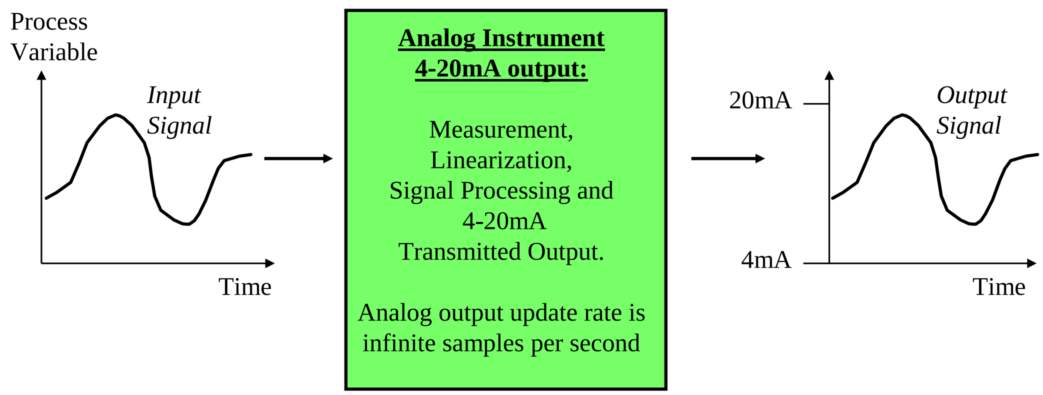
the communication medium is as fast at the field device sampling rate;

the loop controller has a high execution rate (e.g. 5 to 10 times per second, for anti-surge control execution rate can be up to 25 times per second);

the control valve is very responsive (e.g. control valve performance should not inhibit overall loop performance).

Analog Transmitter with 4-20mA Output

- **Analog instruments operate in real time when little or no filtering or signal processing is used**
  - As a consequence they are very responsive

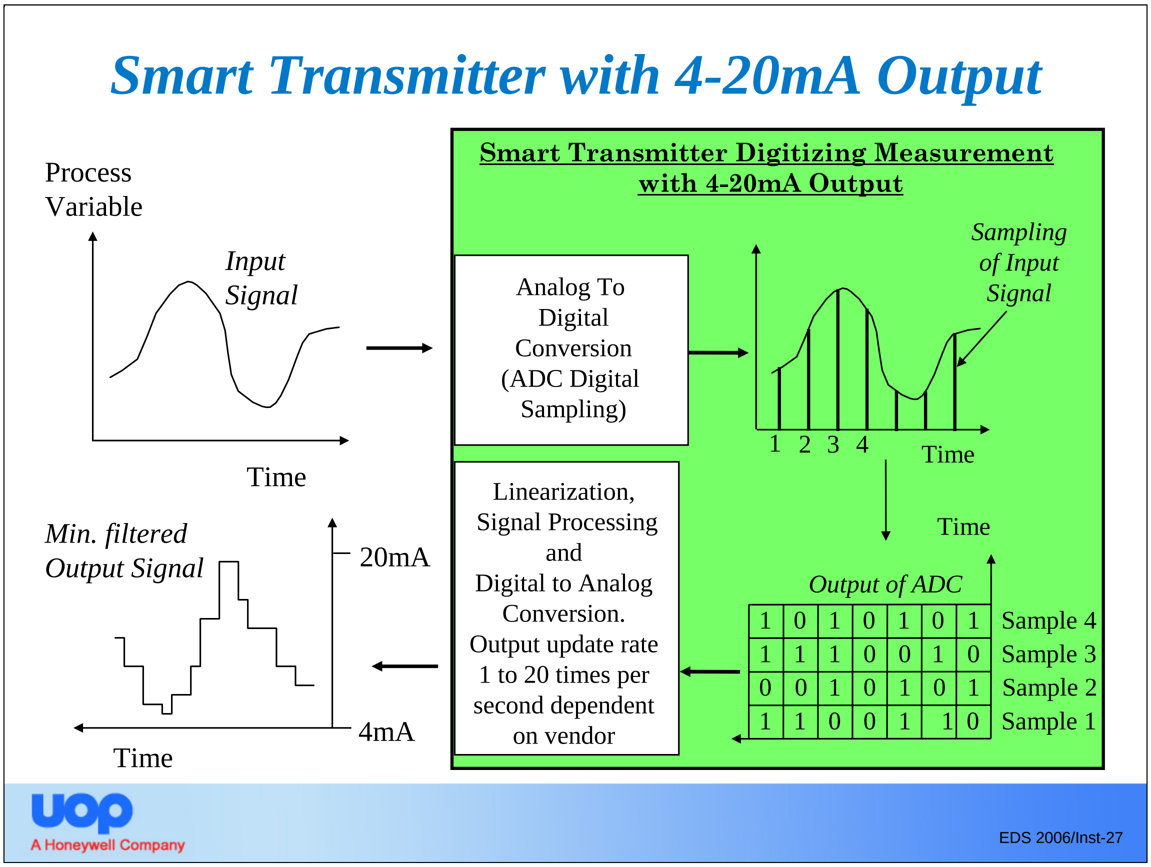


A Honeywell Company

EDS 2006/Inst-26

The main benefit of analog instruments is that they work in real time, unlike digital instruments that sample the process measurement.

The sampling rate for analog instruments is infinite samples per second. Some applications, such as blower anti-surge control, are very fast control loops and the analog instrument is well-suited for these types of applications. Digital instruments with a slow sampling rate may not detect that the equipment has moved in and out of surge because of the quickness of the surge phenomenon.



Digital sampling essentially prohibits a digital instrument from measuring and transmitting the process variable continuously. Sampling rates should be as fast (or faster) than the process changes.

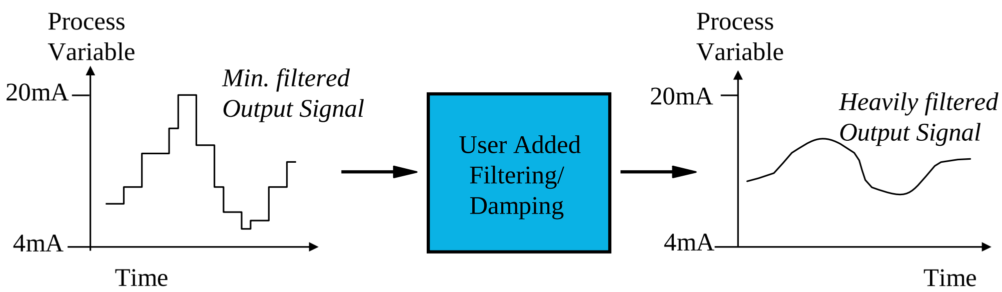
As indicated earlier the digital transmitter is equipped with an analog sensor and digital electronics (microprocessor). The heart of the microprocessor (and thus the digital sampling) is the analog to digital converter.

The digital sampling process produces dead-time in the process measurement. In order to approach real time measurement, dead-time has to be minimized. Instruments with slower sampling rates create additional dead-time. Instruments with slow sample rates (i. e., 1 to 3 samples per second) should not be used except on processes that vary slowly (process variations in seconds not milliseconds).

Due to the action of the digital to analog converter the 4-20mA signal has steps. A further filtering stage is normally required before the 4-20mA signal is transmitted.

Digital Filtering (Damping)

- **Minimum filtering value typically 0.2 sec. Removes noise and other unwanted high frequency signals**
  - **Increased filtering (typically up to 32 sec) removes lower frequency signals (often wanted process variability signals)**
  - **Therefore increased filtering should be used with caution**



EDS 2006/Inst-28

Filtering (also called damping, as higher frequencies are filtered out) is used to "clean-up" the signal before the 4-20mA signal is transmitted.

The minimum filtering value for a differential pressure transmitter is typically 0.2 seconds. The amount of filtering can be increased (increase damping) by increasing the damping value (e.g from 0.2 to 10 seconds). However too much filtering/damping of the measured variable can be detrimental for control of fast processes by filtering-out process variability. The signal is made to look "clean" at the expense of responsiveness.

Analog and Digital Comparison

	Analog	Digital
Accuracy	1.00%	0.10%
Sampling Rate	Infinite, (continuous)	Discrete (1 - 20 samples/sec)
Measures Real Time	Yes	No
Responsiveness	Faster	Slow to Fast (rate dependent)
Calibration	Shop	DCS or Communicator
Ambient Temp Effect	No Compensation	Temperature Compensated
Linearization	Inferior	Superior
Microprocessor	No	Yes
Built-in Diagnostics	No	Yes
Dead Time	Minimal	Yes (varies)

uop
A Honeywell Company

EDS 2006/Inst-29

Some benefits of analog instruments are as follows:

Analog instruments operate in real time and are therefore more responsive than digital instruments. Analog instruments have the equivalent of measuring infinite samples per second of the process variable being measured. For current digital instruments the range of sampling rate is between 1 to 20 samples per second.

Some of the disadvantages of analog instruments are as follows:

Calibration typically requires a skilled technician; and in the process industry, calibration usually requires that the instrument be dismantled and sent to the instrument shop. Measurement accuracy is affected by ambient conditions and the instrument tends to drift with on-stream time. Periodic recalibration is required to ensure measurement accuracy.

Some of the advantages of digital instruments are as follows:

The inclusion of the microprocessor allows re-ranging the transmitter with a hand-held communicator or remotely from the DCS console. Typically a RTD is “built-in” to allow for compensation of ambient temperature effects improving the overall accuracy of the measured variable. Superior linearization and transmitter characterization are some additional advantages of the digital instruments.

Transmitter Reference Accuracy

- **Combined effects of linearity, hysteresis, and repeatability at reference conditions**
- **Means of evaluating/selecting transmitter**
- **Accuracy ranges from 0.075 to 0.2%**
- **Does not predict real operating performance**
- **Does not take into account temperature effects, line pressure, process variations, or transmitter stability over time**

A Honeywell Company

EDS 2006/Inst-30

Reference accuracy alone does not necessarily predict how the transmitter will perform during normal operation or long term.

Overall Real Performance

- **Operating Performance**
  - **Includes reference accuracy, ambient temperature effects, and static pressure effects**
  - **5-15 times greater than reference accuracy when the appropriate errors are included**
- **Evaluate performance based on**
  - **Operating Performance**
  - **Long-term Stability**
  - **Dynamic Performance**

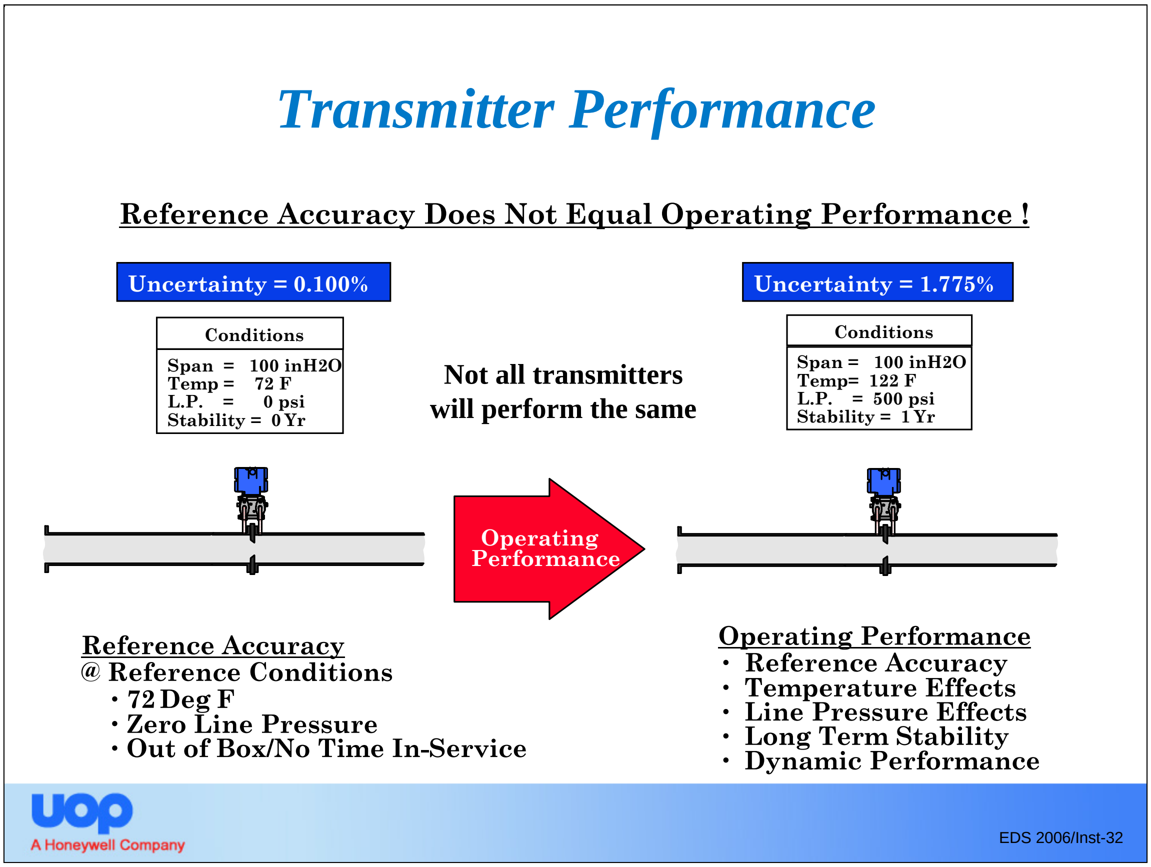
A Honeywell Company

EDS 2006/Inst-31

In addition to reference accuracy, the evaluation and selection of transmitters should take into account operating performance, long term stability, and dynamic performance.

Operating performance includes reference accuracy along with ambient temperature and static pressure effects. The operating performance can be 5 - 15 times greater when taking into account actual operating conditions.

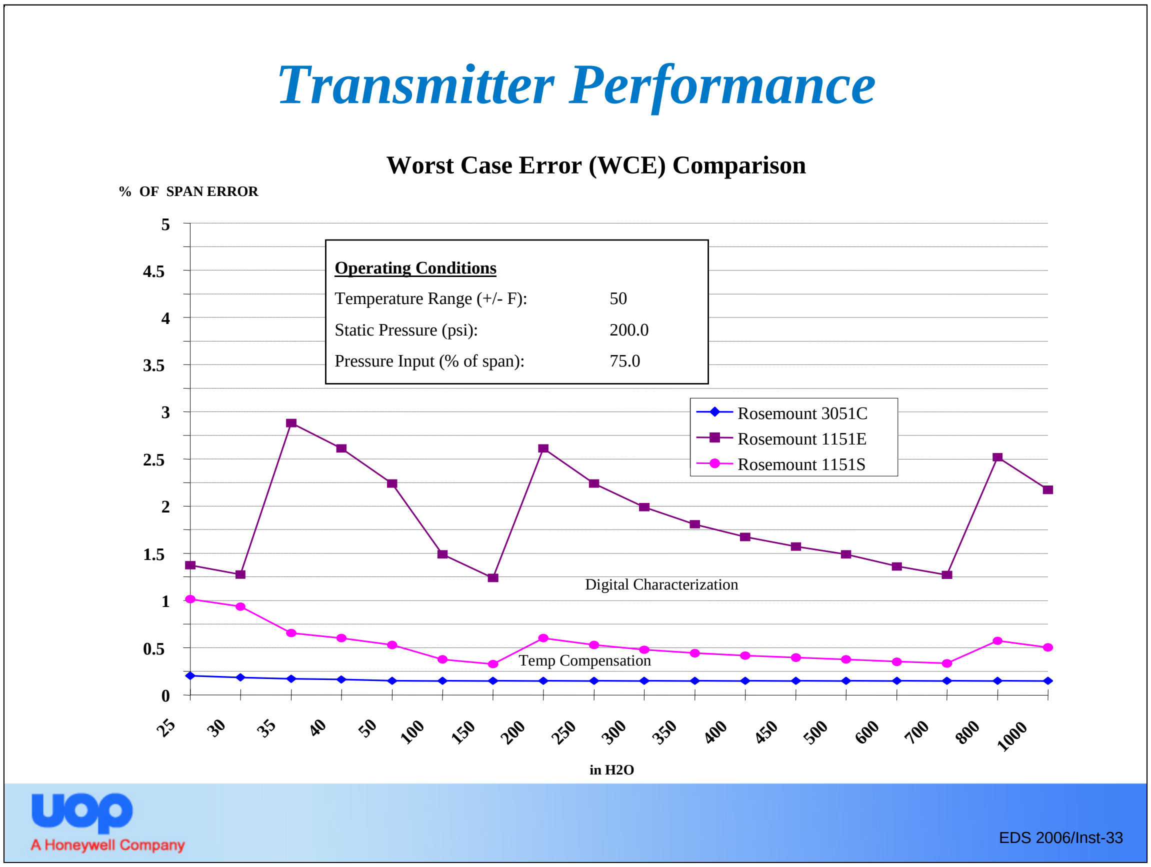
Long term stability and dynamic performance should also be reviewed when selecting the transmitter. Dynamic performance takes into account the update rate and digital communication of the transmitter. For fast acting processes, like flow or liquid pressure control loops, dynamic performance of the transmitter will play an important role in the overall loop performance.



Performance is typically characterized by a comparison of reference accuracy specifications for each transmitter. Published accuracy's are generally 0.075% to 0.100% of span. However reference accuracy alone does not equate to nor predict actual operating performance of the transmitter. In many cases the actual operating performance is 2 - 15 times larger than the transmitter reference accuracy when the appropriate errors are included.

Reference accuracy is the combined effects of linearity, hysteresis, and repeatability without accounting for ambient temperature changes, line pressure, process variations, or transmitter stability over time.

Actual operating performance includes reference accuracy, ambient temperature effects, and static pressure effects. Long term stability and dynamic performance should also be considered in the evaluation of transmitter performance. A basic understanding of actual operating performance is required to adequately compare and select the best transmitter for the application.



Worst Case Error (WCE) as a function of Rosemount transmitter design

For the Rosemount 1151E (electronic analog), which is a purely analog transmitter, the WCE ranges from 1.25 to almost 3%. The peaks represent a change in capsule range.

The 1151S (smart) improved the WCE ranging from 0.3 to1%. This improvement is due mainly to the digital characterization of the pressure sensor.

The latest transmitter, 3051C (fully digital smart), has WCE at less than 0.5%. Improvement for this transmitter design is due to the addition of a temperature sensor in the sensor module compensating for ambient temperature changes as well as an improved pressure sensor (which is also digitally characterized).

Negative Effect of Digital Devices

- **Digital devices introduce Dead Time (T_d)**
  - **Sampling process adds dead time to the loop**
  - **Digital communication adds further dead time**
  - **Dead time worsens the response of a controller**
- **To improve the overall loop response, dead time needs to be reduced – not increased**

EDS 2006/Inst-34

A digital instrument measures more accurately the process variable than an analog instrument, but most of the digital devices do not respond as well as an analog instrument. The slower response is due to dead-time caused by the sampling process; and for instruments with digital communications, this digital communication adds further dead-time to the transmitter response.

For digital instruments the sampling process adds dead-time and the digital communication adds dead-time. By replacing the analog transmitter with a digital transmitter, dead-time has been added to the overall control loop. As a result of adding dead-time, the control loop response will not be improved; in fact, the response will be worse for most digital devices. The degradation of the control loop response depends on the speed of the sampling process and the speed at which the process variable is transmitted to the controller.

If the sampling rate is high (10 - 20 times per second and digital communication is fast(5 - 10 times per second) for the transmitter, then the negative effects of the digital device can be minimized. Also if the loop controller is poorly tuned or the process changes slowly then the negative effects will not influence the overall loop performance.

Analog Versus Digital Performance
(Dead Time)

- **Dead Time (T_d)**
  - **Dead time is seen as the time lag between the process variable being measured <u>actually changing</u> and when the output of the transmitter begins to change**
  - **Sensor signal A/D conversion, sensor processing time, and digital to analog conversion produce the dead time**

UOP A Honeywell Company

EDS 2006/Inst-35

In many applications the replacement of an analog instrument with a digital instrument adds dead time, but the dead time is often masked by other effects like:

Slow scanning PLC's and DCS's (PLC/DCS dead time due to slow scans is larger than instrument dead time).

Poorly performing control valves due to hysteresis and deadband which in itself produces control valve dead-time.

High process dead time (process dead time much larger than instrument dead time)

Large process time constant (i.e your control loop dead time (Td) to process time constant (Tc) ratio is much smaller than 1):

Td (control loop) / Tc(process) << 1

Analog Versus Digital Performance
(Time Constant)

- **Time Constant (T_c)**
  - **Measured by applying a step change to the transmitter sensor and measuring the time it takes for the transmitter output to reach 63.2% of the step change**
  - **The time it takes for the transmitter output to reach 63.2% of the change in the process variable**

A Honeywell Company

EDS 2006/Inst-36

The time constant is representative of a first order lag system. For an induced step change the time constant is defined as the time it takes the output to change 62.3% of its anticipated change.

Analog Versus Digital Performance

	Analog	Digital
Response Time	T_c	$T_d + T_c$
Settling Time	$4(T_c)$	$T_d + 4(T_c)$
Update Rate	Infinite	Varies

UOP A Honeywell Company

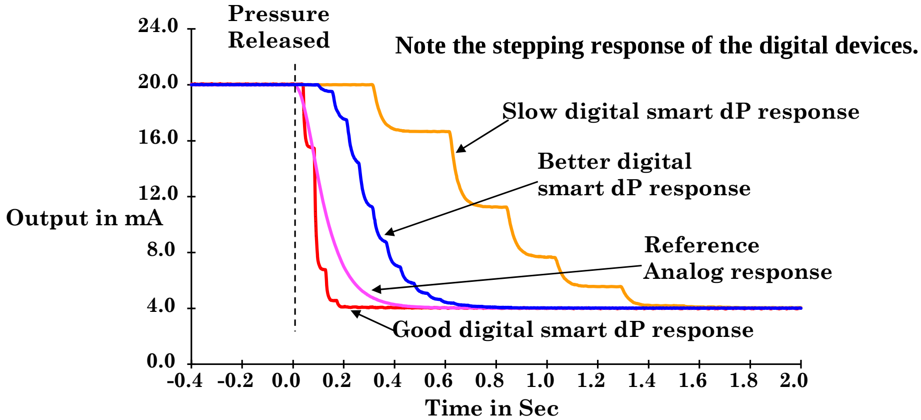
EDS 2006/Inst-37

In an analog transmitter (which has no dead time) the response time is the time constant (Tc).

The update rate is infinite because the 4-20mA output signal in the instrument electronics is always transmitting based on a continuous output signal from the sensor circuit that is always measuring (i.e. no digital sampling). It's important to understand that response time does not represent the time to see the full process change, but rather the time it takes to respond to 63.2% of the initial input step change.

Settling time represents the time it takes to reach 98% of the step change. For digital devices it also includes the instrument dead time.

Analog Versus Digital Performance
(Transmitter Response)



Range of variation in digital transmitter responses, for dP transmitters currently on the market. Step response is based on doing a full range step test (i.e. from 20 to 4 mA)

A Honeywell Company

EDS 2006/Inst-38

The above graph shows actual tests done on dP analog and digital instruments with 4-20mA outputs.

The analog response acts as the reference because in theory a digital instrument replacing an analog one should have at least the same response if not better. In reality for many instruments this is not the case.

Note the stepping response of the digital instruments.

Digital Devices in Closed Loop Control

- **Digital devices provide communications digitally, accurate process measurements, diagnostics, and the capability of measuring more than one variable in a single device**
- **Negative effects of digital device should be negligible provided a high enough sampling rate is provided and the measured process variable is transmitted to the controller in a small amount of time (minimum dead time)**

Because optimizing control loop performance is often neglected, the effects of replacing an analog instrument with a digital one is often missed.

Digital instruments as mentioned before can provide digital communication, diagnostics, very accurate process measurements, the capability of measuring more than one variable; and if a high enough sampling rate is used and if the measured process variable is transmitted to the controller in a small amount of time, then the negative effects of the digital instrument should be negligible in many control loops.

For fast loops with slow responding digital transmitters, control loop performance is often degraded.

Summary of Digital Strengths

- **Microprocessor allows for diagnostics**
- **Easy re-ranging, communication with hand held communicators, PC, and DCS's**
- **More accurate linearization, signal processing and compensation features**
- **Digital devices have the ability to have PID controllers, mathematical functions, logic functions, etc., built-in**

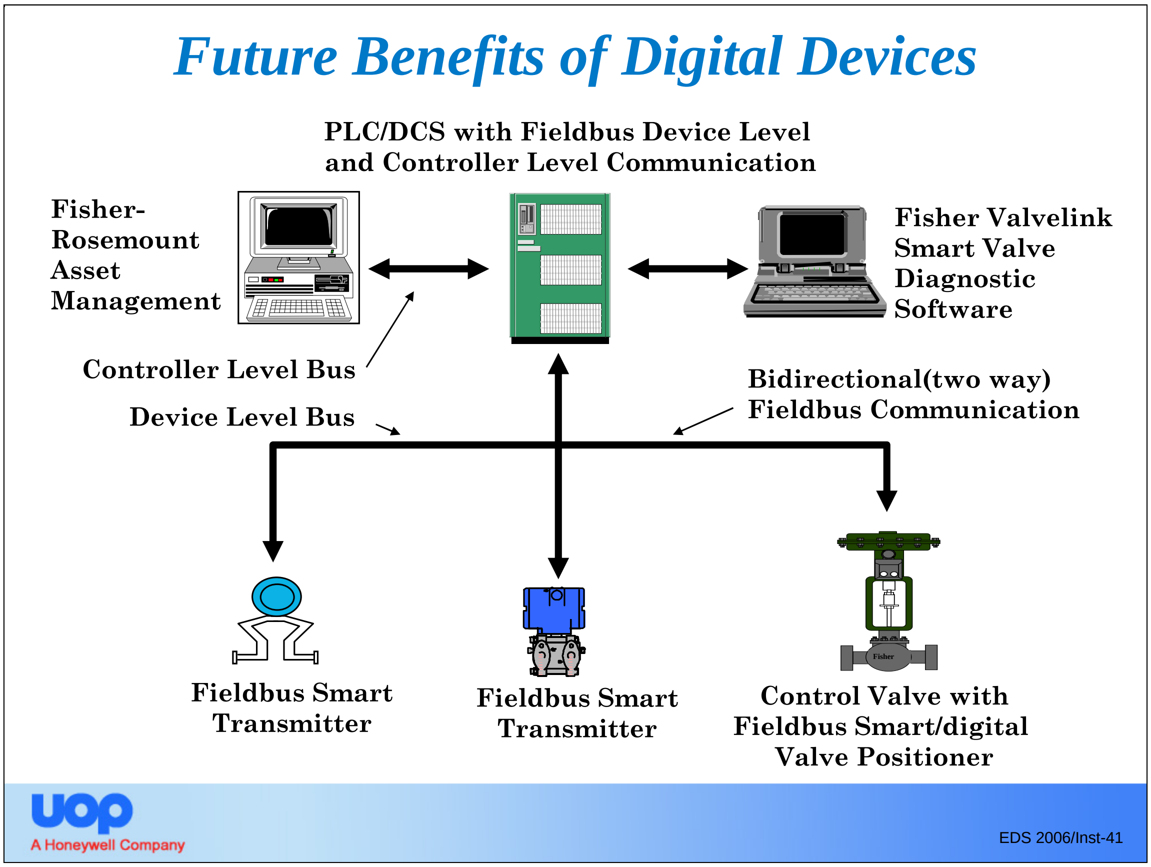
uop
A Honeywell Company

EDS 2006/Inst-40

Having looked at the main weaknesses of a digital instrument when being used in closed loop control, the strengths need to be explored.

By far the main strength of a digital transmitter is that the in-built microprocessor allows for a whole new range of features which are just not practical with analog instruments. Diagnostics, communication, linearization, signal processing and compensation features are among the current advantages of many digital devices.

Some of the other features to be explored are the ability of a digital instrument to have the features often found in PLC and DCS systems such as PID controllers, mathematical functions, and digital functions.



The future is to bring process measurement, process control, instrument diagnostics, instrument calibration, and troubleshooting of field devices to one common platform. This is the basis of the ISA SP50 Foundation Fieldbus standard.

Being on one common platform means that all field and control devices communicate and are configured using a common software language and common hardware components.

Introduction to SP50 Fieldbus

- **SP50 Fieldbus Standard**
  - **Is a digital field device and control bus communication standard as well as a functional control standard for the measurement and control of process variables in hazardous and non hazardous areas to achieve device and control system inter-operability**

A Honeywell Company

EDS 2006/Inst-42

The SP50 Fieldbus standard was initiated by ISA. The next natural step after developing and implementing the 4-20mA standard, was to design a new digital communication and functional standard for field devices measuring or controlling analog variables. This new standard would not only define the digital communications but also define the functions the control and measurement devices could perform. SP50 has been specifically designed for analog measurement and control in both hazardous and non hazardous areas.

SP50 is both a digital device and control bus communication standard as well as a functional control standard for the measurement and control of process variables in hazardous and non hazardous areas to achieve device and control system interoperability.

This standard is now being pursued by the Fieldbus Foundation which is implementing the SP50 standard.

The contender to the SP50 Fieldbus standard is Profibus which has been in existence for a few years. However many of the instrument vendors in the US are supporting the SP50 Fieldbus standard. Only time will tell which will become the “defacto” bus standard for process control.

Fieldbus Objectives

- **Maintain best features of 4-20mA system in developing the SP50 Fieldbus Standard**
  - **Simple two-wire wiring practices**
- **Bus powered devices**
- **Intrinsic safety**
- **Inter-operability of measurement and control devices between vendors**

UOP
A Honeywell Company

EDS 2006/Inst-43

The SP50 ISA committee wanted to keep many of the 4-20mA features for the new SP50 fieldbus standard.

Plant wiring will be basically as per analog loops, i.e. twisted pairs.

Bus systems can supply instrument power.

Bus systems can be IS and use safety barriers.

Bus systems allows for openness in connecting any field device, controller, or other system that uses the SP50 fieldbus standard.

Benefits of SP50 Fieldbus

- **More and better information for predictive maintenance, plant safety, product quality, and regulatory compliance**
- **Multi-drop wiring, for lower installation costs**
- **Support of new intelligent instrument functions and migration of control to field devices for better operating performance**

uop
A Honeywell Company

EDS 2006/Inst-44

The anticipated benefits of the SP50 fieldbus standard over the 4-20mA standard are shown above.

To get the full benefits of the SP50 fieldbus standard more control functions need to be at the field level. The control system should not be structured like a DCS where all control action, trending, alarming, reporting are done in the DCS system.

Fieldbus is not a direct replacement for a DCS, it is a radically different "control system concept" that allows for large amounts of information to pass between all parts of the process (controllers and field devices), analyze and collect data in a process control system, and allow for most of the plant control to be done at the field level.

The field devices will measure variables, set the process alarm points, be able to do diagnostics and set control philosophies. The host systems will collect data, display them on MMI's, operator displays, produce reports, and optimize on overall plant control schemes.

The bus system will reduce wiring costs which is really a small benefit of fieldbus.

Feedback Control Loop

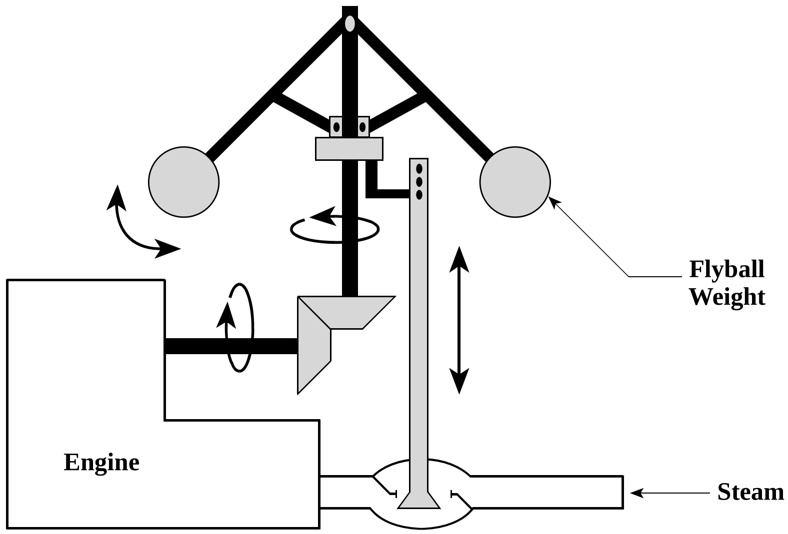
- **Signal Transmission and Transmitters**
  - **Analog – Pneumatic/Electronic**
  - **Electronic Analog/Digital**
  - **Transmitter Performance**
  - **Fieldbus**
- **Loop Components**
  - **Process**
  - **Measuring Means**
    - **Temperature, Flow, Pressure, Level, Analysis**

UOP A Honeywell Company

EDS 2006/Inst-45

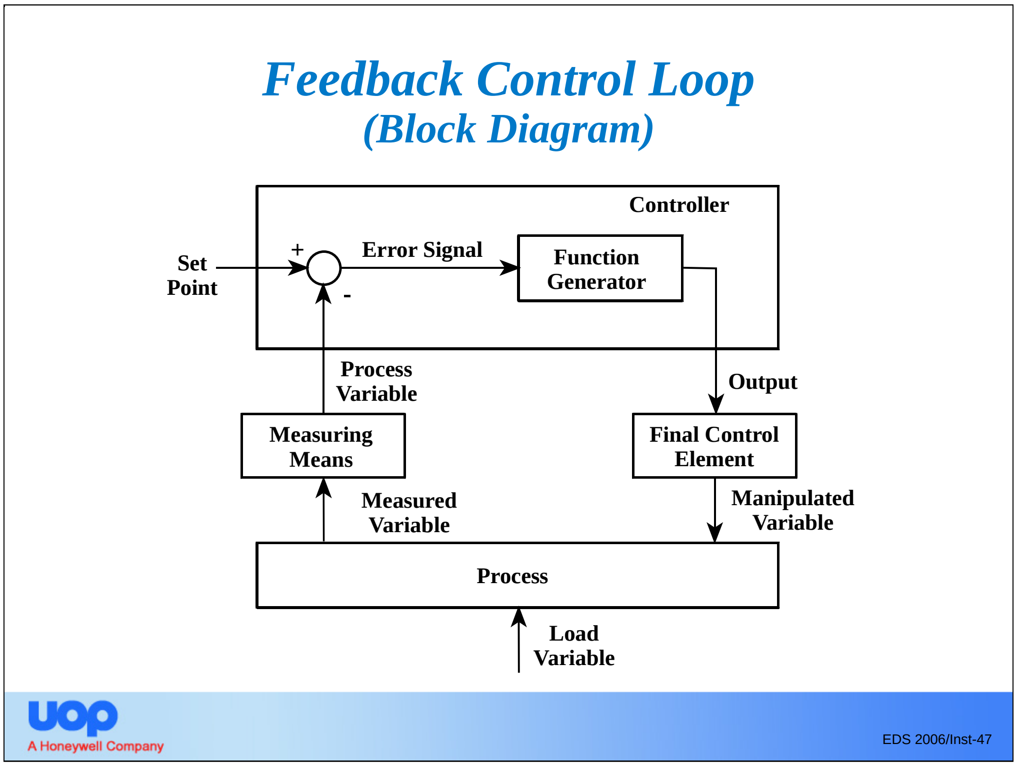
Feedback Control Loop

(Flyball Governor)



The first significant application of closed loop, automatic feedback control was the use of a fly-ball governor on a steam engine in the late 1700's. In this application, the fly-ball governor uses the centrifugal force generated by rotating fly-ball weights, along with a system of linkages and levers, to operate a valve that controls the steam flow to the engine.

On start-up the system is at rest and the steam valve would be in the wide-open position. The operator would introduce steam to the system by manually opening an isolation valve (not shown) on the steam supply. The steam provides work and the engine speed would be sensed by the fly-ball governor through the series of linkages. The fly-ball weights, due to the centrifugal force, would move outward and upward from the center of rotation. This in turn pulls upward on the valve stem closing the steam valve, thus limiting the amount of steam supplied to the engine. The speed set point of the engine could be adjusted manually by adjusting the length of the linkage connected to the valve stem.



Mechanical feedback control was applied to the process industry at the beginning of the 20th century. Today the most widely used automatic feedback control is the basic 3-mode PID controller. Even though other more sophisticated control techniques are available today (especially with distributed control systems), feedback control accounts for more applications than any other.

The block diagram illustrates the various components that make-up the basic feedback control loop. Any feedback loop will consist of a *Process* (flow, pressure, temperature, etc), *Measuring Means* to measure the *Process Variable*, and some form of *Comparator* to compare the value of the *Process Variable* with the desired *Set Point*.

The resulting difference, or *Error Signal*, is sent to a *Function Generator* (PID mathematical algorithm) which operates a *Final Control Element* (typically a control valve). The *Final Control Element* adjusts a *Manipulated Variable* (flow, pressure, speed, etc) as necessary to minimize the *Error Signal*, and bring the *Process Variable* closer to the *Set Point*. How quick this happens is contingent upon the process itself and the P, I, and D constants (tuning parameters) selected for the *Function Generator* (mathematical algorithm).

Feedback Control Loop

(Examples)

Example Number	{1}	{2}	{3}	{4}
Process	Bathtub	Toilet Tank	Shower	Water Heater
Type	Batch	Batch	Continuous	Continuous
Process Variable	Water Level	Water Level	Water Temperature	Water Temperature
Measuring Means	Eye	Float	Touch	Thermostat
Set Point Adjustment	Hand	Linkage	Hand	Thermostat
Comparator	Brain	Linkage	Brain	Thermostat
Function Generator	Brain	Linkage	Brain	Thermostat
Final Control Element	Water Valve	Water Valve	Water Valve	Gas Valve
Manipulated Variable	Water Flow	Water Flow	Water Flow	Gas Flow
Load Variable	(See Note)	(See Note)	Water Pressure	Water Flow

Note: Normally no load variable; however, load variable effect could be created by water leakage or variation in water pressure. A Load Variable is any variable, other than the Set Point, that has an affect on the Process Variable.

UOP A Honeywell Company

EDS 2006/Inst-48

These are just a few examples of feedback control in our daily lives and some feedback loops contain a human component in the loop. Example 1 illustrates a person adjusting a water valve to fill a bathtub. The person filling the tub watches the water level and shuts off the water flow as the level approaches the desired level in the tub. The level information is fed back visually to the operator, who eventually takes action as the level approaches the desired level.

Likewise in example 2 when a float ball in a toilet tank moves a valve to fill the tank to the specified level, feedback control is involved.

Examples 3 and 4 also illustrate feedback control. The third example involves a human component (similar to example 1), where a person taking a shower will adjust the flow of hot/cold water to control the temperature of the water. The last example illustrates the use of a thermostat to operate a gas valve to control the temperature of a hot water heater.

Feedback Control Loop Examples
(Comparisons)

- **Bathtub/Shower examples by definition**
  - **manual and open loop control**
- **Toilet Tank/Water Heater examples**
  - **automatic and closed loop control**
- **First 2 are batch – Last 2 are continuous**
- **All are processes having measuring means, comparator, set point, function generator, and final control element**
- **All examples are subject to load variables**

uop A Honeywell Company EDS 2006/Inst-49

These examples not only exhibit some similarities, they also exhibit some basic differences which helps define various types of feedback control loops. Filling of the bathtub and adjusting shower temperature are by definition manual open loop feedback control. In both instances operator intervention is required (manual control). Likewise open loop implies that no mechanism is available to automatically adjust. In other words, the operator must be present to prevent mis-operation.

Filling of the toilet tank and temperature control on the hot water heater are by definition automatic closed loop feedback control. Some mechanical device is utilized to measure and provide feedback (other than the operator) to manipulate the process automatically. As long as the loop components are operational, the operator need not be present at all times. Reliability then becomes an issue.

Another difference is whether the process is batch or continuous. The bathtub and toilet tank are examples of batch processes, while the other two examples are examples of continuous processes. The refining and petrochemical industries are primarily continuous processes.

Also all cases are subject to disturbances in the Process, (these disturbances are often called Load Variables), and require corrective action by the control system. Changing water supply pressure is an example of a load variable.

Process

- **Even though processes are different, the principles of the feedback control loop are applied to all**
- **The laws of physics and dynamics govern all and each is represented by the same mathematical equations, differing only by the coefficients used in the equations**
- **Many common processes are computer simulated and dynamically modeled**

UOP A Honeywell Company EDS 2006/Inst-50

Although these four processes are different, they all have feedback loops and the same principles can be applied to all of these processes. In fact the process can be almost anything and each will follow the same laws of physics and dynamics. Therefore each process can be described by the same mathematical equations differing only in the coefficients used in the equations. We will investigate some of the simpler first order processes and use some practical approaches to simply the procedures to model the process and determine optimal tuning parameters for controlling the process.

Process
(cont'd)

- **Each process has two external inputs to the feedback control loop**
  - **Set Point - servomechanism response**
  - **Load Variable- regulator response**
- **Load variables play a major role in process control causing transients from steady state operation (dynamics)**
- **Process and Manipulated Variable may differ leading to more difficult control**

uop
A Honeywell Company

EDS 2006/Inst-51

Examining the previous examples of the feedback control loops, continuous processes (examples 3 and 4) have two external inputs to the feedback control loop. Both the Set Point and Load Variables are External Inputs and the control loop must respond to a change in either of these inputs. Any process can have more than one Load Variable. A Load Variable is any variable, other than the Set Point, that has an affect on the Process Variable.

In Example 3 a change in hot water pressure upstream of the valve will cause a larger differential pressure across the valve. The net effect is an increase in temperature as a result of more hot water flowing through the valve. This increase in flow is a direct result of an increase in differential pressure.

For real continuous processes true steady state seldom exists. Load Variable changes and upsets in the process must be countered and corrected and the dynamics of the system must be considered. However the majority of the processes are approximated by the same laws of physics and dynamics, and the instrument engineer is concerned with "tuning" the loop to match the dynamics of the process.

A third point worth noting about the feedback control loop is that the Process Variable and the Manipulated Variable may not be the same. See Example 4. The more indirect relationship between MV and PV is, the more difficult the controller's task may be.

Process Measurement

- **To control must measure Process Variable**
  - **Primary element converts intrinsic property into a measurable signal**
  - **Transmitter converts this signal into standard instrument signal**
- **Transmission of standard instrument signals**
  - **Pneumatic, 3 - 15 psig (0.2 - 1.0 kg/cm^2(g))**
  - **Electronic, 4 - 20 mA (milli-amperes)**
  - **Digital protocol, proprietary/fieldbus**

uop A Honeywell Company

EDS 2006/Inst-52

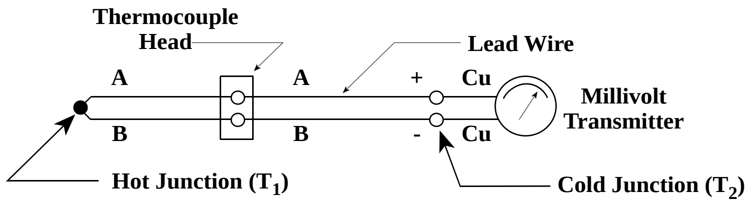
A multitude of devices can used to measure Process Variables and transmit the signal to the control room. In general a primary element is located in the process unit and converts an intrinsic property of the process into a measurable signal. An associated transmitter "reads" this measurable signal and converts the signal from the primary element into a standard instrument signal. Standard instrument signals used in most applications today are the 4 - 20 ma electronic signal and the 3 - 15 psig pneumatic signal.

For an orifice plate, a mathematical relationship exists between flow through the orifice and the differential pressure generated across the orifice plate as a function of flow rate. Once the orifice plate and transmitter are designed and calibrated, variation in flow across the orifice plate is transmitted to the control room where 4 - 20 ma represents 0 to 100% of design. As long as the flow is less than what the system has been designed for, the Process Variable is measurable.

Typically the primary element is located in the field; and depending upon the equipment used, the transmitter may or may not be located in the field. Temperature transmitters are milli-volt transmitters and are frequently located in the control room.

In the Process Industry, several main types of Process Variables exist (flow, pressure, temperature, level, analytical) along with several varieties of primary elements used to measure these variables.

Thermocouple Temperature Devices



ISA Type	A(+)	B(-)
E	Chromel	Constantan
J	Iron	Constantan
K	Chromel	Alumel
T	Copper	Constantan

An EMF is developed when opposite junctions of certain dissimilar pairs of metal are exposed to different temperatures. EMF is proportional to $T_1 - T_2$

uop
A Honeywell Company

EDS 2006/Inst-53

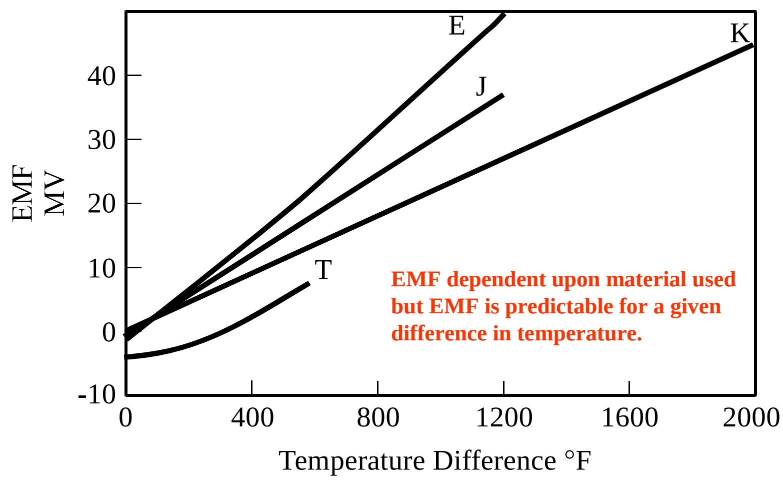
The primary element used most often for temperature measurement is the thermocouple (T/C). T/C operation is based on the Seebeck effect, in which an electromotive force (EMF) develops when opposite junctions of certain dissimilar pairs of metals are exposed to different temperatures. This EMF varies as the temperature difference between the two junctions varies. The magnitude of the EMF depends upon the material used; however, for a given material combination (T/C type), the relationship between EMF and temperature difference is predictable.

The cold junction is typically located in the control room; therefore T/C extension wire is installed from the field termination to the control room termination, and the circuitry must include a mechanism for cold junction temperature compensation. Once the temperature difference is determined as a function of the EMF generated and the cold junction temperature is compensated for, the hot junction temperature is known.

UOP will typically specify type E for most process applications and will specify type K for high temperature applications (fired heaters). Type J was frequently specified in the 60 - 80’s, but because of iron oxidizing, UOP now specifies type E. Type T is often used for cold temperature measurements.

Chromel, Alumel, and Constantan are proprietary alloys, but are widely available. The thermocouple is a relatively simple inexpensive, rugged and reliable device.

Thermocouple EMF



A Honeywell Company

EDS 2006/Inst-54

As stated earlier UOP now specifies type E for most process applications because of the chromel-constantan material is not subject to oxidation as is the type J. Another advantage of type E is the increased sensitivity as shown above.

UOP will typically limit type E to temperatures up to 1200 °F (650 °C). For higher temperatures type K will typically be specified not to exceed 2000 °F (1100 °C).

Although it is difficult to see with this graph, the thermocouple EMF non-linear with respect to the temperature difference. However this tends not to be a set-back for their use in most applications. This can be compensated for by signal characterization in the overall circuitry or by providing a temperature transmitter with a smaller temperature span to improve the measurement reliability.

Other materials of construction for thermocouples are available, but most of UOP's processes are covered by types E, J, and K.

Limits of Error – Thermocouples

			Limits of Error	
Calibration	**Thermocouple Type**	**Temperature Range**	**Standard**	**Special**
J	Iron/Constantan	32 °F to 530 °F	±4 °F	±2 °F
		530 °F to 1400 °F	±0.75%	±0.4%
K	Chromel/Alumel	32 °F to 530 °F	±4 °F	±2 °F
		530 °F to 2300 °F	±0.75%	±0.4%
T	Cppper/Constantan	[-328 °F to 32 °F	±2 °F or ±1.5%	[----]
		32 °F to 260 °F	±4 °F	±2 °F
		260 °F to 700 °F	±0.75%	±0.4%
E	Chromel/Constantan	[-328 °F to 32 °F	±3 °F or ±1.0%	[----]
		32 °F to 600 °F	±3 °F	±2 °F
		600 °F to 1600 °F	±0.5%	±0.4%
R	Platinum 13% Rhodium/	32 °F to 1100 °F	±2.5 °F	±1 °F
	Platinum	1100 °F to 2700 °F	±0.25%	±0.1%
S	Platinum 10% Rhodium/	32 °F to 1100 °F	±2.5 °F	±1 °F
	Platinum	1100 °F to 2700 °F	±0.25%	±0.1%

A Honeywell Company

EDS 2006/Inst-55

One of the disadvantages of the thermocouple is its accuracy of measurement. Thermocouple "Limits of Error" summarizes the accuracy of the measurement. For a type E thermocouple that may be used to measure the Platforming Reactor inlet temperature (say 1000 °F), the limit of error is ±0.5% or in this example ±5 °F. Therefore the actual temperature lies between 995 and 1005 °F. This is not a major concern because the measurement is repeatable; and for the Process, the operator is not concerned about the absolute temperature measurement. His concern will center around the target Octane number of the Platforming product. If the real temperature is 996 °F and the Octane number of the product is 99 (with a Target of 100), the operator will raise the inlet temperature approximately 4 °F per difference in Octane number required. In other words the operator will increase the reactor inlet temperature set point from 1000 °F to 1004 °F. As long as the thermocouple measurement is repeatable the absolute temperature measurement is not a necessity in many applications.

Limits of Error – Thermocouple Wire

			Limits of	Error
ANSI Type	Thermocouple Type	Temperature Range	Standard	Special
J	Iron/Constantan	32 °F to 400 °F	±4 °F	±2 °F
K	Chromel/Alumel	32 °F to 400 °F	±4 °F	±2 °F
T	Cppper/Constantan	32 °F to 260 °F	±2 °F	±1 °F
E	Chromel/Constantan	32 °F to 400 °F	±3 °F	±2 °F
R	Platinum 13% Rhodium/	32 °F to 400 °F	±2.5 °F	±1 °F
	Platinum			
S	Platinum 10% Rhodium/	32 °F to 400 °F	±2.5 °F	±1 °F
	Platinum			

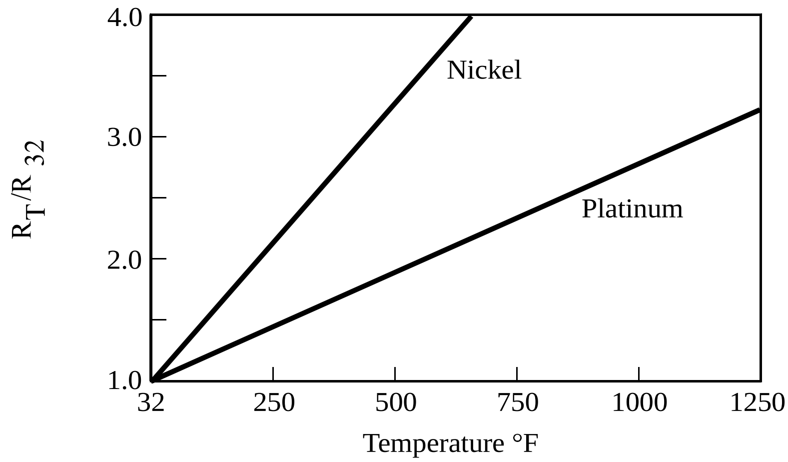
uop
A Honeywell Company

EDS 2006/Inst-56

Limits of Error also exist for thermocouple extension wire. For the example of the Platforming reactor inlet temperature, the limit of error for the type E thermocouple was ±5 °F. The limit of error for type E thermocouple extension wire is ± 3 °F. Therefore the temperature measurement lies between 1000 ± 8 °F if thermocouple extension wire is installed to the control room.

If thermocouple extension wire is not installed (a pair of copy wires can be used), then the cold junction is moved to the field. The transmitter may also be located in the field, but the cold junction compensation circuitry is more complex due to the wider range in ambient temperature. Accessibility also becomes an issue, especially for high temperature applications.

RTD Resistance

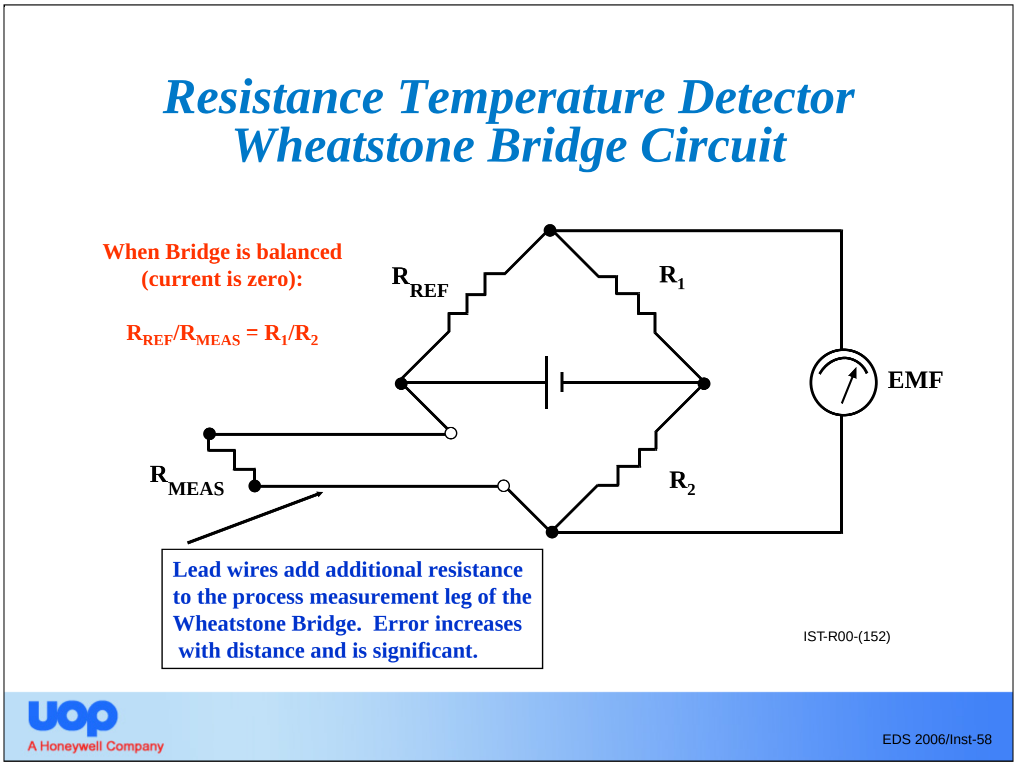


A second type of primary element commonly used for temperature measurement is the Resistance Temperature Detector (RTD). The RTD senses heat based on the principle that a change in temperature results in a proportional change in the electrical resistance of the wire. The most widely used material is Platinum, and is essentially a long thin wire wound into a small coil.

The Platinum RTD has a stable, accurately known linear relationship between resistance and temperature as shown in the graph and is an extremely accurate temperature measurement device. The typical design for a RTD has a known resistance of 100 ohms at 32 °F (0 °C).

Once the resistance of the RTD is measured, the actual temperature is known, unlike the T/C which actually measures a temperature difference along with a cold junction temperature.

Nickel RTD’s are also available, but are not widely used throughout industry.

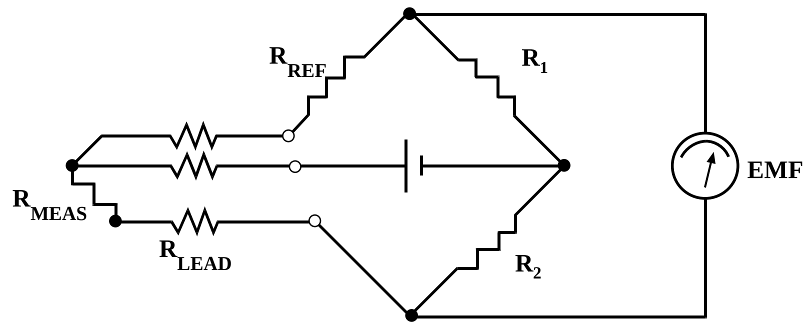


A Wheatstone bridge circuit with a variable reference resistance (R_{ref}) is employed to accurately measure the resistance. As the temperature being measured by the RTD changes, the resistance (R_{meas}) changes. If the bridge is unbalanced, then current will flow across the bridge. The Wheatstone bridge circuitry will adjust R_{ref} in the proper direction to drive the current to zero balancing the bridge.

The principle behind the Wheatstone bridge is such that the ratio of the two known resistances is equal to the ratio of R_{ref} / R_{meas} when the bridge is in balance. Once the bridge is balanced and R_{ref} is determined, R_{meas} can be determined.

However typical installations with lead wires ranging from tens of feet to hundreds of feet introduce a significant variable error in the temperature measurement. Because the circuit is quite sensitive to resistance, the resistance of the copper lead wires must be taken into account. Both the length of the lead wires and the temperature of the lead wires will introduce error in the temperature measurement.

Resistance Temperature Detector Three-Wire Circuit



Adding lead wires for both the bridge and reference legs balances the bridge equally, eliminating the error.

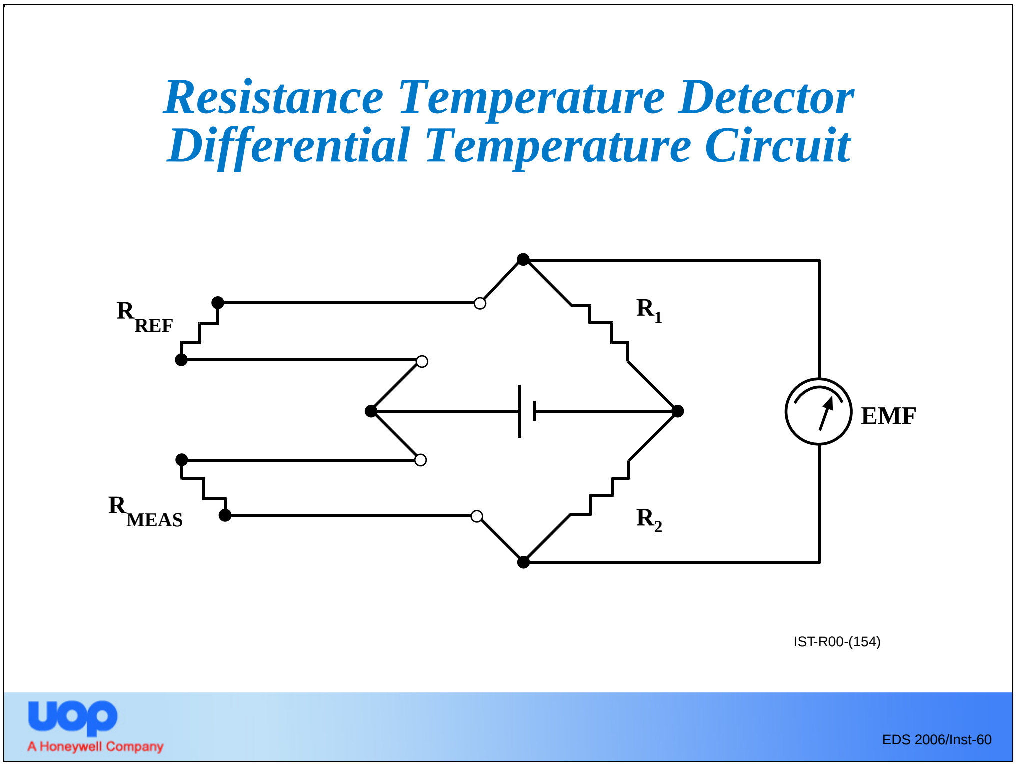
IST-R00-(153)

A Honeywell Company

EDS 2006/Inst-59

Three-wire RTD’s counterbalance the effects on the measurement and reference legs of the Wheatstone bride. In essence the bridge on both the measurement leg and the reference leg is extended to the field. The length, resistance, and temperature of the leads are essentially equal and make the error negligible.

The three-wire copper wire system offsets the cost of the two-wire thermocouple extension wire system.



Temperature devices can also be modified to meet special or specific requirements. In some distillation control applications a differential temperature across a fixed number of column trays is employed to compensate for changes in column pressure. Binary distillation is a primary example. (This control scheme will be reviewed in greater detail later on in the session.)

Utilizing two-two wire RTD's arranged in such a manner as shown, the reference resistance R_{REF} and the measured resistance R_{MEAS} yield the temperature difference across the trays as measured by the Wheatstone Bridge.

Similar set-up can be employed with Thermocouples. Cold junction compensation is not required in this application. The "reference" junction becomes in essence the cold junction and the end result is temperature difference.

Thermocouple and RTD Comparison

	Theromcouple	RTD
Accuracy	Limits of Error wider	Limits of Error narrower
Ruggedness	Excellent, will not affect life of probe	Sensitive to strain, vibration, shock
Temperature Range	[-328 °F to 4200 °F	[-50 °F to 1100 °F
Size	>0.010" sheath diameter	>0.062" sheath diameter
Drift	Periodic calibration required	0.02 to 0.2 °F per year
Resolution	mv/degree, lower signal to noise	ohms/degree, higher signal to noise
Cold Junction	Required	not Required
Lead Wire	Mandatory match	3-lead copper wire (typical)
Response	millisecond response time	> 1 second response time
Sensitivity	Tip sensitive	Thermal mass prevents tip sensitivity
Linearity	Non-linear	Linear
Cost	Lower	Higher

uop
A Honeywell Company

EDS 2006/Inst-61

As a comparison between thermocouples and RTD's, each has its pluses and minuses.

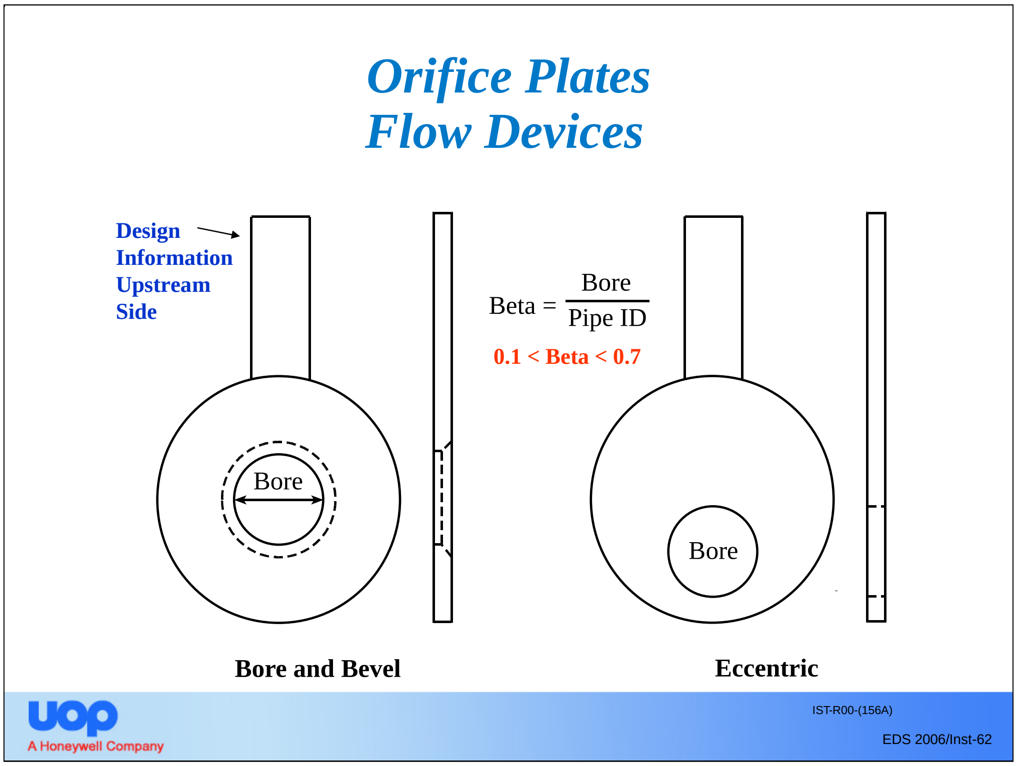
The thermocouple is much more rugged, faster responding to temperature transients, tip sensitive, and covers a much broader temperature range. Cost wise the thermocouple is less expensive, but cost are offsetting somewhat with the overall installation costs associated with the thermocouple extension wire.

On the other hand the thermocouple limits of error are wider, requires periodic maintenance (calibration check), requires cold junction temperature compensation, and has a non-linear signal (characterization of the signal is required).

The RTD is much more accurate in absolute temperature measurement, has a linear signal with respect to temperature, and is a direct measurement of the temperature (cold junction compensation not required).

However the RTD is sensitive to shock, vibration, strain, etc, is limited to temperature less than 1100 °F, has a slower response to temperature transients, and requires a three-wire (minimum) copper system offsetting its overall installation cost advantage.

UOP typically specs Thermocouples and will spec RTD's only where an accurate temperature measurement is required.

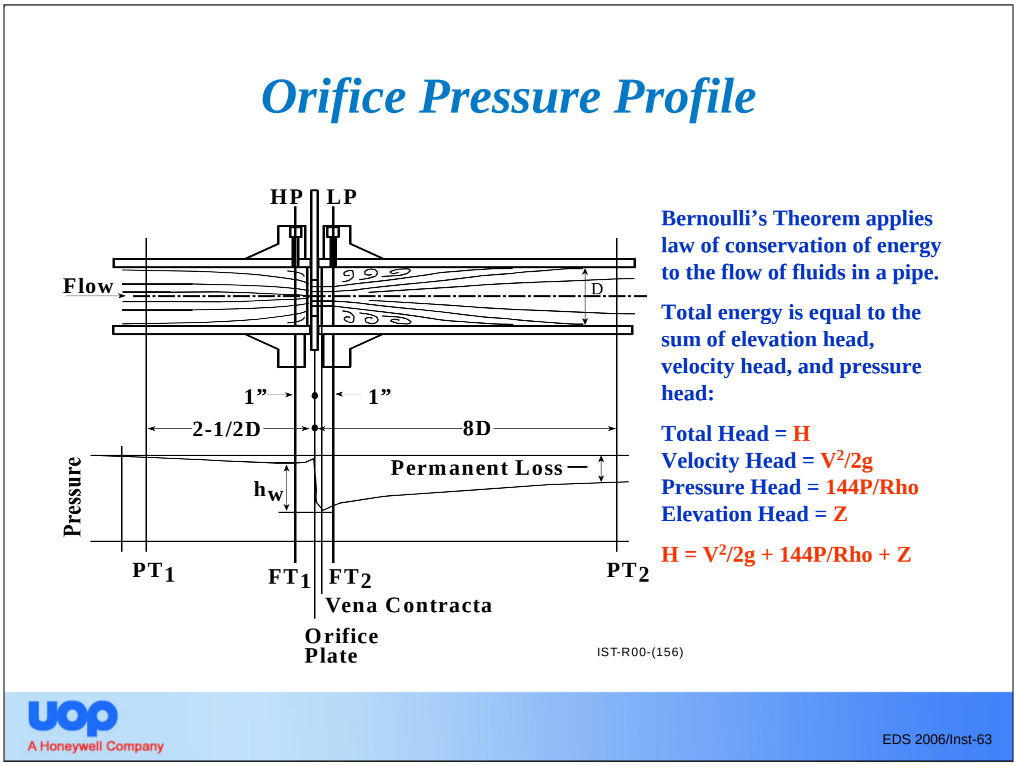


One of the more common process measurements encountered is the "flow rate". The industrial work horse for flow measurement, accounting for over 50% of the total flow devices, is the orifice plate and pressure differential transmitter.

The square edge, beveled, concentric orifice plate is used to measure either liquids or gases. Variations of this type of plate are known in industry to allow for certain fluid constraints. Weep holes are often drilled above or below the bore to allow for entrained liquids or gases to escape.

As shown above the eccentric plate is often used by UOP to measure % vaporization at a fired heater outlet. The eccentricity of the plate prevents a build-up of liquid upstream of the plate. Typical heater design has 40 to 60% vaporization at the heater outlet, the plate effectively is a measure of heat input in terms of vaporization (This application will be reviewed in more detail later on in the session.)

The orifice plate handle is often stamped with pertinent design information on the upstream side of the plate and is characterized by the "BETA RATIO". By definition: Bore Diameter/Pipe Inside Diameter is equal to BETA. An abundance of correlated data sets the beta limits between 0.1 and 0.7.



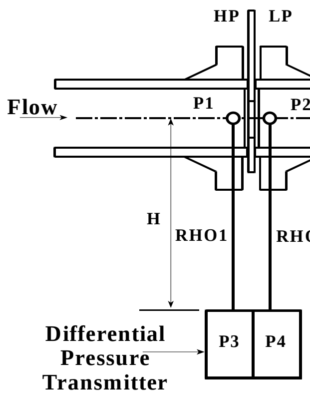
Flow through an orifice plate is based on Bernoulli's principle as it applies the law of conservation of energy to the flow of fluid in a pipe. The total energy at any particular point in the system is equal to the sum of the velocity head plus the pressure head plus the elevation head:

$$H = V^2/2g + 144P/Rho + Z$$

The Total Head of the fluid up-steam of the orifice plate is equal to the Total Head of the fluid as the fluid passes through the orifice plate; and for a horizontal meter run, the elevation head (Z) cancel. The cross-sectional area of the orifice plate is smaller than the cross-sectional area of the pipe, resulting in an increase in fluid velocity through the orifice plate (velocity = volumetric flow divided by cross-sectional area). The law of conservation of energy therefore indicates that the pressure head decreases as the fluid passes through the orifice plate. Down stream of the orifice plate the cross-sectional area expands once again and the fluid velocity decreases while the pressure head increases. Approximately 8 pipe diameters down steam of the orifice plate, all possible pressure recovery has taken place. The difference between the pressure upstream at this point and the pressure up-stream of the orifice plate is the permanent pressure loss or the energy absorbed by the system.

Flange taps (FT) or pressure taps (PT) can be used to measure a dP and correlated to flow. The industrial application has standardized on orifice flange taps 1 inch up-stream and 1 inch down-steam of the orifice plate.

Transmitter Static Head Effects



$$\begin{aligned} dP &= (P3-P4) \\ &= (P1+H \times RHO1)-(P2+H \times RHO2) \\ &= (P1-P2)+H(RHO1-RHO2) \end{aligned}$$

(where RHO1 and RHO2 are fluid densities)

A Honeywell Company

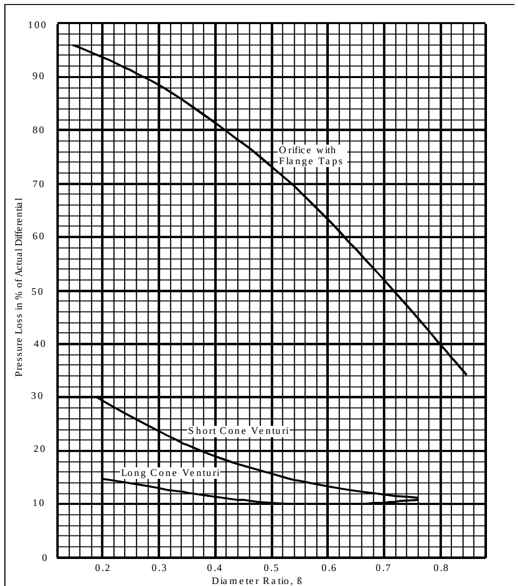
EDS 2006/Inst-64

For the typical liquid flow measurement installation the transmitter not only measures the orifice plate differential, but also measures the static head in the transmitter impulse lines. If the fluid densities are different (different composition or different temperature), then the differential pressure at the orifice plate will not be equal to the differential pressure at the transmitter. The resultant error increases with an increasing difference in densities and/or an increasing difference in elevation between the orifice plate and transmitter.

As a percent of flow this error decreases as the orifice differential increases. Therefore the higher the differential the less error in flow measurement. However an upper limit on orifice differential is governed by the economics of pumping the fluid. The industrial standard for most applications designs the orifice differential at 100 inches water column at the meter maximum. The meter maximum generally is chosen so the normal flow is between 70 and 80% of the meter maximum.

A square root relationship exists between flow and pressure differential. If the flow was at 70% of meter maximum, the measured differential pressure across the orifice plate would be 0.7 squared or 0.49 times 100 inches, which is 49 inches. This equates to about 50% of the transmitter calibrated span. Keep this fact in mind as we will see most instrumentation specified at 50% of its calibrated span or 50% of its capacity as in the case of control valves.

Permanent Pressure Loss



A Honeywell Company

EDS 2006/Inst-65

As we discussed earlier the pressure differential at the orifice plate taps and the pressure differential at the pipe taps (2-1/2 pipe diameters up-stream and 8 pipe diameters down stream) can be correlated as a function of beta ratio and also in terms of permanent pressure loss.

For the typical orifice flange installation, the above graph illustrates permanent pressure loss in % of actual pressure differential measured at the orifice flanges (y-axis) as a function of orifice plate Beta Ratio (X-axis).

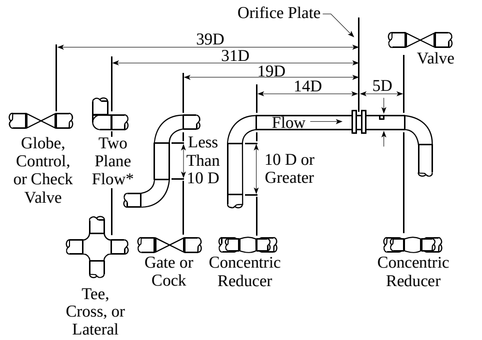
As for our typical meter installation, if the beta ratio is 0.70, the meter maximum is designed for a 100 inches water column differential at maximum flow, and the flow is at meter maximum; then the permanent pressure loss is 52% of the measured pressure differential. For this case at a flow rate equal to the meter maximum the permanent pressure differential is 0.52 times 100 inches water column or 52 inches water column. In terms of pressure units:

(Inches H20)(fluid density)(ft^2/144 inches^2)(ft/12 inches) = pounds per inch^2

(52 inches)(62.4 pounds/ft3)(ft^2/144 inches^2)(ft/12 inches) = 1.88 psi

With Beta ratios ranging from 0.1 to 0.7, the graph illustrates that permanent pressure loss will range from about 50% of the measured differential up to about 97%.   The graph also illustrates that the smaller the beta (higher resistance for a given flow rate) the greater the energy consumption will be.

Orifice Minimum Straight Run Requirements



*Where Distance Between Elbows is 10 D or Less

Note:
Dimensions are for Flange Tap Orifice Surfaces Only, Based on AGA-ASME Report No. 3

A Honeywell Company

IST-R00-(159)

EDS 2006/Inst-66

As stated previously, an orifice plate is characterized by its beta ratio and this ration will range from 0.1 to 0.7. Insufficient experimental data exist outside of this range, and the accuracy of the correlation can not be confirmed. Following UOP's engineering practice for line sizing, most installations will not require Beta ratios greater than 0.7. However in some cases where the line sizing is sized "tight" (approaches the upper limit of pipe pressure drop per unit length), the meter run may require swaging to a larger pipe size in order to keep the beta ratio below 0.7.

In addition to the beta ratio requirements, a certain amount of straight run pipe up-stream and down stream of the the orifice plate is required to insure that a fully developed turbulent flow profile exists at the orifice. This allows for application of the flow correlation developed from the experimental installations.

Also the amount of straight run pipe is a function of the beta ratio. The larger the bore for a given pipe size (increasing beta ratio), the greater the upstream pipe requirement.

The information provided above is UOP's recommended practice and is based on a beta ratio of 0.7. This is a conservative approach.

Basis for Orifice Minimum Straight Run Requirements

- **Upstream pipe diameter varies with Beta**
- **Minimum requirements based on Beta = 0.7**
- **Installations with shorter upstream pipe runs with smaller betas are not incorrect**
- **Being conservative allows for replacement of existing orifice plates with larger betas without the need to modify piping meter runs in the future**

UOP
A Honeywell Company

EDS 2006/Inst-67

UOP selected the upstream pipe requirements based on a beta ratio of 0.7. If the beta ratio is say 0.45, the correlated experimental data would require less straight run requirements than recommended in the previous slide.

UOP has chosen the data for 0.7 to be on the conservative side. For future debottlenecking and expansion projects, where the orifice bore may need to be increased to meet the process requirements of the expansion, changes to meter run piping will be held to a minimum; thus minimizing the cost associated with the expansion project.

Venturi Tube

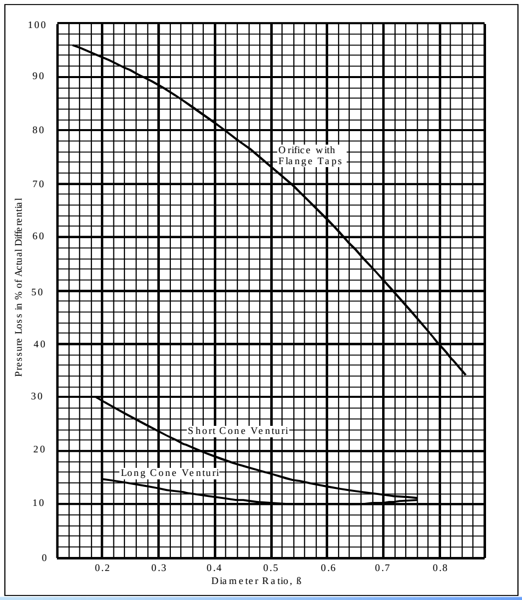
- **Alternative to the orifice plate**
- **Venturi’s taper allows for stream lining flow into/out of throat area**
- **Eliminates turbulence and boundary layer separation as in an orifice plate**
- **Pressure/Velocity heads nearly reversible (permanent pressure loss is minimized)**
- **Capital costs are high, but energy savings may be significant (centrifugal compressor)**

A second type of primary flow element is the venturi meter. The venturi is tapered both at the inlet and outlet and is designed to guide the flow streamlines in toward the throat area (minimum cross-sectional area), and to expand them back out to a fully developed velocity profile without developing the turbulence and boundary layer separation that occurs with an orifice plate. Like the orifice plate the venturi is also characterized by a beta ratio, with the limits of 0.4 to 0.7. (Beta ratio is defined as the Throat Diameter/Pipe ID.)

Due to the tapered design the static and kinetic energy exchange (pressure and velocity heads) is more reversible and the overall permanent pressure loss is held to a minimum. On the other hand the cost of a venturi meter is much higher relative to the orifice plate.

Venturi meters are often designed in centrifugal compressor circuits where line sizes are on the order of 10 inches or larger. For compressor circuits less pressure drop across the primary flow element translates into an overall reduction in horsepower requirements to operate the compressor. Payouts for the increase in capital cost between the venturi and orifice through a reduction in operating costs are exceptional fast.

Permanent Pressure Loss



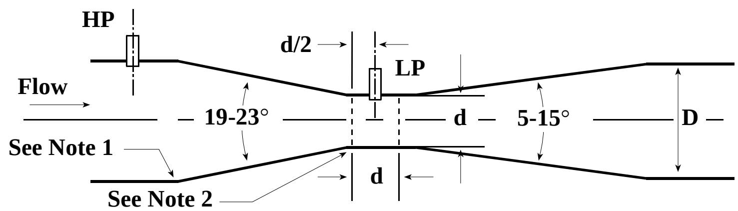
EDS 2006/Inst-69

Reviewing the plot of Permanent Pressure Loss in % of Actual Differential vs. Beta Ratio for a venturi, the permanent pressure loss is less than 20% over the recommended beta ratio range of 0.4 to 0.7 and less than 30% with beta ratios down as low as 0.2.

Herein lies the advantage of the venturi in terms of energy savings in the operation of centrifugal compressors.

The disadvantage of the venturi is its overall capital cost. Also on expansion projects, an orifice plate can be re-bored to a larger diameter or replaced inexpensively with another orifice plate. However the throat diameter of the venturi can not be altered without also making changes to the inlet and outlet cones and overall length.

Dimensions of Classical Venturi Tube



Notes:
1. Round to 0 to 1.375 D Radius
2. Round to 3.5 to 3.75 d Radius
3. LP Tap Radius <=d/6
4. For Herscel Venturi, Exit Cone Angle is 7-8°

A Honeywell Company

EDS 2006/Inst-70

For the classical venturi tube, the above depicts the typical dimensions. Atypical design could put the beta close to 0.7 in a 20 inch pipe. This implies that the throat diameter is approximately 14 inches. Base on these requirements and the above dimensional requirements for the classical venturi tube, the over length is on the order of 6.5 ft (approximately 2 meters in length).

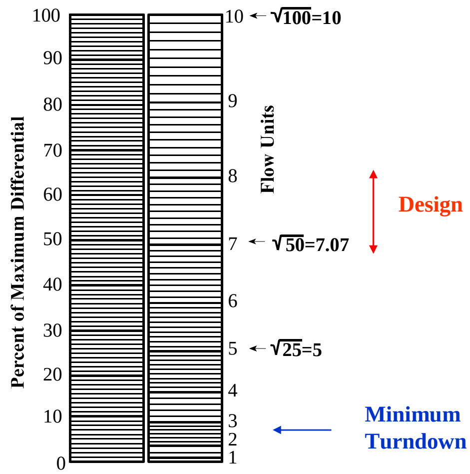
Square Root Relationship
Head vs. Flow

Flow is proportional to the square root of the differential pressure measured across the primary element.

Accuracy of the overall measurement usually expressed in terms of '% of full scale'.

Therefore a much larger error exists in flow at reduced rates than at the design rate.

Turn down usual limited to 4:1



uop
A Honeywell Company

EDS 2006/Inst-71

Flow is proportional to the square root of the differential pressure measured across the primary element. The above square root chart illustrates the <u>measured differential in % of maximum differential</u> with respect to <u>flow rate</u> from 1 to 10 units of flow (10 units of flow representing meter maximum).

Meter accuracies are generally ± 2% of full scale. If the meter maximum was 100 gpm and the normal flow was 75 gpm, the actual flow rate is 75 ± 2 gpm (which is 75 gpm ± 2.67%). If turndown is at 50% of normal, the actual flow rate is 37.5 gpm ± 2 gpm (which is 37.5 ± 5.33%). The bracketed accuracy of the measurement is getting much larger at reduced flows. UOP will typically limit the design of differential head type devices to a 4 to 1 turndown. Modifications to the design will be incorporated if the range of flows exceed 4 to 1 turndown. Various modifications can be employed, depending upon the design criteria. These modifications include re-spanning transmitters, installation of two transmitters (1 high span and 1 at a low span), and even the installation of parallel flow meter runs for extreme cases.

Additional Flow Meters

- **Turbine flow meter**
  - **Consists of straightening vanes, rotor with small imbedded magnet, and bearings**
  - **speed of rotor proportional to fluid velocity**
- **Electromagnetic flow meter**
  - **Fluid flows through a magnetic field and fluid must be conductive (hydrocarbons are not)**
  - **Voltage produced proportional to fluid velocity**

uop A Honeywell Company

EDS 2006/Inst-72

A multitude of other technologies are available for flow measurement and each has its unique place in the industry. Various laws of physics and electrical laws are used to correlate flow to some intrinsic property.

The turbine meter is an in-line flow measuring device that has reported accuracies of ± 0.25%. These are often specified and used on raw material and product steams for inventory control. These meters are costly compared to orifice plate installations, have a significant increase in pressure drop (potential increase in operating costs for pumped systems), and are potentially high maintenance items because of the rotating parts.

The “mag” meter on the other hand has no moving parts and essentially “zero” pressure drop because of no internal obstruction within the body. However for this meter to be able to measure flow the fluid must be conductive. Gases, steam, and most hydrocarbon liquids are non-conductive and do not produce any induced voltage when flowing through the meter’s magnetic field. Like the turbine meter, the “mag” meter is costly when compared to the orifice plate installation.

Additional Flow Meters
(continued)

- **Vortex flow meter**
  - **Consists of a triangular 'bluff body' spanning the inside diameter of the meter**
  - **Vortices are shed in a regular oscillating pattern**
  - **Frequency of oscillation proportional to volume**
- **Ultrasonic flow meter**
  - **based on the principle that the velocity of sound in a fluid in motion is the resultant of the velocity of sound in the fluid at rest plus/minus the velocity of the fluid itself**

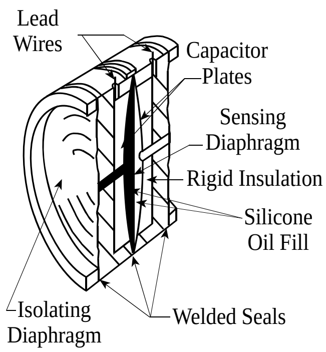
UOP
A Honeywell Company
EDS 2006/Inst-73

Another primary flow element that has gained much popularity over the past 10 years or so is the Vortex meter. A "bluff" body is installed in the cavity of the meter, which generates vortices in the flowing stream (down steam of the bluff body) that are shed at a rate proportional to the flow rate. Early designs were limited to temperature limits not to exceed 400 °F and could not tolerate any systems with minor amounts of pipe vibration. Current technology has addressed these issues, which account for the vortex meter gaining wide-spread acceptance.

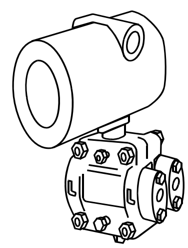
The last flow meter that we will briefly discuss is the ultrasonic flow meter. Based on either the Doppler effect or transit time of a sound wave to measure the velocity of a fluid. The transit time meter performs well in clean services whereas the Doppler effect meter performs well in dirty services.

For these last four devices, they either have internal moving parts or internal electrical parts that are subject to failure. When compared to the orifice plate which has neither, bypassing the flowmeter installation is required in order for maintenance to be conducted. Meter failure typically requires that the primary element be removed from the process piping. This generally is not the case for an orifice plate.

Diaphragm Type Differential Pressure Instrument



The δ–Cell



Rosemount Electronic DP Transmitter

uop
A Honeywell Company

EDS 2006/Inst-74

The orifice meter utilizes a differential pressure transmitter to convert the differential pressure across the orifice plate into a standard instrument signal (can be a digital, electrical, or pneumatic). Typical transmitter designs today use a capsule made from two very thin and flexible metal diaphragms on either side of a heavy support member that is designed to withstand a differential pressure equal to the full design pressure of the transmitter.

Some designs use strain gauges to measure the stress on the diaphragm generated from the differential pressure; some measure the resonant frequency of a wire that varies with the differential pressure; and some (as shown above) measure changes in capacitance as a function of differential pressure.

The δ–cell uses the latter technology to measure differential pressure. Pressure up stream of the orifice plate and pressure downstream of the orifice plate exert pressure on their respective isolating diaphragms. In turn the incompressible silicone fluid distorts the sensing diaphragm, which makes up part of the "capacitor". This distortion, on the order of 0.0004 inch maximum, changes the capacitance of the cell with respect to differential pressure. The two lead wires are connected to the electronic capsule on the top works.

Level Instruments
(Displacers)

- **Based on principle of buoyancy (effective weight is less when suspended in a fluid)**
- **Consists of cylindrical displacer suspended by a hanger in a chamber housing, torque arm, knife edge bearing, and torque tube (i.e. essentially a weigh scale)**
- **Zero point: displacer suspended in vapor space (weight$_{0\%}$ = weight of displacer)**
- **100% point: displacer immersed in liquid**

UOP A Honeywell Company

EDS 2006/Inst-75

Based on Archimedes' principle (a body immersed in a fluid is acted upon by a net force that is vertically upward and equal in magnitude to the weight of the fluid displaced by the body) UOP specifies external displacers for measurement lengths of 60 inches or less.

The displacer, which is not a float, is in a chamber typically external to the vessel. As the liquid level in the vessel changes, the fluid level in the external chamber rises/falls and the displacer is more/less immersed in the fluid. The total movement of the displacer from 0 to 100% immersion is less than 1/4 of an inch.

One manufacturer's design consists of a cylindrical displacer, hanger, torque arm, knife edge bearing, and torque tube. This assembly is essentially a weigh scale that is used to measure the effective weight of the displacer over the range of the displacer (0 to 100% immersion) as illustrated with the bullet points.

Typical displacer lengths available from most manufacturers are 14, 32, 48, 60, 72, 84, 96, and 120 inches.

Level Instruments
(Displacers, cont’d)

- **At 100% : weight$_{100\%}$ = weight of displacer - weight of hydrocarbon displaced)**
- **Effective weight over the range of displacer**
  - **EW = weight$_{0\%}$ - weight$_{100\%}$**
  - **EW = weight of hydrocarbon displaced**
- **For a given EW at a specified density (D):**
  - **Volume of HCBN = V$_{HCBN}$ = EW/D**
- **At constant cross-sectional area (A) :**
  - **Height of HCBN = V$_{HCBN}$/A**

uop
A Honeywell Company

EDS 2006/Inst-76

As shown over the range of the displacer the effective weight is equal to the weight of hydrocarbon displaced. Knowing the density of the fluid and the cross-sectional area of the cylindrical displacer, the effective weight measurement is proportional to the height of the fluid in the chamber.

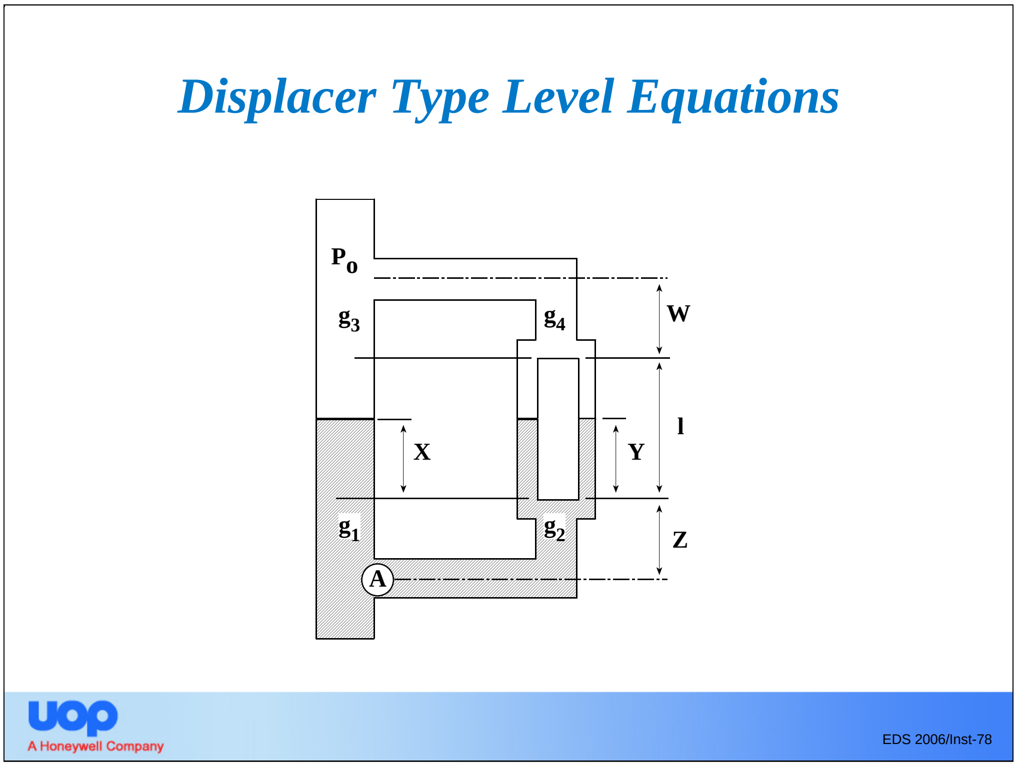
Level Instruments
(Displacers used for interface measurement)

- **Zero point: displacer suspended in HCBN ($weight_{0\%}$= $weight_{dispacer}$- $weight_{HCBN\ disp'd)}$**
- **100% point: displacer suspended in water ($weight_{100\%}$= $weight_{displacer}$- $weight_{H_2O\ disp'd}$)**
- **Effective weight over the range of displacer**
  - **EW = $weight_{0\%}$ - $weight_{100\%}$**
  - **EW = $weight_{H_20\ disp'd}$ - $weight_{HCBN\ disp'd}$**
- **Interface is contingent upon differences in gravity between the two fluids, > 0.1**

uop
A Honeywell Company

EDS 2006/Inst-77

The displacer can also be utilized for interface measurement. The most common application is an interface between a lighter hydrocarbon phase and a heavier water phase. Most designs for the common displacer will handle gravities (and gravity differences) between 0.1 up to 1.5. If the gravities of the two fluids are nearly equal, the interface level will not be distinguishable.



The complete chamber assembly comes in 4 orientations. The above diagram represents the Top/Bottom orientation. The Top/Side, Side/Bottom, and Side/Side orientations are the three remaining orientations, and define the location of the flanged connections. Depending upon the vessel style (horizontal vs. vertical), room availability, and actual measurement length and location, 1 of the 4 orientations will be selected to meet the design criteria and installation requirements.

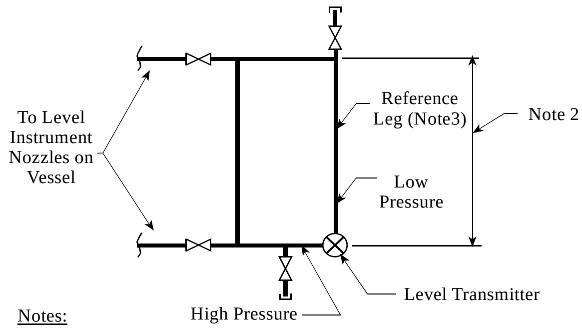
The standard transmitter has the advantage of a simple zero check. Any time the level is below the displacer, the transmitter should show 0% output. The zero setting is not affected by temperature or composition changes. The actual level measurement are affect by these changes. If the vessel temperature is relatively higher than the temperature of the fluid in the external chamber, the level in the vessel will be different than the level in the displacer due to difference in gravities.

$$Y = [X(g_1-g_3) + Z(g_1-g_2) + (w-l)(g_3-g_4)]/(g_2-g_4)$$

If well insulated, then g_1 and g_2 are equal and g_3 and g_4 are equal and essentially negligible relative to the liquid density. Therefore Y = X.

Differential Pressure Level Instrument

(Installation Detail)



Notes:
1. All Piping is 3/4"
2. Contractor Shall Advise Length of Reference Leg
3. Fill Reference Leg With Sealing Liquid Specified

A Honeywell Company

EDS 2006/Inst-79

For level measurement lengths greater than 60 inches, a differential pressure instrument is employed to measure the static head of the vessel fluid. The high pressure tap on the transmitter is connected at the lower vessel connection. The static pressure, P_{high} = LIQ HEAD + P_o, is the pressure exerted on the high pressure side of the transmitter. The low pressure tap is connected at the upper vessel connection. The static pressure, $P_{low} = P_o$, is the pressure exerted on the low pressure side of the transmitter. The differential pressure is:

$$\Delta \mathbf{P} = P_{high} - P_{low} = \text{LIQ HEAD} + P_o - P_o = \text{LIQ HEAD}$$

Vapors at or near the dew point of the vapor pose a unique situation with this type of application. As vapors condense in the impulse line on the low pressure side (reference leg) of the transmitter, liquid build-up in the impulse line provides a negative static head effect. In the worse case a true 100% liquid level would indicate a 0% level (static head on both sides of the cell would be the same provided the densities were equal). Filling of the reference leg could also occur, if during an upset, the vessel level is becomes higher than the upper vessel connection.

To compensate for this type of application or mis-operation, the reference leg is filled with the same fluid as that inside the vessel prior to or during unit commissioning. The transmitter must have zero elevation capabilities.

Differential Pressure Level Instrument

- **Measure static head as fluid level changes**
  - **Static head = level span x fluid specific gravity**
- **Differential pressure instrument**
  - **High side = static head + vessel pressure**
  - **Low side = vessel pressure**
  - **ΔP = static head**
- **Fill fluid required on low side of transmitter whenever fluid is at or near bubble point**
- **Transmitter capable of zero elevation**

A Honeywell Company

EDS 2006/Inst-80

The differential pressure is:

$$\boldsymbol{\Delta P} = P_{high} - P_{low} = \text{LIQ HEAD} + P_o - P_o = \text{LIQ HEAD}$$

The calibrated span for the transmitter, similar to the flow transmitter, is in terms of “inches water column” or its metric equivalent. The calibrated span is determined by multiplying the measurement length (level span) times the fluid operating specific gravity.

The fill fluid can be something other than the actual liquid in the vessel. However this material must be compatible with the process. Contamination of the process could lead to catalyst poisoning or other ill effects on the process. In colder climates a 50-50% mixture of glycol and water can be used.

Zero elevation is a feature with most transmitters that compensates for the negative differential pressure induced by filling the reference leg with a sealing fluid. As long as the correct transmitter has been specified, the negative differential pressure present in the reference leg is “zeroed” out during instrument calibration.

Analytical Type Devices

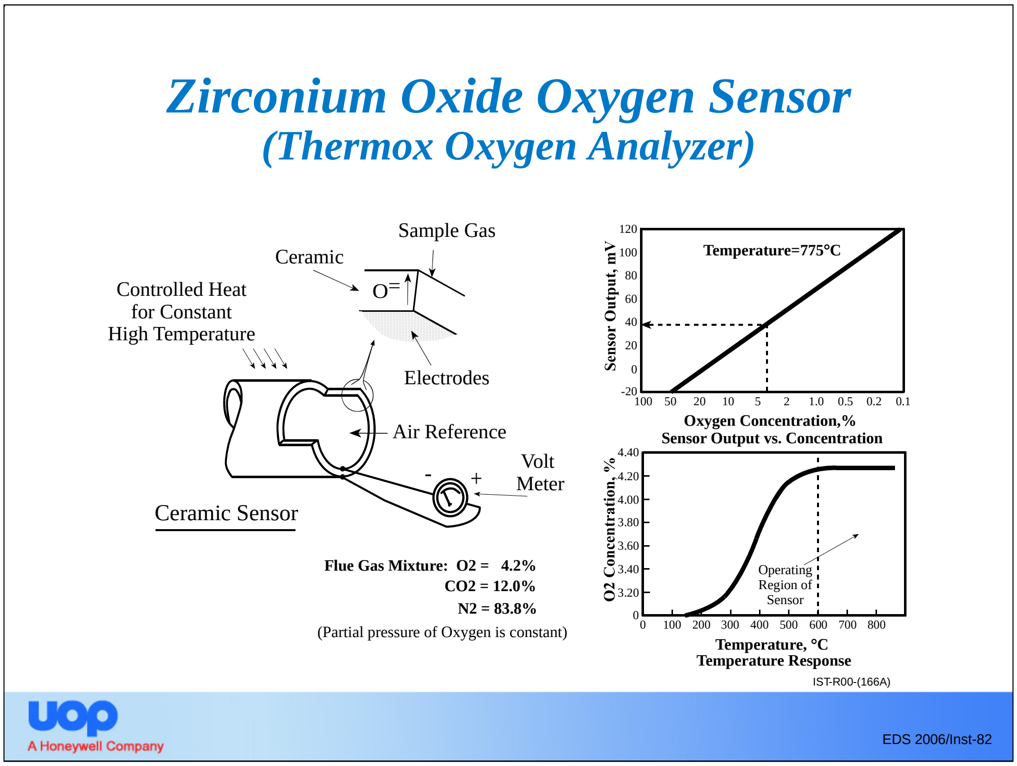
- **Open Loop Devices**
  - **Relies on operator to take corrective action**
  - **Quicker analysis than supplied by Lab**
- **Commonly specified analyzers**
  - **Moisture monitors**
  - **Oxygen Analyzers**
  - **Gas/Liquid Chromatographs**
- **Sampling system, if required, will add complexity to the operation of the analyzer**

A Honeywell Company

EDS 2006/Inst-81

Process analyzers are usually specified by UOP as open loop devices and requires the operator to take corrective action on the process. The advantage of on-line analysis is a much quicker update of the process measurement in comparison to laboratory analysis. Their main justification comes in during upset conditions. Being able to supply the operator with a much quicker analysis during upsets, allows the operator to make adjustments in a more timely manner minimizing off-spec product.

Moisture, oxygen, hydrogen, and chromatograph analyzers are among the more common analyzers specified by UOP. Some analyzers are in-line, that is the probe is installed in the process piping. Others have additional piping to route a slip-stream to and from the process and some even incorporate an elaborate sampling system to “cleans” the process material of any contaminants which may interfere with the operation of the analyzer. Lag time is minimized by designing the slip-stream with a relatively high flow rate.



Oxygen analyzers are used most commonly in the flue gas from fired heaters. The sensor is made of zirconium oxide ceramic coated platinum. Instrument air is used as a reference on one side of the ceramic, and the flue gas on the opposite side.

At the operating temperature of about 1285 degF, a voltage is developed between the two sides of the ceramic in accordance with the Nernst Equation.

Thermox Oxygen Analyzer

Nernst Equation: $E = AT \text{ Log } (20.9/0_2)$

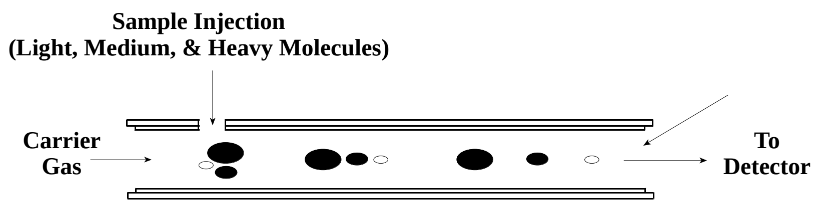
Where 'A' is a constant, 'T' is the cell temperature, and 'O_2' is the oxygen content in the sample gas.

uop A Honeywell Company EDS 2006/Inst-83

The zirconium oxide analyzer comes in an in-situ version and an extraction type arrangement. The latter version requires an eductor to withdraw the sample from the heater stack.

UOP also specifies portable oxygen analyzers for checking oxygen breakthrough during catalyst regeneration steps. The portable oxygen analyzer, a Teledyne Micro-Fuel Cell, produces an electric current proportional to the O2 concentration in the sample by means of the above chemical reactions.

Gas-Liquid Chromatography



A Honeywell Company

EDS 2006/Inst-84

Gas-Liquid Chromatography is the most widely used form of Chromatography in the Process Industry. With similarities to Paper Chromatography, the typical chromatograph is based on the differences in rats of adsorption/desorption of the various molecular components.

The chromatograph is composed of a long, thin, packed column and accessories to vaporize liquid samples. The packing is coated with a liquid which absorbs and desorbs the sample components. A sample, either liquid or gas, is injected and flushed through the packed column by a carrier gas to the detector. Because of the adsorption/desorption mechanism, the lighter components travel through the packed column at a faster rate than the heavier molecules, thus separation of the molecules is achieved.

Specific chromatograph designs can distinguish between some components in ranges on the order of 100's ppm levels. In some UOP advance process control applications, the chromatograph specifications require some 30 to 35 component analyses. In more basic applications, one or two key component analysis is required in the operation of fractionating columns. As with most instrumentation, the more complex the system the more costly that system will be.

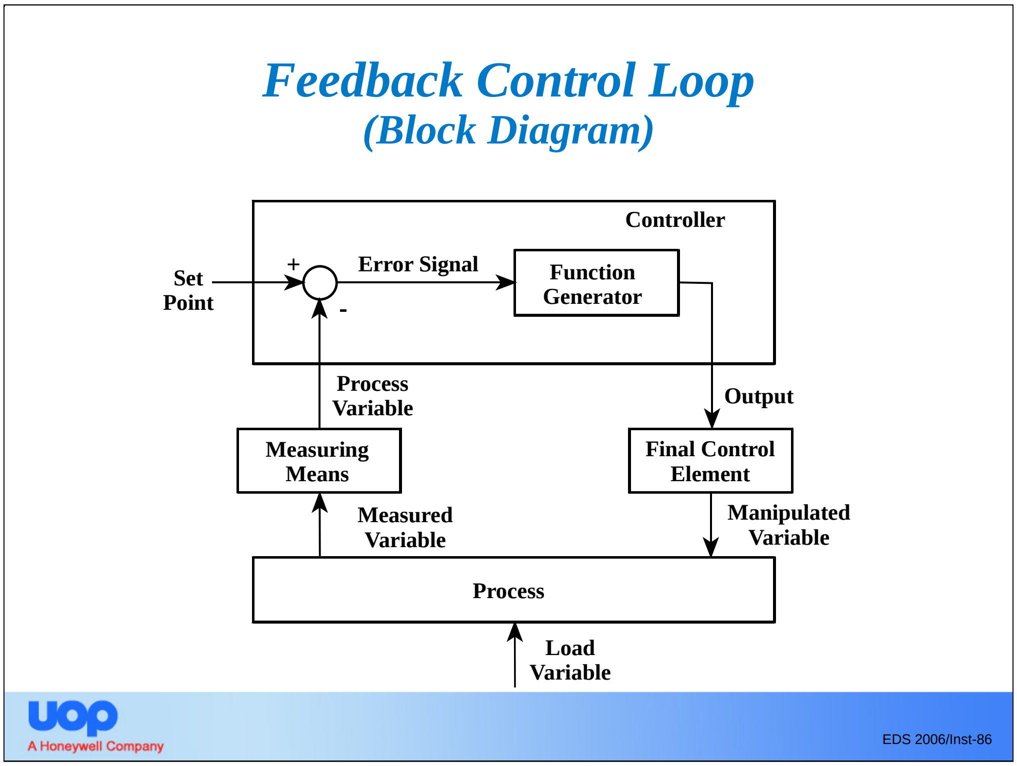
Feedback Control Loop

- **Loop Components**
  - **Process**
  - **Measuring Means**
    - **Temperature, Flow, Pressure, Level, Analysis**
  - **Final Control Element**
    - **Control Valve Body Types**
    - **Control Valve Performance and Accessories**
    - **Trim Characteristics**
    - **Hydraulics and Sizing Equations**

A Honeywell Company

EDS 2006/Inst-85

We have now completed the review of the most common types of instruments to measure such variables as temperature, flow, pressure, etc. These devices are used to measure the process variable. Next we will investigate the most common final control element - THE CONTROL VALVE. There are several body types, performance issues, trim characteristics and sizing issues.



Thus far we have investigated the Process and the Process Variable side of the controller. This is the means of being able to measure the process variable, which is an input to the process controller.

Now we will look at the Final Control Element, which is the mechanism that varies the Manipulated Variable in order to move the Process Variable in the direction of the Set Point.

Final Control Element

- **Control valve is the most often used final control element in the Process Industry**
- **Control valve dissipates energy in order to control the process (adds operating cost)**
- **Alternatives to the control valve**
  - **Variable speed/variable stroke, pumps**
  - **Variable speed turbines, centrifugal compressor**
  - **Variable pitch fans, air condensers**

uop
A Honeywell Company

EDS 2006/Inst-87

The most often used final control element in the Process Industry is the control valve. The control valve is essentially a variable resistance, located somewhere in the process line between Point A (the source) and Point B (the destination), and dissipates energy of the system in order to "throttle" the manipulated variable. For a pumped system the dissipated energy adds operating costs, by increasing incrementally the horsepower requirements of the pump. Good engineering practices can optimize the amount of wasted energy, yet maintain some guarantee of additional throughput above the normal design flow rate.

Alternatives to the control valve are variable speed drivers and controllers. In the case of centrifugal compressors, variable speed turbines are often part of the compressor drive package. As we say earlier with the selection of a venturi flow meter over the conventional orifice plate, the venturi flow meter reduces the overall pressure drop of the system (by a couple of psi at the most).

With a variable speed compressor, the work required to operate the compressor can be minimized without the energy dissipation across a control valve. This energy savings can be a significant reduction in overall operating costs.

Control Valves

- **Single-seated globe style valve for services requiring 2 inch and smaller body size**
- **Quarter turn rotary globe style valve for sizes > 2 inch (limited in flange rating)**
- **Cage-guided globe style valves where rotary globes are not suitable**
- **Butterfly (wafer) valve for low delta P**
- **Multi-stage angle valve for high delta P**
- **Globe valve for high noise in vapor service**

uop
A Honeywell Company

EDS 2006/Inst-88

Control valves come in a variety of styles. Globe, rotary globe, ball, and butterfly are among the most common styles. Other styles and modifications to the most common styles are available for special applications.

UOP will specify the single-seated, sliding stem, globe valve with a pneumatic spring/diaphragm actuator for 2 inch and smaller body sizes. This essentially will be 1, 1-1/2, and 2 inch control valves. The single-seated valve is available from some manufacturers up to and including 6 inch body sizes. However actuator sizing, dependent upon the maximum valve differential pressure, becomes an issue for 3 inch and larger valves.

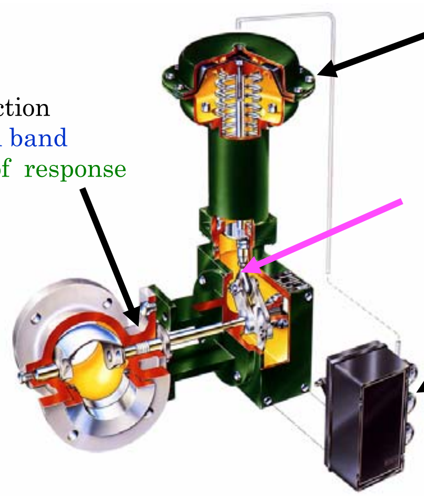
For 3 inch and larger valves, UOP will specify the quarter turn rotary globe style (eccentric globe) control valves. These valves often have limitations for maximum flange rating (600 RF) and differential pressure shutoff. Also the rotary globe is not capable of reducing noise via a noise reducing trim package.

When the rotary globe control valve can not meet the mechanical requirements of the system, then UOP will specify the cage-guided globe style control valve. This style of valve is a “balanced” design and actuator sizing is minimized due to this ‘balanced’ design.

Control valve performance is also an issue and is greatly affect by the various components and accessories associated with the valve design.

Control Valve Performance Overview

Valve Dynamic Performance Variables



uop
A Honeywell Company

EDS 2006/Inst-89

There are a number of control valve components that effect the control valves ability to provide good position control.

Hysteresis is the effect of a valve following a non-linear oval curve when first fully opening and then fully closing.

Dead band is the effect that occurs when the control valve signal is changing, but the valve is not moving. Usually valve dead band dominates over hysteresis.

Dynamic response is how fast a valve moves.

The positioner, actuator, mechanical linkages and packing all contribute to the overall hysteresis, deadband and dynamic response of a valve.

Effect of Control Valve Performance

- **Poor control loop performance, for many loops, can be directly linked to the poor dynamic performance of the control valve**
- **Because of poor control valve dynamic performance, most control loops increase rather than decrease process disturbances**
- **Typically users size a control valve based on process capacity requirements without regards to any dynamic performance criteria**

UOP
A Honeywell Company

EDS 2006/Inst-90

In the past several years more attention has been given to control loop performance and the dynamic performance of the control valve itself. The control valve is a critical part of the control loop and not all control valve and control valve accessories are created equal. If the valve is in need of maintenance or an inferior valve has been installed in the process, poor control valve performance can increase process disturbances.

Along with defining the hydraulics and valve sizing, UOP has a standard specification in which we try to address dynamic performance in terms of valve accessories and specific response capabilities.

We can think of a control loop as an instrument chain. Like any other chain, “the whole chain is is only as good as its weakest link”. We must insure that the control valve is not this weak link.

Control Valve Performance
(Control Valve Step Resolution)

- **The minimum step change in input signal to which the control valve system will respond while moving in the same direction**
  - **This phenomenon is caused by the tendency for a control valve to stick after coming to rest**
  - **This is also known as stiction**

We will cover some of the definitions in our latest standard specification, so that we can have a better understanding of control valve performance. How much does the input signal have to increase in order for the valve to move? Control Valve Step Resolution defines the minimum step change in input signal before the valve does actually move.

Control Valve Performance
(Control Valve Static Dead Band)

- **The range through which the input signal to the control valve can be changed, without the control valve moving position**
  - Dead Band **results from various phenomena such as backlash and stiction (friction) and causes the valve system to require extra input change after a reversal of direction before actual movement is resumed**

uop A Honeywell Company

EDS 2006/Inst-92

In a typical control loop the control valve is modulating i.e. constantly moving its position to satisfy the process setpoint. Excessive static dead band is seen as limit cycling. Often it is impossible to tune out the limit cycling without making the control loop very unresponsive.

Control Valve Performance Criteria

- **Control Valve Dead Time (T_d)**
  - **Time it takes after a change in input signal for the valve to start moving**
- **Control Valve Step Response Time (T_{63})**
  - **Time it takes after an input step change, for the valve to have moved to 63% of the step change**
  - **This includes the valve's dead time (T_d)**
- **Control Valve Hysteresis**
  - **Combined effect of step response and dead band that prevents changes in valve travel**

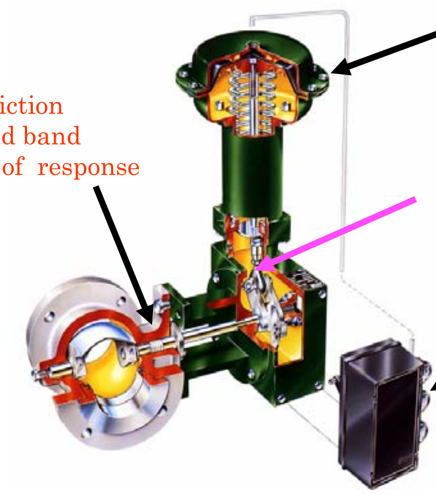
UOP A Honeywell Company

EDS 2006/Inst-93

As was the case for transmitter performance once again performance is a measure of dead time and the time constant. For the control valve the time constant is defined as the Control Valve Step Response Time.

Control Valve Performance Overview

Valve Dynamic Performance Variables



Actuator design
effects speed of response

Packing friction
effects dead band
and speed of response

Mechanical linkages
effects dead band

Positioner design
effects dead band, and
speed of response

uop
A Honeywell Company

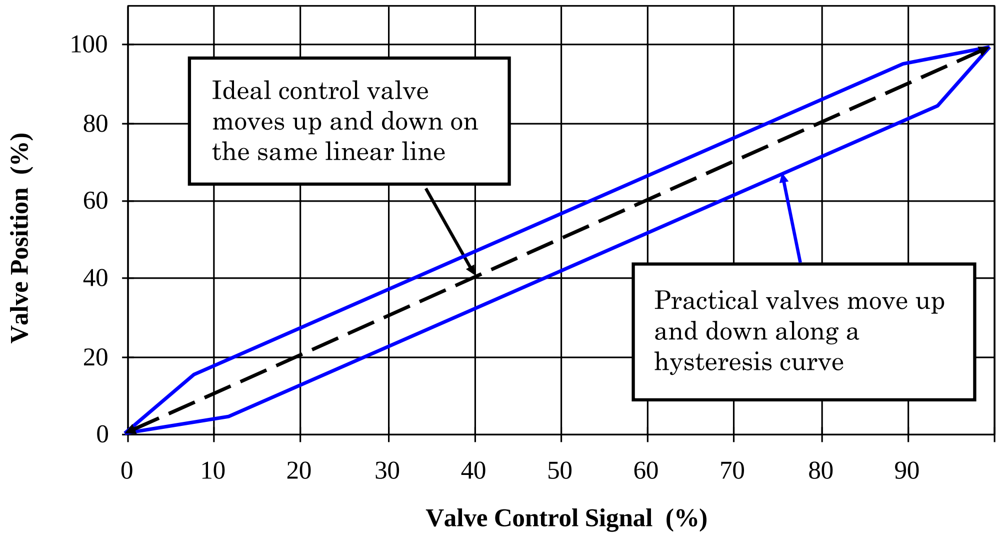
EDS 2006/Inst-94

Several key components of the valve affect Valve Dynamic Performance:

Actuator design and valve packing friction affect speed of response.
Mechanical linkage and valve packing friction affect dead band.

Control Valve Performance

(Control Valve Hysteresis)

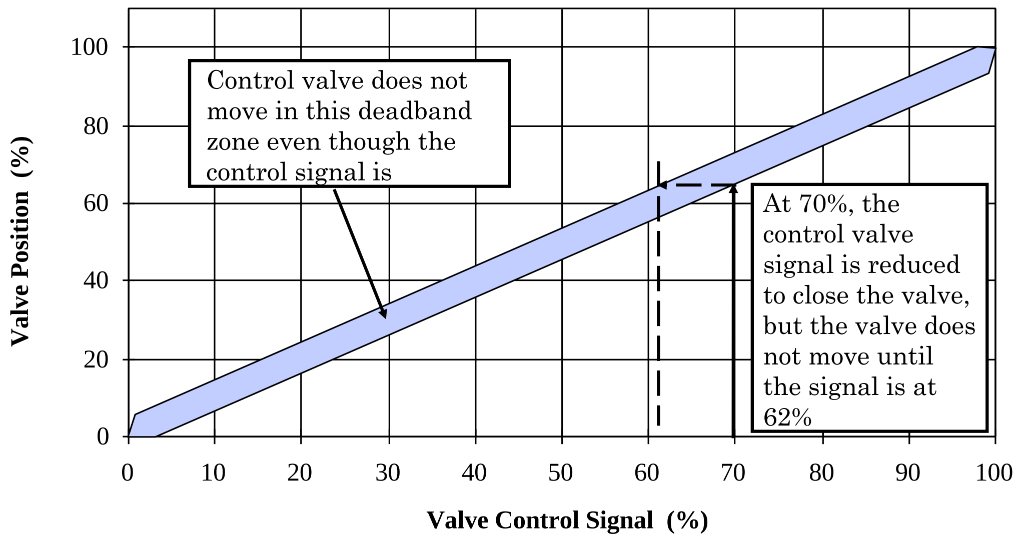


Hysteresis is the effect that a control valve moves up and down on an oval curve as shown above. The main difference with hysteresis compared to deadband is that the valve always moves (although on a curved line rather than the ideal linear line) when the control signal changes.

With deadband the control signal changes but the valve doesn't move.

Control Valve Performance

(Control Valve Dead Band)



UOP A Honeywell Company

EDS 2006/Inst-96

With dead band we have an effect that worsens control loop performance.

This is due to the deadband zone (the shaded zone) where the control valve does not move although the control valve signal is changing.

As an example, the control valve signal is at 70%. Due to the control loop error signal, the controller decides to start closing the valve. Because of the dead band zone, the control valve does not start to close until the control signal drops to 62%. This is very detrimental to good control and valve performance.

Control valves should be selected with the smallest dead band possible (<1%).

Control Valve Performance
(Control Valve Actuator)

- **Control Valve Actuator**
  - **Moves valve stem and plug relative to the controller signal**
  - **Provide fail-safe position for the valve**
  - **Typically operates on a pneumatic signal**
- **Actuator Types**
  - **Pneumatic spring diaphragm**
  - **Pneumatic spring return piston**
  - **Double-acting piston**

uop A Honeywell Company

EDS 2006/Inst-97

In closed loop control, the output from the controller is directed to the final control element and thus "throttles" the manipulated variable. An actuator by definition is "a pneumatic, hydraulic or electrically powered device that supplies **force and motion** to open or close a valve". It is this valve accessory that positions the valve plug relative to the control signal.

The pneumatic spring diaphragm actuator is the most popular actuator for sliding stem control valves, such as the single-seated and cage-guided globe valves discussed earlier. Various styles include direct acting (increasing air pressure pushes down diaphragm and extends actuator stem) and reverse acting (increasing air pressure pushes up diaphragm and retracts actuator stem). The actuator spring opposes the air pressure pushing on the diaphragm and it is this opposing spring force that will move the valve to its fail mode on loss of control signal.

The pneumatic spring return piston actuator provides a much higher steam force output than the spring diaphragm actuator. For services with high differential shutoff pressures, the piston style actuator must be used. The piston style actuators take advantage of higher instrument air supply pressures, thus providing the higher steam force.

Control Valve Performance
(Control Valve Actuator, cont'd)

- **Actuator sizing criteria (spring diaphragm)**
  - **Required force (thrust = T_m) must seat valve at maximum specified shutoff pressure(dP_{max})**
  - **T_m = (valve seat cross-sectional area) (dP_{max})**
  - **A = T_m/(maximum control signal air pressure) (A = Actuator diaphragm cross-sectional area)**
  - **Once A is known, pick next largest actuator**
- **Dynamic forces, process dP, actuator spring, stem packing, hysterisis, etc. prevent exact positioning**

UOP
A Honeywell Company

EDS 2006/Inst-98

Actuators are sized by comparing the required force to stroke a control valve with an actuator that can supply the force. The major force required to operate a globe valve include the sum of the static unbalance of the valve plug, seat load, and packing friction.

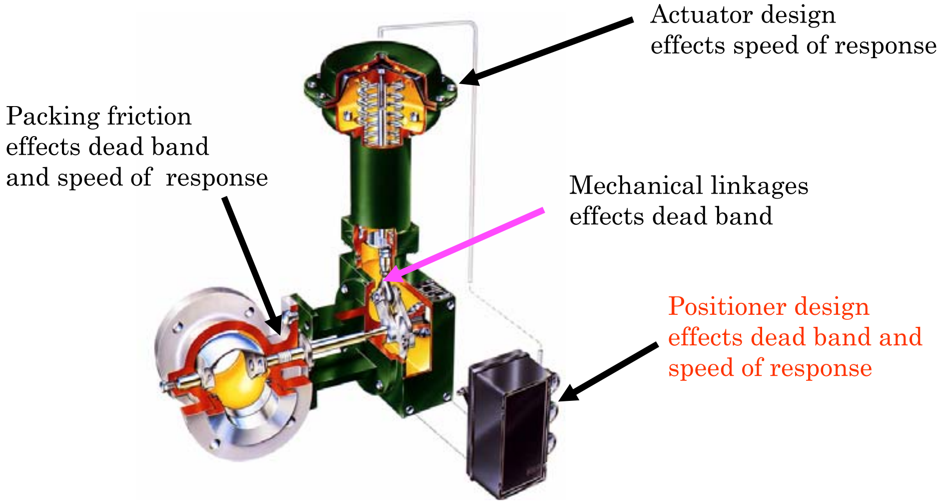
The unbalance force results from the process fluid when the valve is in the closed position with maximum differential pressure imposed on the control valve. The unbalanced area for the selected trim must be provided in order to determine the required force. As the control valves get larger for a given differential pressure, the required actuator size gets larger.

Seat load is also determined by shutoff requirements. Leak classes have been developed, and UOP typically specifies Class IV shutoff. Class IV shutoff by definition sets the Maximum Leakage Allowable at 0.1% of rated capacity.

Packing friction is determined by stem size, type of packing, and the quantity of compressive load placed on the packing either by the bolting or by the process.

Control Valve Performance Overview

Valve Dynamic Performance Variables



uop
A Honeywell Company

EDS 2006/Inst-99

The Positioner is a position controller that is mechanically connected to a moving part of a final control element and that automatically adjusts its output to the actuator to maintain a desired position in proportion to the input signal.

Recent advancements in positioner design has improved on control valve dead band and control valve speed of response. Choosing the right positioner can eliminate some of the effects of dead band and improve the speed of response of the control valve.

Control Valve Performance
(Control Valve Positioner)

- **Purpose of Control Valve Positioner**
  - **To maintain and control valve position**
  - **To compensate for control valve dead band**
- **Dynamic performance of control valve is influenced by the type of positioner**
  - **Single stage pneumatic spool valve positioner**
  - **2-stage high performance pneumatic positioner**
  - **Digital high performance positioner**

UOP A Honeywell Company

EDS 2006/Inst-100

The mechanical feedback from the actual valve position allows the positioner to adjust its output to the actuator to maintain a desired position in proportion to the process controller input signal within the limitations of the system. Most spring diaphragm type actuators can withstand air pressures up to 60 psig without damage to the diaphragm. This is the limiting factor in determining thrust requirements for a given actuator size.

The most important feature of a good positioner for increased dynamic performance is that it be a high gain device. Unless the positioner is sensitive to small input signal changes, the valve assembly will not be able to respond to minor disturbances in the process variable. Therefore the positioner must be designed such that it responds quickly to these small changes in input signal.

A second feature of a good positioner is that it must supply the power (in the form of air supply) needed to move the valve quickly. This power comes in the form of rapid air flow as needed to move the valve.

The simplest positioner is the single state pneumatic spool valve positioner. In order to meet the second feature, the spool valve must be modified in order to supply the rapid air flow to the actuator. However this modification increases overall air consumption for the control valve assembly.

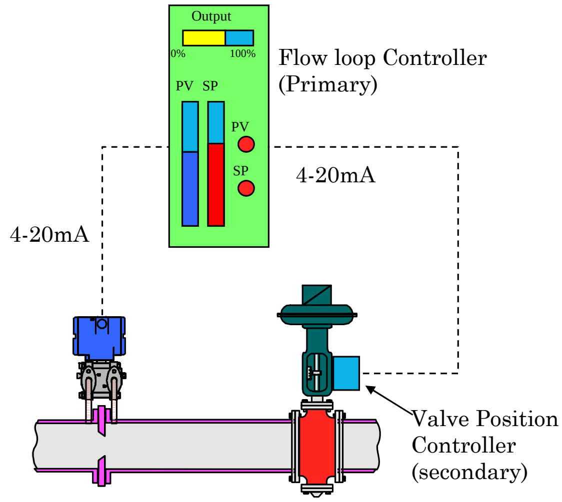
Two-stage positioners, although more complicated, give excellent dynamic performance with minimal steady-state air consumption.

Control Valve Performance
(Positioner/Cascade Loop)

Positioner is the secondary controller in a cascade loop

The control valve position controller must have a speed of response (dynamic response) superior to that of the loop (primary) controller.

This is very important in fast control loops such as flow.



A Honeywell Company

EDS 2006/Inst-101

For any control loop, the response time of the control and measuring equipment must be less than the response time of the process.

For the control valve positioner, it's speed of response must be superior to that of the loop controller.

For flow loops or other fast loops, only high performance positioners will help provide good valve performance.

Control Valve Performance
(Positioner Comparison)

	Position Control	Speed of Response	Design Complexity	Micro-Processor
Single Stage	±1.5%	Slow	Simplest	NO
Two Stage	±1.0%	Fast	Complex	NO
Digital HP	±1.0%	Fast	Complex	YES

uop
A Honeywell Company

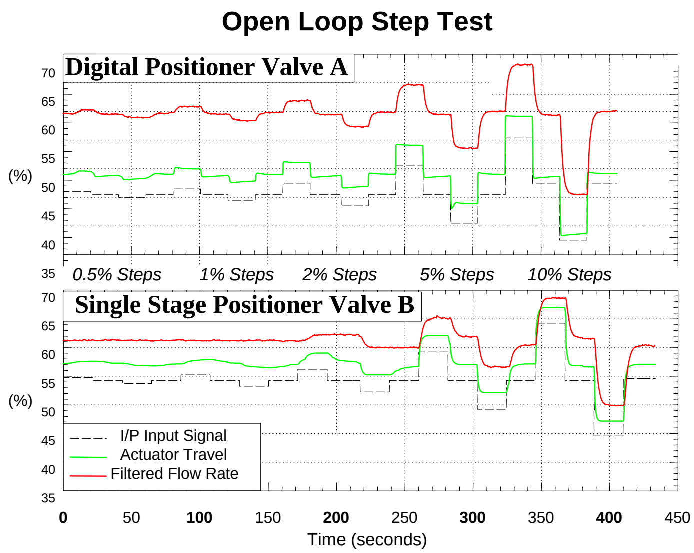
EDS 2006/Inst-102

Because of its simplicity, the single state positioner is the most common positioner provided with control valve assemblies. A single stage positioner may have the simplest design but provides the worst position control. With the increasing emphasis upon economic performance of process control, better performing positioners should be considered. In the past, maintenance personnel often liked the single stage positioner for it's simplicity and ease of maintenance. The performance of the single stage spool valve, internal to the positioner, in controlling the air pressure in the control valve actuator, limits the ability of the control valve to provide position control to no better than +/- 1.5%.

The two stage design provides better position control and if set up correctly can provide position control to within +/- 1 %. It is more difficult to set up than a single stage. Good quality dry and oil free air is required for good performance otherwise the air quality effects the performance of the first stage.

A digital valve controller (positioner) should provide the same level of control as a two stage positioner with the added benefits of the digital interface for set-up and diagnostics. Digital high performance positioners are invariably easy to set up, provide very good valve position control and often have the capabilities to perform control valve diagnostic tests.

Control Valve Performance
(Positioner Testing)



The above flow loop test response shows that valve "A" responds to 0.5% step changes in control signal. Valve "B", typical of many installed control valves, starts responding correctly to step changes of magnitude 5% and greater in control signal.

Unless users start specifying dynamic performance criteria when purchasing control valves , then they will always have to tolerate the type of control valve response shown by Valve "B". Dead band is a major contributor to excess process variability, and the control valve assembly is often a primary source of dead band in the control loop.

In both valves, the actuator stem motion (green) changes in unison with the input signal (black) changes. However for Valve B, 2% step changes resulted in the valve faithfully moving in conjunction with changes in input signal. This can also be seen in Valve A with changes in flow rate (red). Step changes on the order of 0.5% in input signal resulted in changes in flow rate; where as with Valve B, step changes in the order of 2% were required to obtain a corresponding change in flow rate response.

Control Valve Performance
(Dynamic Response)

- **Dynamic Performance (dynamic response)**
  - **A measure of how well a control valve performs at constantly changing valve position to meet the demands of modulating, multi-frequency disturbances to a closed loop process**
  - **One way of measuring control valve dynamic performance is to see how well it tracks the controller set point**
  - **Deviations away from the set point is known as variability (high variability,measured in %, shows poor control)**

UOP A Honeywell Company

EDS 2006/Inst-104

The real test of a control valve is how well it handles disturbances (i.e, load changes) and how well it tracks the controller setpoint.

The following is a rough measure of variability, deviation away from set point:

Very good variability is 0.5 to 2 %

Good variability is 2 to 4 %

Poor variability is 5 to 10 %

Extremely poor variability is >10%

Therefore during normal operation with the process in automatic, a control audit could be used to measure variability. The various control loops could be evaluated based on criticality and variability. Once identified as a control loop with high (poor) variability and deemed as a critical loop, an investigation could be launched to identify and improve the dynamic response of the control loop.

Control Valve Performance
(Requirements)

- **Control valve performance specification**
  - **Primary objective of the specification is to control a process variability to within the required variability limits**
- **Unfortunately there is no one performance specification for all applications**
- **Throughout Industry a large variation in the performance of control valves exist**
- **Users must define the specification and then find valves that can meet the specification**

A Honeywell Company

EDS 2006/Inst-105

Control valves need to do more than just work (i.e move eventually with a large change in control signal). Hysteresis, dead band and speed of response are all parameters that effect the control valves dynamic performance.

For Control Valve Dynamic Performance:

Need to define limits on hysteresis and dead band.

Need to specify speed of response.

It's a control valves ability to have good dynamic performance that reduces variability (i.e. control a variable, flow, pressure etc. to specific limits of deviation)

Is there a standard? Unfortunately not yet, although the Instrumentation, Systems, and Automation Society (ISA) is looking at such a standard.

Control Valve Performance Specification
(Definitions)

- **Step Resolution**
  - **The minimum step change in input signal to which the control valve system will respond while moving in the same direction**
- **Dead Band**
  - **The range through which the input signal to the control valve can be changed, without the control valve moving position**
- **Overshoot**
  - **Maximum amount in excess of step change**

UOP
A Honeywell Company

EDS 2006/Inst-106

The only current dynamic performance standard is the one developed by Entech.

The Entech standard was developed for the paper mill industries and was developed by Entech engineering consultants to allow them to develop loop performance criteria for control loops in the paper industry. They found that loop controllers were very difficult to tune for optimum control because of poor performance in the control valves.

The performance criteria are geared for control valves in the paper industry.

The Entech standard may not be the ultimate standard for all control valves in the process industry, however this type of standard should be developed by each user of control valves to produce a specification that meets their requirements. As such the Entech standard is a good model to use.

It should be noted that this Entech standard cause many control valve vendors to redesign their valves to meet this standard.

UOP has formulated a standard for control valve performance and has used the Entech standard as a model and applied it to the Process Industry. These are some of the definitions and criteria that appear in UOP's standard.

Control Valve Performance Specification
(Definitions cont'd)

- **Dead Time (Td)**
  - **Time it takes after a change in input signal for the valve to start moving**
- **Step Response Time (T63)**
  - **Time it takes after an input step change, for the valve to move 63% of the step change**
- **Step Response Time (T86)**
  - **Time it takes after an input step change, for the valve to move 86.5% of the step change**

UOP
A Honeywell Company

EDS 2006/Inst-107

The only current dynamic performance standard is the one developed by Entech.

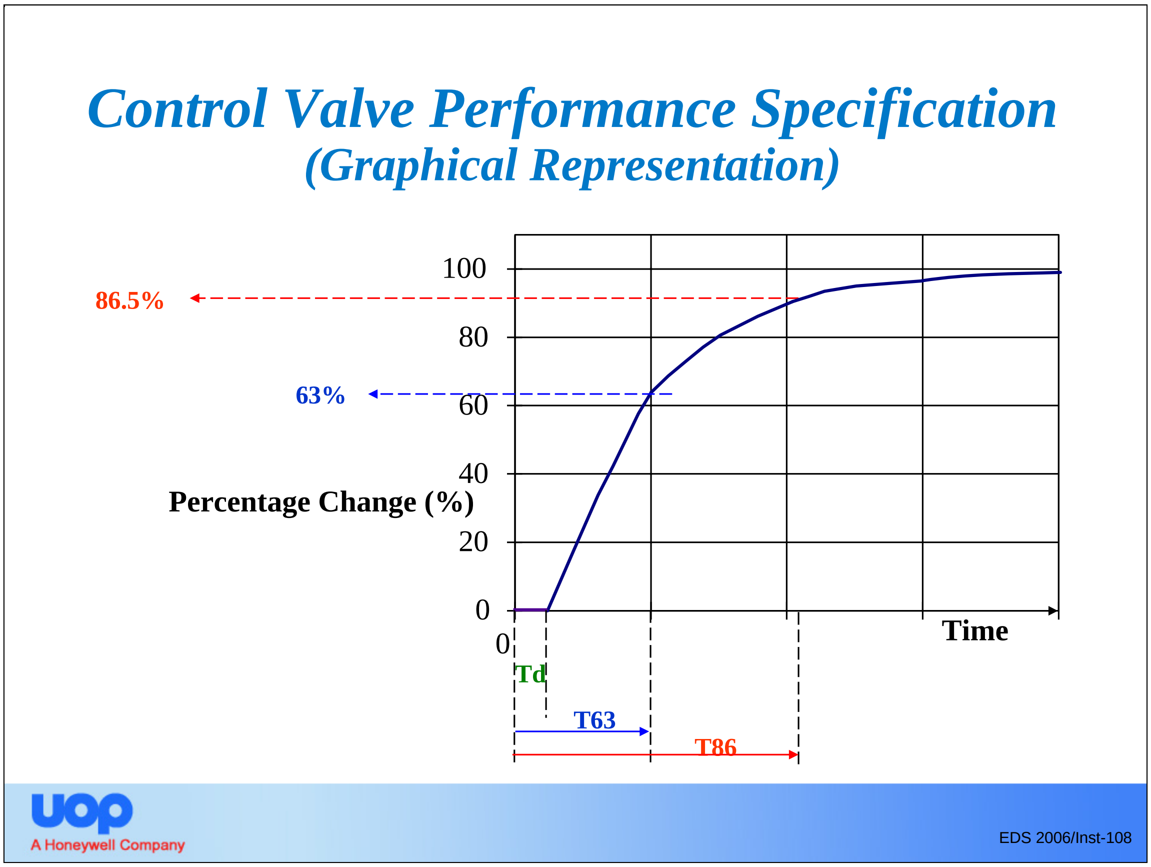
The Entech standard was developed for the paper mill industries and was developed by Entech engineering consultants to allow them to develop loop performance criteria for control loops in the paper industry. They found that loop controllers were very difficult to tune for optimum control because of poor performance in the control valves.

The performance criteria are geared for control valves in the paper industry.

The Entech standard may not be the ultimate standard for all control valves in the process industry, however this type of standard should be developed by each user of control valves to produce a specification that meets their requirements. As such the Entech standard is a good model to use.

It should be noted that this Entech standard cause many control valve vendors to redesign their valves to meet this standard.

UOP has formulated a standard for control valve performance and has used the Entech standard as a model and applied it to the Process Industry. These are some of the definitions and criteria that appear in UOP's standard.



The graphical representation depicts a first order response with dead time and graphically represents Td, T63, and T86.

T63 includes Td (deadtime).

T86 also includes Td.

Control Valve Performance Specification
(Speed of Response – Fast Loops)

Speed of Response:

Any step change in the range of 2 to 10% of full valve travel

Dead Band:

Less than 0.5% of full valve travel

Step Resolution:

Less than 0.25% of full valve travel

Overshoot:

Less than 10%

Speed of response for control valves in flow, differential pressure, and pressure control loops

Valve Size (inches)	**Td** (sec)	**T63** (sec)	**T86** (sec)
0 to 2	0.25	0.5	0.75
3 to 6	0.5	1.0	1.5
8 to12	0.75	1.5	2.25
14 to 20	1.0	2.0	3.0
22 to 24	1.25	2.5	3.75

UOP
A Honeywell Company

EDS 2006/Inst-109

Control valve step response and dead time criteria according to UOP control valve standard.

Using digital valve controllers with diagnostics, control valve step responses can be measured.

Control Valve Performance Specification
(Speed of Response – Slow Loops)

Speed of Response:

Any step change in the range of 2 to 10% of full valve travel

Dead Band:

Less than 0.5% of full valve travel

Step Resolution:

Less than 0.25% of full valve travel

Overshoot:

Less than 10%

Speed of response for control valves in temperature, level, hand, and analytical control loops

Valve Size (inches)	**Td** (sec)	**T63** (sec)	**T86** (sec)
0 to 2	0.5	1.0	1.5
3 to 6	1.0	2.0	3.0
8 to12	1.5	3.0	4.5
14 to 20	2.0	4.0	6.0
22 to 24	2.5	5.0	7.5

UOP
A Honeywell Company

EDS 2006/Inst-110

This is the same information by category for slower loops, such as temperature, level, etc.

Measuring Valve Performance

- **How well is the valve performing?**
- **Is the valve sized correctly?**
- **How well is the valve controlling the process?**
- **Does the valve need maintenance?**
- **Would a digital positioner with built-in diagnostics answer these questions?**

YES, YES, YES

UOP
A Honeywell Company

EDS 2006/Inst-111

It is for the reasons listed above that valve manufacturers have developed digital valve controllers.

These digital valve controllers overcome many of the problems inherent to pneumatic positioners.

Drawbacks With Pneumatic Positioners

- **Single stage positioners do not provide tight position control**
- **2-stage positioners improve position control but are more difficult to set-up and maintain**
- **Strap-on test equipment is required to perform valve diagnostic tests**
- **Difficult to measure accurately the valve travel**

A Honeywell Company

EDS 2006/Inst-112

It is for the reasons listed above that valve manufacturers have developed digital valve controllers.

These digital valve controllers overcome many of the problems inherent to pneumatic positioners.

Advantages of Digital Positioners

- **Improved valve position control**
- **Contains a microprocessor to perform valve control functions, communicate with a host device, and diagnostics**
- **Auto-calibration only takes a few minutes**
- **Monitor any process or valve status alarms, as well as valve position and input control signal while in service**

A Honeywell Company

EDS 2006/Inst-113

Digital valve controllers/positioners (DVC):

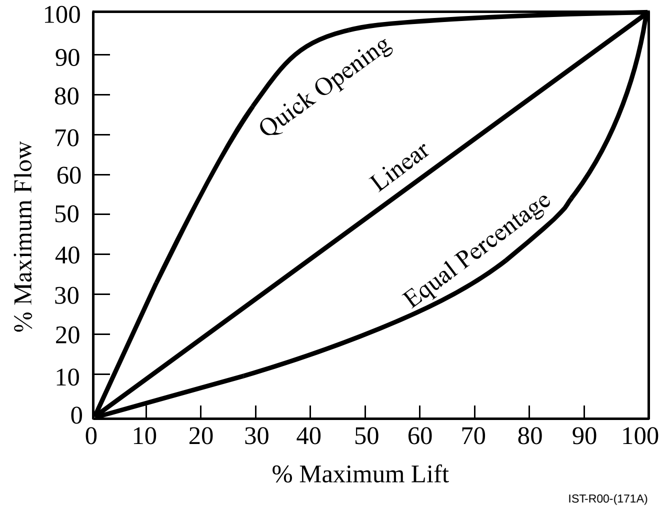
DVC's replace the conventional pneumatic positioner, usually can be retrofitted to replace existing pneumatic positioners, and contain a microprocessor to perform valve control functions, diagnostics and communication with a host device (PC, DCS, Hand held communicator).

The digital positioner provide better valve position control.

Auto calibration of a valve only takes a few minutes. The valve can be monitored in service to assess any process or valve status alarms, as well as monitor valve position and input control signal.

The advantages of the digital devices are endless.

Control Valve Characteristics



A Honeywell Company

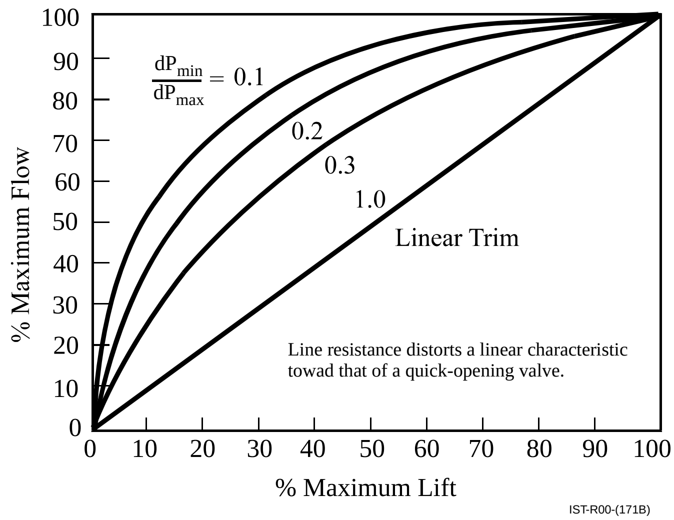
EDS 2006/Inst-114

By definition the inherent characteristic of a control valve is defined as the relationship between valve capacity and the closure member travel as the valve is moved from the closed position to rated travel at a constant pressure drop across the valve.

Bearing in mind valve flow is a function of both the valve travel and the pressure drop across the valve (as we shall see later), flow characteristic tests are performed at a constant pressure drop to provide a systematic means of comparing one valve characteristic design to another. The most common characteristics are linear, equal percentage, and quick opening.

One needs to evaluate the process hydraulics over its entire anticipated flow range in order to evaluate which trim would be best suited for the application. The key to the trim selection is based on the installed characteristic of the valve. Remember that the above characteristics are based on constant pressure drop across the valve. The installed characteristic may be distorted contingent upon the real valve pressure drop variation with respect to flow.

Installed Characteristic – Linear



uop
A Honeywell Company

EDS 2006/Inst-115

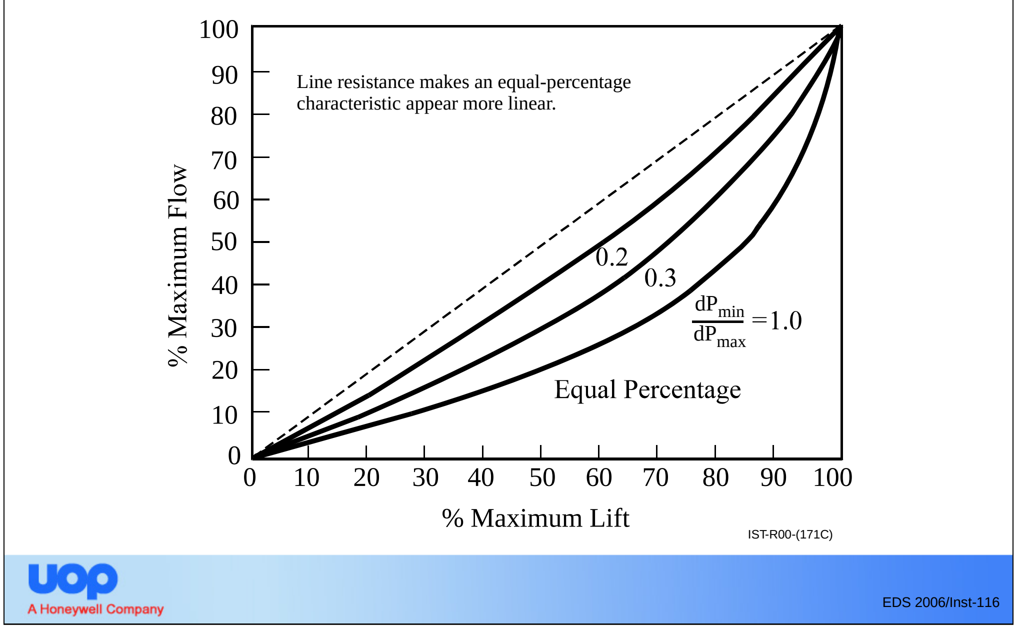
If at high flow rates the valve pressure drop is low relative to the valve pressure drop at low flows, then the ratio of dP_{min}/dP_{max} is less than 1. The smaller the ratio the more the inherent linear characteristic is distorted; and the installed characteristic will mimic that of a quick-opening style trim. Pipe and equipment resistance will distort a linear characteristic toward that of a quick-opening characteristic provided substantial differential pressure is consumed by the system.

Essentially if the differential pressure across the valve is relatively constant, ie

dP_{min}/dP_{max} is equal to 1, then a linear trim would have a linear install characteristic. This implies that the gain of the valve is constant. For a process in which the ratio is approximately 1, the process would be considered linear and a linear valve (a valve with constant gain) would be the appropriate choice.

Therefore examining this ratio as provided in the hydraulic analysis is a good basis for trim selection and we would want to choose a linear trim if this ratio is near 1.

Installed Characteristic – EQ%



Again if at high flow rates the valve pressure drop is low relative to the valve pressure drop at low flows, then the ratio of dP_{min}/dP_{max} is less than 1. The smaller the ratio the more the inherent equal percentage characteristic is also distorted; and the installed characteristic will mimic that of a linear style trim. Pipe and equipment resistance will distort an equal percentage characteristic toward that of a linear characteristic provided substantial differential pressure is consumed by the system.

Essentially if the dP_{min}/dP_{max} is relatively small, ie dP_{min}/dP_{max} is less than or equal to equal to 0.3, then an equal percentage trim would have a linear installed characteristic. This implies that the installed gain of the valve would be constant. For a process in which the ratio is small (<0.3), the process would be considered non-linear and a non-linear valve (a valve with a variable increasing gain) would be the appropriate choice.

Therefore examining this ratio as provided in the hydraulic analysis is a good basis for trim selection and we would want to choose an equal percentage trim if this ratio is much less than 1.

For processes where the ratio is say between 0.3 and 0.7, there is no obvious selection of one trim over the other. In these cases either trim selection will be adequate for the service.

Basic Flow Equation for Liquid Service

$Q = C_v \, (dP/G)^{0.5}$ where

Q = Flowrate, gpm at flowing temperature
G = Specific Gravity at flowing temperature
dP = Differential Pressure, psi

Minimum Design Pressure Drop

The valve minimum design pressure drop is the greatest of the following:

1) 50% of system friction drop, exclusive of the control valve, at the normal flow rate;
2) 10% of pump differential (the pump differential is the differential obtained if the control valve drop is set in accordance with item 1 above);
3) 25 psi.

A Honeywell Company

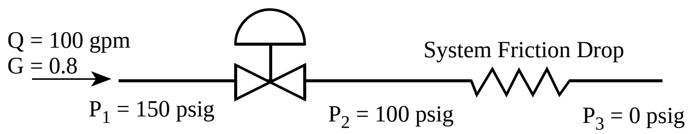
EDS 2006/Inst-117

The basic equation for liquid flow indicates that flow is proportional to the square root of the differential pressure drop across the valve. Cv is defined as the valve flow coefficient and by definition is " the number of gallons per minute of water which will pass through a given flow restriction with a pressure drop of 1 psi". Basically it is a capacity index with the valve in the wide open position, and allows the engineer to rapidly and accurately estimate the required size of a restriction in any fluid system. For a 1 inch valve the Cv = 12. Therefore if the dP is 1 psi, then 12 gpm would be flowing if the fluid was water at 60 °F. If the dP is 4 psi, then 24 gpm would be flowing.

UOP's normal practice is to calculate the required valve flow coefficient (required Cv) at the normal conditions and double the calculated Cv to obtain the "Approximate Valve Cv". Therefore at the normal flow condition, the valve should be at approximately 50% of its capacity. The minimum and maximum Cv's should also be estimated to ensure that the valve selected is adequate.

Rearranging the basic flow equation and solving for Cv, we can see that for any given flow rate the differential pressure across the valve is required. For a pumped system the rules stated above are based on a system where the design flow is 110% of the normal throughput. If the design flow is say 125 or higher these rules may require modifications.

Maximum Flow with 50% Frictional Drop for Valve



Basic Equations

$$C_V = Q\,(G/dP)^{0.5} \quad \text{and} \quad Q = k\,(dP)^{0.5}$$

Valve Sizing and Selection

$$C_V = Q\,(G/dP)^{0.5} \quad \text{Valve Cv} = 2 \times \text{Cv}$$

IST-R00-(172A)

A Honeywell Company

EDS 2006/Inst-118

The following example will help to illustrate the effects of line losses and the variable pressure drop across the control valve as flow rates change. Two basic equations can be used to define the liquid filled system. The first is the basic flow equation, rearranged solving for Cv. The second is the relationship of flow as a function of the pressure drop of the system.

This example shows 100 psi frictional drop in the system; therefore, allowing for a valve dP of 50 psi (rule 1) requires that the source pressure be equal to 150 psig. If the source pressure is constant, what is the maximum flow through the system when the valve is wide open? We are given two equations with two unknowns. The two unknowns are the maximum flow through the system (Q_{max}) and P_2 at Q_{max}.

Once the size of the valve is selected, the two equations can be set up with the two unknowns and solved simultaneously to determine what the maximum throughput through the system is with a constant source pressure.

Maximum Flow with 50% Frictional Drop for Valve Example

- **Valve Sizing, Selection & Maximum Flow**
  - **See Figure 40 following page 39 in Text Material**

uop
A Honeywell Company

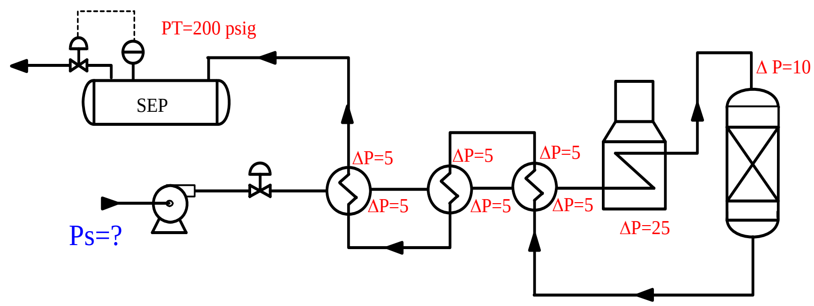
EDS 2006/Inst-119

The step by step solution is provided in Figure 40 following page 39 of the text material. The first step is to determine the calculated Cv at the normal flow rate of 100 gpm. The calculated Cv is 12.6. The Approximate Valve Cv is determined by doubling the calculated Cv. Therefore the Approximate Valve Cv is 25.2. This would lead to a valve selection of a 1-1/2 inch single seated globe valve with a rated Valve Cv equal to 25.

The next step is to use the basic flow equation with the Cv equal to 25. This first equation has two unknowns. With the Cv at 25 (valve wide-open) the flow then becomes Q_{max} (first unknown) and the downstream pressure P_2 is the second unknown. Likewise the second equation, the system pressure drop relationship, can be set up in terms of Q_{max} and P_2.

The two equations can be solved simultaneously. The solution yields Q_{max} equal to 115.3 gpm or a 15% increase in throughput. Even with the valve wide open the available pressure drop across the valve at normal conditions is taken up as frictional drop in the system, allowing for only an additional 15% increase in flow rate.

Reactor System Hydraulics



Piping:	**ΔP_{frict} = 35 psi**
Equipment:	**ΔP_{equip} = 65 psi**
@ flow = 100% (normal),	**ΔP_{total} = 100 psi**

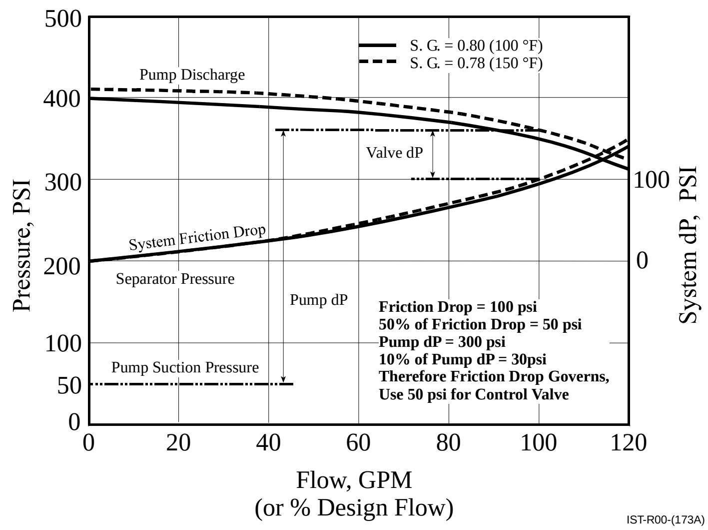
EDS 2006/Inst-120

Heat & Weight balance data provides the design basis for the typical reactor system. Equipment and Piping Specialists will design the equipment and pipe sizes along with providing pressure drop data for the system.

In the example above the separator pressure P_t (terminal pressure) is given as 200 psig. Based on the H&WB data the piping and equipment hydraulic pressure drop is estimated to be 100 psig at 100% (normal) flow. What is the pressure drop allotment across the control valve and what is the source pressure P_s?

Using the previously defined rules the pressure drop across the valve at normal flow is 50 psig. Therefore the discharge pressure at the pump will be 350 psig. This value will be used in selecting the appropriate pump for this service. But what happens to the system at different flows other than normal flow? UOP will typically design the system for a maximum flow of 110% of normal and a minimum flow of 60% of normal.

Low Pressure Pump, Motor Drive



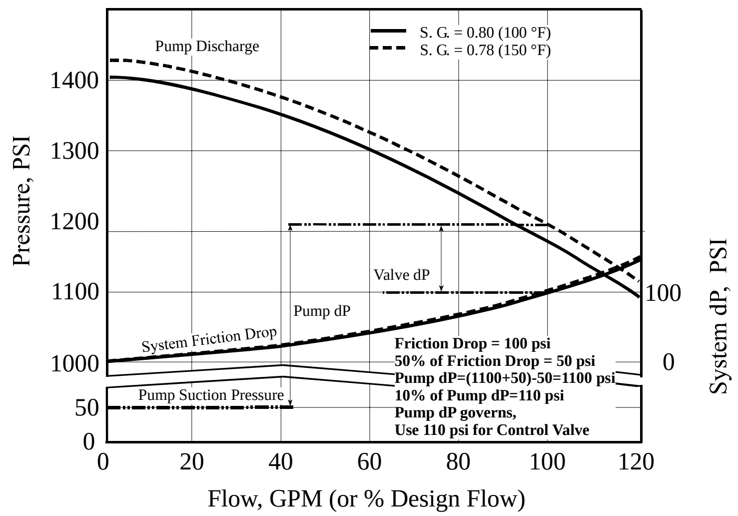
EDS 2006/Inst-121

A system pressure curve as a function of flow rate can be developed to illustrate the overall hydraulics of the system. The system friction drop curve can be generated using the basic friction drop relationship. We know that at 100% flow the system pressure drop is 100 psi. At 60% of normal flow the system pressure drop will be 36 psi and at 110% the system pressure drop will be 121 psi. Therefore the system friction drop curve can be generated on the curve.

We have also determined that at 100 % normal flow that the pump discharge will be 350 psig. Knowing the flowing gravity of the fluid, the pump head requirement can be calculated and a pump selected to provide this head. Once the pump head curve has been determined, then the pump discharge pressure can be generated and plotted on the system pressure curve at various flow rates.

The differential pressure between the pump discharge pressure curve and the system friction drop curve is the pressure differential across the control valve. Therefore the Cv requirement for design (110%) and turndown (60%) can be determined. This data then allows for determining the rangeability of the control valve along with valve sizing.

High Pressure Pump, Motor Drive
(Pump Head & Friction Drop Considered)



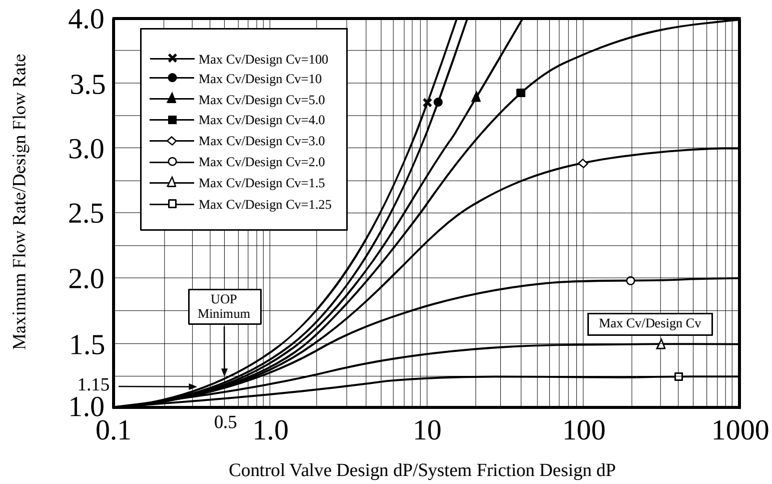
UOP A Honeywell Company

IST-R00-(175A)

EDS 2006/Inst-122

The last example is similar to the second example with the exception of the inclusion of rule 2 in determining the pump discharge curve. This positions the pinch point to the right of the 110% of design. Therefore the system is a workable solution. However notice that the pressure drop at the normal flow rate is higher (more energy waste is required at the normal flow) but this a consequence for meeting the design flow rate with a higher head pump.

Control Valve Hydraulics



A Honeywell Company

EDS 2006/Inst-123

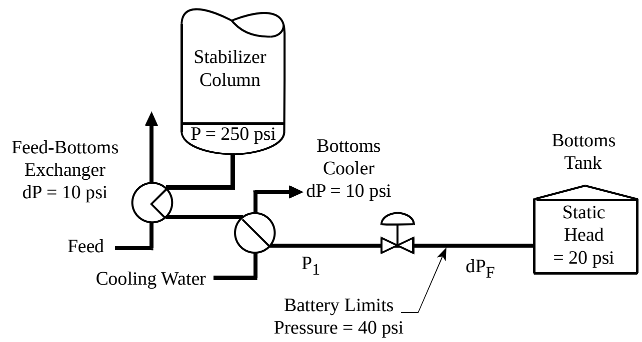
The graphical representation illustrates how the relationship between design valve differential and design friction drop affects the relationship between flow and valve Cv. This family of curves assumes constant supply pressure upstream and constant termination pressure downstream of the system.

As the ratio of design valve differential to design system drop increases (x-axis), the ratio of maximum to design flow (y- axis) approaches the ratio of maximum to design Cv. In other words as the majority of the pressure differential is available for the control valve with minimal pressure drop due to piping and equipment, Q is proportional to valve Cv. Therefore if the valve is 10 times larger then the flow is 10 times larger.

This also illustrates that over-sizing the control valve, will not necessary over-compensate for bad hydraulics. As an example if the design valve differential is only 20% of the frictional drop (instead of 50%), even a valve Cv 100 times larger than required will have a difficult time trying to meet a 110% design throughput.

Hydraulics – Stabilizer Bottoms

(Non-pumped circuit)



1. If Pipe dP Inside Battery Limits = 5 psi, P_1 = (250 - 10 - 10 - 5) = 225 psi
2. Valve dP = (225 - 40) = 185 psi
3. If Tank Static Head = 20 psi, dP_F = (40 - 20) = 20 psi
4. Total Friction Drop = (10 + 10 + 5 + 20) = 45 psi
5. Valve dP / Total Friction Drop = 185 / 45 = 4.1

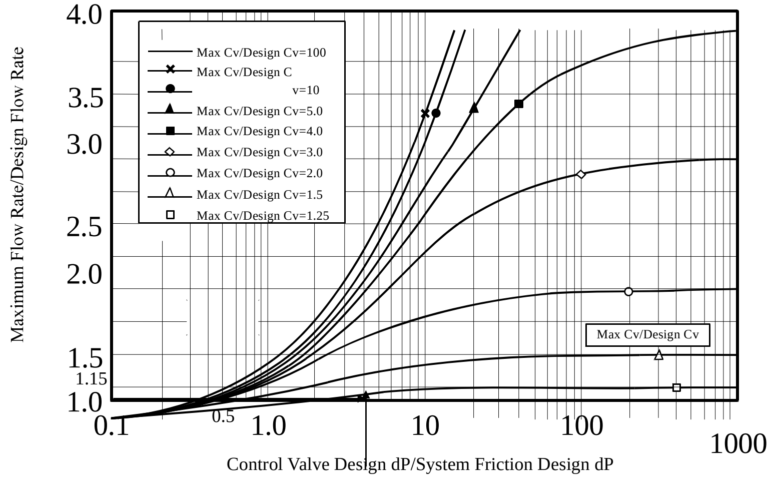
A Honeywell Company

EDS 2006/Inst-124

For a non-pumped circuit, the system must be analyzed for pressure drop considerations. UOP will typically use the doubling of the calculated Cv to determine the required valve Cv, but other methods could be applied as well.

In the above system, the Valve dP/Total Friction Drop was calculated to be 4.1. The previous graph can be used to determine what size valve is required to obtain a 115% increase in throughput based on this ratio equal to 4.1

Control Valve Hydraulics



EDS 2006/Inst-125

In order to be able to obtain a 115% increase in throughput with the ratio of 4.1 (x-axis) the Max Cv/Design Cv of the valve is approximately 1.25. Therefore if the Design Cv (calculated Cv at normal flow) is 8, a 1 inch valve with a valve Cv equal to 12 would suffice for this system.

Control Valve Sizing Equations

- **Control Valve Sizing Handbook**
  - **Liquid Flow Equations**
    - **Non-vaporizing (sub-critical flow)**
    - **Vaporizing (critical flow)**
    - **Non-turbulent**
  - **Gas and Vapor Flow Equations**

uop A Honeywell Company

EDS 2006/Inst-126

Masoneilan's control valve sizing Handbook has been provided. The various equations and terminology is provided for Liquid/Gas applications along with the sub-critical/critical/choked equations.

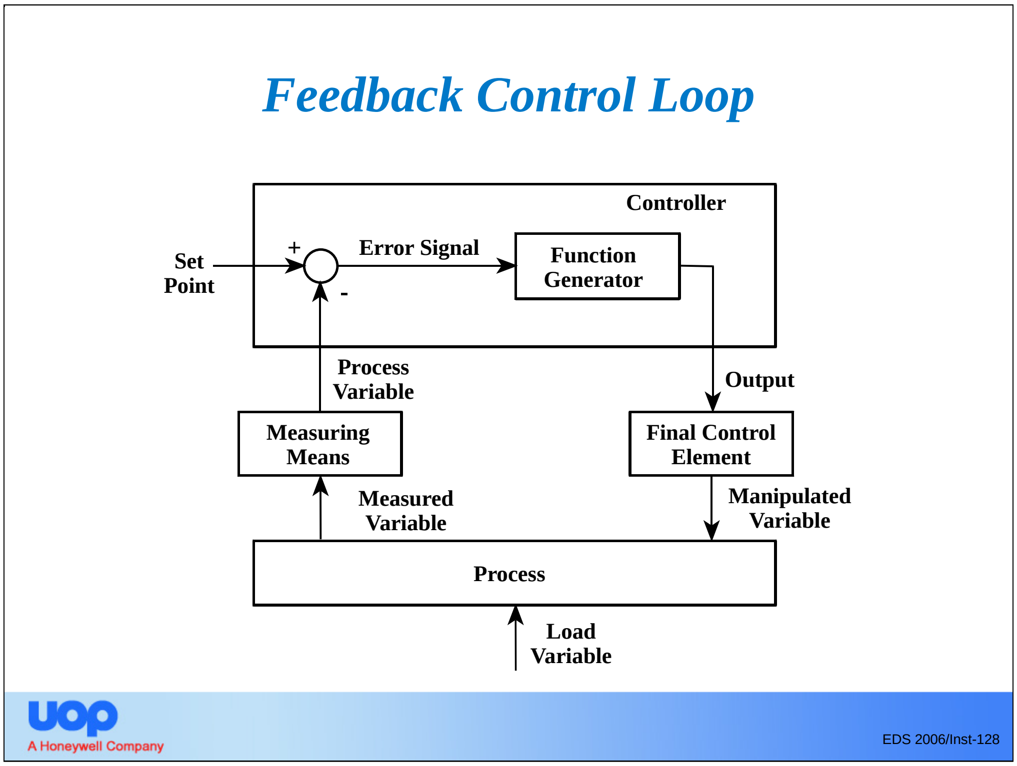
The equation that we investigated earlier was the sub-critical liquid equation. The critical liquid equation introduces an F_L factor, the liquid pressure recovery term and a modified pressure differential limit based on fluid vapor pressure and critical pressure. The factor is a function of valve geometry. Like the orifice plate, the fluid accelerates past the seat and forms a vena contracta (area of highest velocity and lowest pressure). As the fluid decelerates back to normal velocity some of the pressure drop is recovered. A control valve with a smooth shaped outlet (rotary and butterfly valves) show a higher percentage of pressure recovery than a valve which has more changes in flow direction (globe valves).

While pressure recovery in a flow measurement device is beneficial, pressure recovery in a control valve is not. A high pressure recovery corresponds to a low pressure recovery factor, which increases the possibility of critical flow. If the pressure anywhere in the system falls below the vapor pressure of the fluid, then vaporization will occur. If during pressure recovery the pressure rises above the vapor pressure of the fluid, then the vapors collapse back into the liquid phase. This phenomenon is known as cavitation. Flashing occurs if after pressure recovery, the static pressure is still below the fluid vapor pressure.

Feedback Control Loop

- **Loop Components**
  - **Final Control Element**
    - **Control Valve Body Types**
    - **Control Valve Performance and Accessories**
    - **Trim Characteristics**
    - **Hydraulics and Sizing Equations**
  - **Controller**
    - **P, PI, and PID Modes**
    - **Applications**
    - **Tuning Methods**

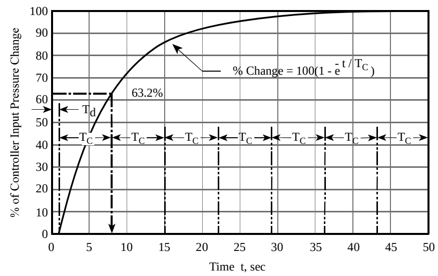
This concludes the section on the final control element. The next topic and last part of the feedback control loop is the controller itself.



In terms of the block diagram the controller has two inputs and 1 output. The two inputs are the Process Variable and Set Point. The output of the controller is connected to the final control element and modulates the Manipulated Variable. The Feedback Controller generates an error signal (difference between Set Point and Process Variable) and adjust the controller output as defined by the PID algorithm.

We have discussed dynamic performance of both the transmitter and final control element, but have not discussed the "transmission means" of the Process Variable or Controller Output. The means of transmission would not matter if it did not affect the loop performance. Historically the means of transmission was initially a 3 to 15 psig pneumatic signal representing 0 to 100% of the signal range. The pneumatic world, as we shall see, was slow and limited process control to the field for most control loops. We the advent of electronic controllers and transmitters, the transmission means was an instantaneous 4 - 20 ma electrical signal representing 0 to 100% of the signal range. This lead to the concept of centralized control rooms. In today's digital environment, the transmission means is still fast, but is not continuous as we say with the update rates in the transmitters. This will have an overall affect on loop performance if the process variability is as fast or faster than the transmitter update rate.

Pneumatic Tubing Time Response
First Order Lag plus Dead Time
(1000 feet of 1/4 inch tubing)



Time t, sec	Time Constant	Change, % Full Scale	Controller Input, % Full Scale	% Full Scale from Final
0	0	0	0	100.0
1	0	0	0	100.0
8	1	63.2	63.2	36.8
15	2	23.3	86.5	13.5
22	3	8.6	95.0	5.0
29	4	3.1	98.2	1.8
36	5	1.2	99.3	0.7
43	6	0.4	99.8	0.2
50	7	0.1	99.9	0.1

A Honeywell Company

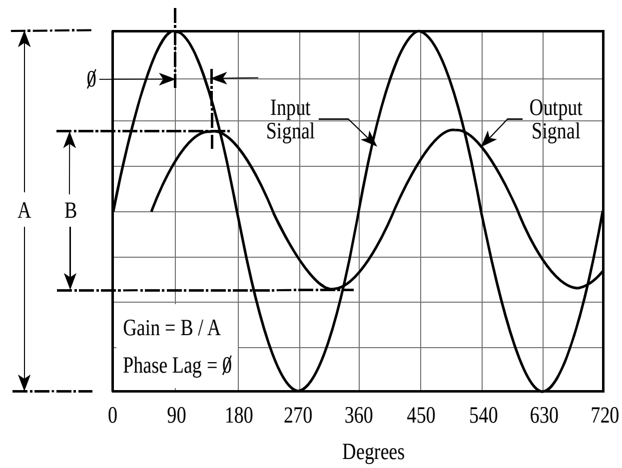
EDS 2006/Inst-129

Pneumatic systems are not commonplace in today's environment, but a brief discussion on this system is useful to demonstrate the principles behind control theory. In a compressible medium (instrument air), the speed of a pressure change (transmitter or controller output) is limited to the speed of sound (≈1100 feet per second). If the pneumatic transmitter is located 1000 feet from the controller, nearly 1 second would elapse before the pressure at the controller would even start to change. In addition each unit length of air tubing has a capacitance (a given volume) and resistance to flow. The net result is that once the pressure at the controller does start to change, it changes in an exponential manner (first order lag with dead time).

The above data illustrates a step change in a transmitter's output and the measured pressure at the inlet to the controller as a function of time. The plot of the data illustrates the transient response of a single time constant process plus dead time. The dead time is the time from the initial step change in transmitter output until the controller input signal first start to change. The time constant of the system is the time required from the initial change in controller input signal until it reaches 63.2% of the total change. These two times, dead time and time constant, are important in the determination of system response and stability.

For the 1000 feet of tubing the dead time is shown to be 1 second and the time constant for the system is 7 seconds.

Pneumatic Tubing With Sinusoidal Input



uop A Honeywell Company

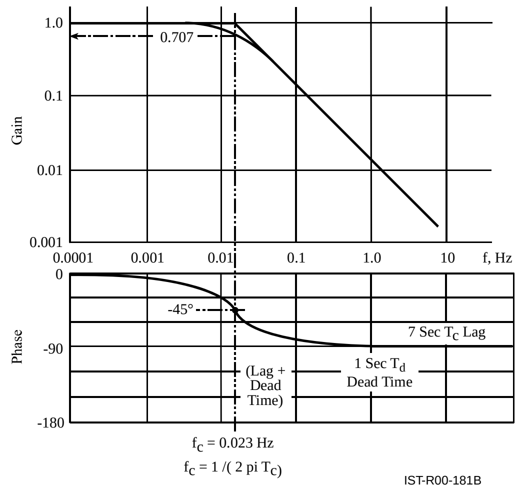
EDS 2006/Inst-130

Response can also be measured in terms of frequency. To determine the frequency response of any component, a sinusoidal signal of varying frequency is applied to the input side of the component. On the output side of the component, the magnitude and phase shift of the output signal are measured for each input frequency.

At low frequency input signals the resulting magnitude of the output signal is typically equal to the magnitude of the input signal and the phase lag is 0 degrees, i.e. the output signal faithfully reproduces the input signal. The component or system being tested can be pneumatic, electrical, mechanical, or even the “Process”. However as the input frequency increases, it becomes more difficult for the system or component to faithfully reproduce the input in terms of the output signal. Depending upon the system or component in question, eventually the magnitude of the output signal begins to diminish and a phase lag less than 0 degrees develops.

Graphically the ratio of the magnitudes is plotted as a function of the input frequency on a log-log scale, and the phase lag is plotted as a function of the input frequency on a semi-log scale. These characteristic curves are commonly known as a Bode Plot.

Pneumatic Tubing Frequency Response
(1000 feet of 1/4 inch tubing)



A Honeywell Company

EDS 2006/Inst-131

The logarithmic plot of gain vs. frequency is advantageous from the standpoint that the product of individual component gains in a loop is equal to the overall loop gain, i.e, G_L=G_1xG_2x… and can be quickly multiplied to obtain the overall gain at each frequency.

If two straight lines are drawn asymptotic to the gain curve, the intersection of the two segments occurs at the “corner frequency”. The frequency at this point is directly related to the time constant found in the step response previously discussed. The corner frequency in radians per unit time is equal to the reciprocal of the time constant:

$$1/Tc = 2\pi f$$

where:

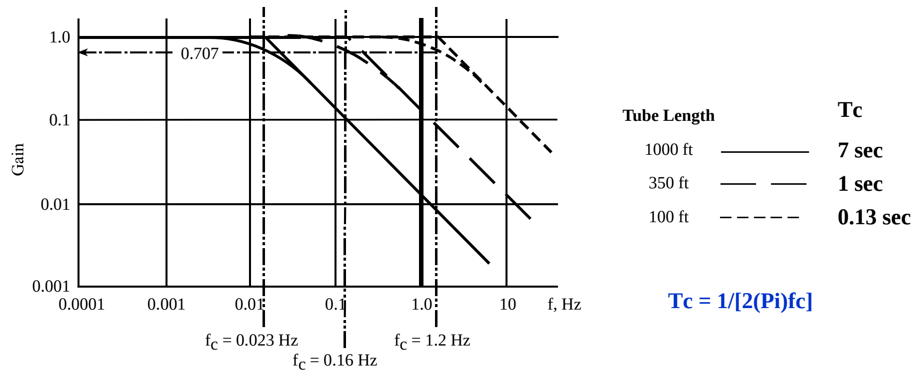
Tc = time constant

f = corner frequency, Hz

For the system in question the corner frequency was determined to be 0.023 hertz. Solving for Tc yields a time constant for the system equal to 7 seconds.

Pneumatic Tubing Frequency Response

(Variable Tubing Length)



EDS 2006/Inst-132

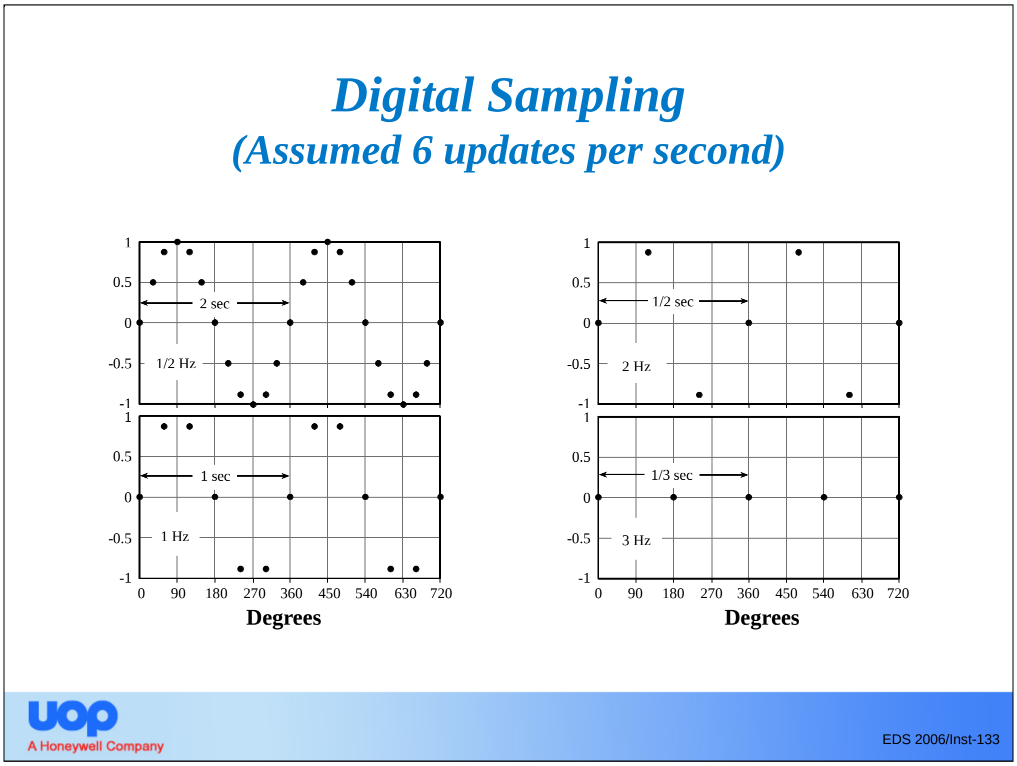
For the 1000 feet of pneumatic tubing the corner frequency is found and the resultant Tc is determined to be approximately 7 seconds. In terms of the corner frequency (0.023 Hertz) this is much to slow a response for good control. UOP requires a frequency response of at least 1.0 Hertz.

Shortening the pneumatic tubing to 350 feet results in a Tc = 1 second and a frequency response equal to 0.16 Hertz. In order to meet the 1 Hertz requirement for good control the pneumatic tubing run has to be limited to something less than 150 feet. The frequency response for 100 feet is shown to be 1.2 Hertz.

This is the main reason why many pneumatic systems for flow and pressure control loops are mounted in the field as close to the control valve and transmitter as possible

In terms of electrical devices, electronic analog transmitters and controllers and temperature devices the frequency response is essentially instantaneous.

The advent of the electrical devices gave way to centralized control rooms.



The use of electronic analog transmitters have given way to microprocessor base instruments. Commonly known as smart transmitters, the 4 - 20 ma signal is continuously updated at discrete intervals.

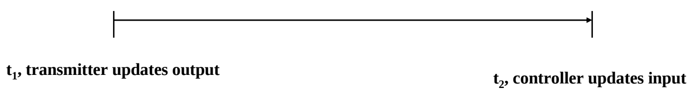
If the process variable is oscillating in a sine wave fashion and the update rate is high compared to the frequency of oscillation, then the updated signal yields a fairly representative picture of the process variable (1/2 and 1 Hz).

Again as the frequency of the process variable increases and approaches the the update rate of the transmitter, an increasing distorted view of the process variable results (2 and 3 Hz).

For most fast responding processes, transmitters with slow update rates degrade the control system and in some instances are completely unacceptable (compressor anti-surge control).

Control Algorithm Performance

Control Algorithm Performance		
Control Service	**Update Rate**	**Execution Rate**
Flow	4 updates per second	250 milliseconds
Liquid Pressure	4 updates per second	250 milliseconds
Gas Pressure	2 updates per second	500 milliseconds
Differential Pressure	4 updates per second	250 milliseconds
Temperature	2 updates per second	500 milliseconds
Level	1 update per second	1 second
All Others (Analysis, pH, etc)	2 updates per second	500 milliseconds



A Honeywell Company

EDS 2006/Inst-134

In terms of the execution rate for the distributed control system, a similar phenomenon exits with the input to the controller. If the transmitter updates at t_1 and the controller updates at t_2 before the transmitter updates again, then the difference in time (dead time) determines how "old" the process variable is, and in the worse case if completely out of sink, the controller is making adjustments on the oldest data. If the update rate of the transmitter is doubled, then the time difference (dead time), is cut in half. Therefore the update rate of the transmitter needs to be somewhat faster than the controller execution rate. UOP recommends that the update rate of the transmitter be twice as fast as the update rate of the distributed control system. In general for control systems which meet these guidelines, the signal transmission and control will not be the limiting factor in control performance of the process.

The above table is based on experience with various distributed control systems available in industry today. Control algorithm execution time is used as a measure of performance. Flow and liquid pressure loops are fast acting processes and require the faster execution times. Lower transmitter update rates and higher controller execution times increase the control loop dead time, the delay before the output reflects a change in the input.

Proportional Only Controller

$$O_c = G\,e + B$$

where:

O_c = Controller Output
G = Gain
e = Error
B = Bias (Manual Reset)

$$e = (PV\text{-}SP)$$

where:

PV = Process Variable
SP = Set Point

Gain = change in Output/change in Input
conversely, G may be represented in terms of Proportional Band (PB%)
PB% = 100% / Gain

uop
A Honeywell Company

EDS 2006/Inst-135

The basic feedback controller has three modes of control: PID. Proportional (P), Integral (I), and Derivative (D) action operates on the error generated in the controller and adjusts the output to the Final Control Element accordingly. Error is defined as the difference between the Process Variable and Set Point. Any combination of PID can be implemented in most controllers.

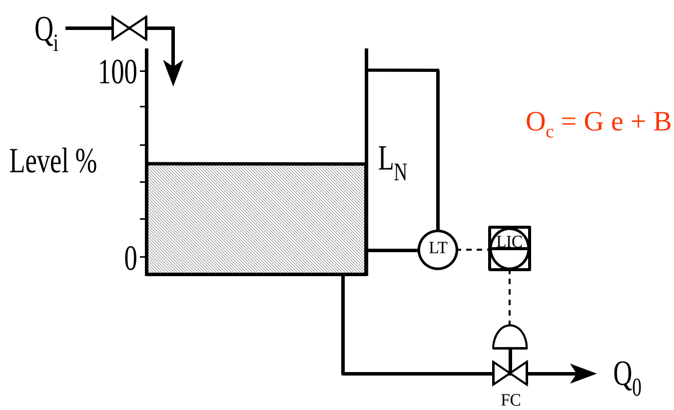
We shall investigate the response of each individually. The proportional only controller is the simplest mode and is defined above:

<u>controller output is equal to the controller gain times error plus bias</u>

The equation which represents the proportional only controller is an equation for a straight line. The gain of the controller is the slope of the straight line and the bias term is the y-axis intercept. Gain then is defined as change in output per change in input. With the input being the error signal, if the error is 1% and the gain is 1, then the output will change 1%. If the gain is 2, then the output will change 2% for a 1% error. Essentially the output will change proportionally to the error being generated by the process.

At UOP, the proportional only control mode is used most often for level measurement. Surge drums are primary examples where exact level control is not mandatory in the operation of the unit.

The Level Process

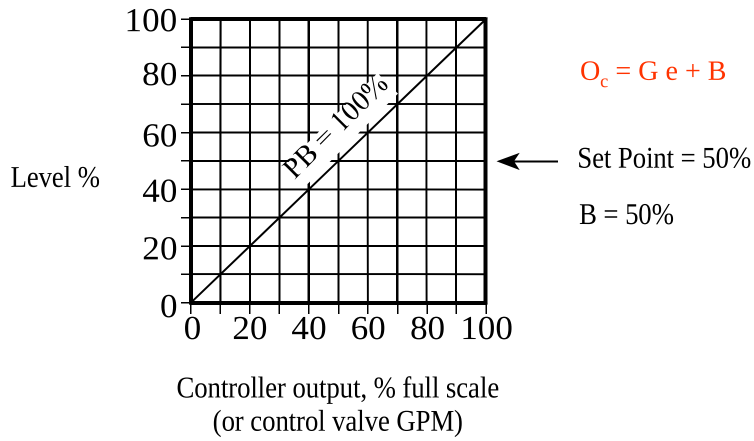


In examining the proportional only controller per UOP's typical design criteria, lets say the normal flow rate to the vessel is 50 gpm. If a valve is selected with exactly twice the capacity as the normal rate, the controller output can be viewed as a flow rate in gpm.

During operation of the P only controller the operator will typically line-out the process in manual. If the inflow is 50 gpm, then the operator would manually adjust the output to the valve until the outflow is equal to the inflow. Let's say this occurs at a level measurement of 50%. This matches the engineering design of the system with the level at 50% flow at design and the control valve at half capacity.

The operator switches to automatic control. PV is equal to SP, and no error exists. The output of the controller remains at 50% until something disturbs the process. For our equation of controller output, this scenario sets the bias of the controller to 50%, independent of the gain of the controller.

The Level Controller
(Proportional Only)



A Honeywell Company

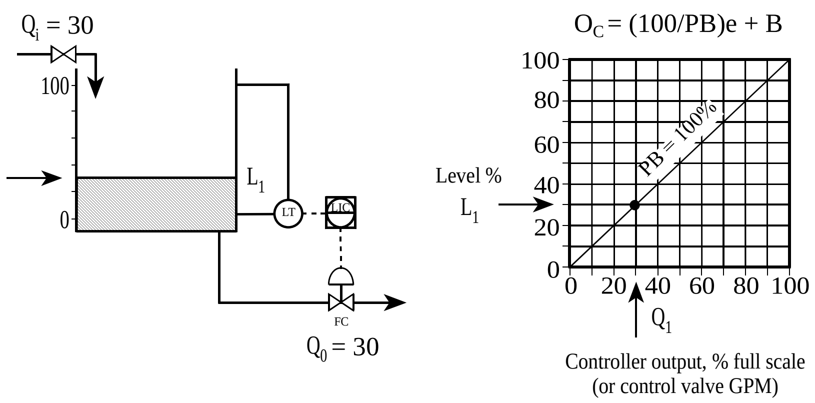
EDS 2006/Inst-137

In terms of the controller at the point of switching to automatic control, the PV and SP are equal and e = zero. The Bias of the controller is set equal to the output of the controller. This is commonly called “bumpless” transfer.

If the gain of the controller is 1 (100% PB), then the equation is represented as straight line with a slope equal to 1 passing through the 50% points.

If the inflow decreases such that the level moves to 30%, then an error would exist: e = 30 - 50 = -20. The output of the controller would be equal to 30% or in terms of flow 30 gpm. The system for p-only control eventually matches input flow to output flow, but will always have an offset between PV and SP.

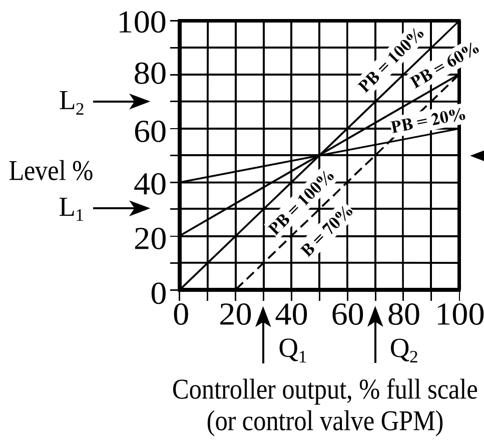
The Level Controller
(Proportional Only)



EDS 2006/Inst-138

Assuming that the controller was placed in automatic with a 50% set point when the level was at 50% sets the B to 50%, fixes the controller equation. If the inflow changes from 50 to 30 gpm, then the error generated is -20% and the controller output is 30% or 30 gpm and outflow matches inflow at steady state. What happens to the system now if we have a step change in inflow from say 30 gpm to 70 gpm?

The Level Controller
(Variable Gain & Bias)



Offset = PV -SP = (PB/100)(Oc-B)

Two ways to reduce offset
1. **Increase Gain**

or

2. **Increase Bias**

A Honeywell Company

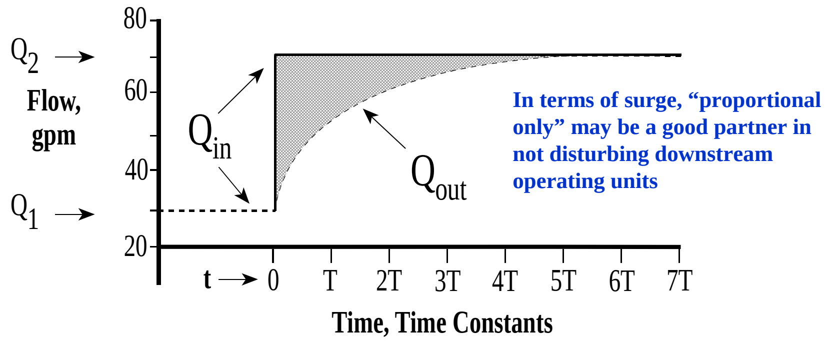
EDS 2006/Inst-139

Steady state error , the deviation between the lined-out value of the process variable and the set point, is also known as offset. This offset can be reduced by increasing the gain of the controller (smaller PB). Theoretically the gain could be increased to infinity and thereby decrease the offset to zero. However increasing the gain will eventually lead to oscillation and control loop instability.

Rearranging the controller algorithm in terms of the steady state error or offset, indicates that the offset could be eliminated by adjusting the manual bias B equal to the output of the controller. As indicated this would shift the control line to the right for the system where the inflow had increased to 70 gpm. The problem with this solution is at low inflows the tank level will go to zero and the valve will remain at 20% open.

Proportional control is often selected for level control when inventory control is not a major concern for process control. In other words the vessel will be used as a surge drum minimizing upsets to downstream operations.

The Level Controller
(In-Flow Step Change)



uop A Honeywell Company

EDS 2006/Inst-140

For the example of changing from an inflow of 30 to 70 gpm, the inflow is represented by a step change at time "zero". Depending upon the gain of the controller as the level increases, the output of the controller begins to increase and as a result the outflow from the vessel increases. The process follows a first order lag response and is represented by the Qout curve. The area between the two curves, Qin and Qout, represents the accumulated level in the tank.

If the gain could be adjusted to have the offset equal to zero then the area under the curve would be zero and as soon as Qin changes Qout would change instantaneously.

In terms of downstream effects on operation, the outflow is gradually increased and does not "slug" the changes in inflow to the downstream equipment.

Proportional Plus Integral Controller

$$O_c = G[e + (1/T_i) \int_0^t e\,dt]$$

where:

O_c = Controller Output
G = Gain
e = Error
T_i = Integral Time, minutes per repeat

$$e = (PV-SP)$$

where:

PV = Process Variable
SP = Set Point

Integral Time, (T_i) = minutes per repeat
conversely, T_i may be represented in terms of Reset (I), repeats per minute

$$I = 1 / T_i$$

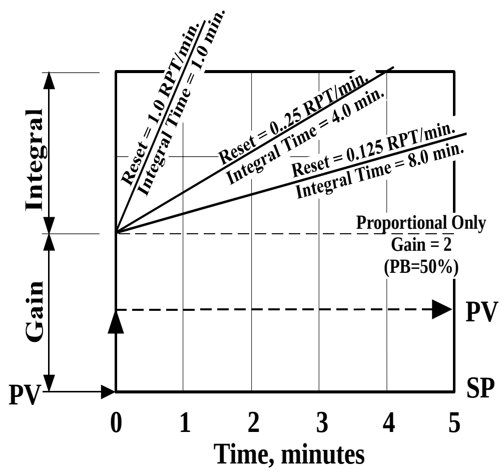
A Honeywell Company

EDS 2006/Inst-141

With the proportional only controller, we discovered earlier that we could reduce the offset by manually readjusting the bias. By replacing this manual bias with the integral term (I in PID) the form of the equation is similar except the manual reset is replaced by the integral action and automatically moves the controller output in the direction that minimizes the offset. The integral time is defined as minutes per repeat. One minute per repeat means that in each one minute interval, the Integral action of the controller will cause a change in controller output that is equal to the change in output that results from the Proportional action alone.

As the integral time gets larger and larger the reset action diminishes. If in fact we set the integral time to infinity, the reset action of the control is essentially zero, i. e.. proportional only control.

Open Loop Integral Response (Gain = 2.0)



A Honeywell Company

EDS 2006/Inst-142

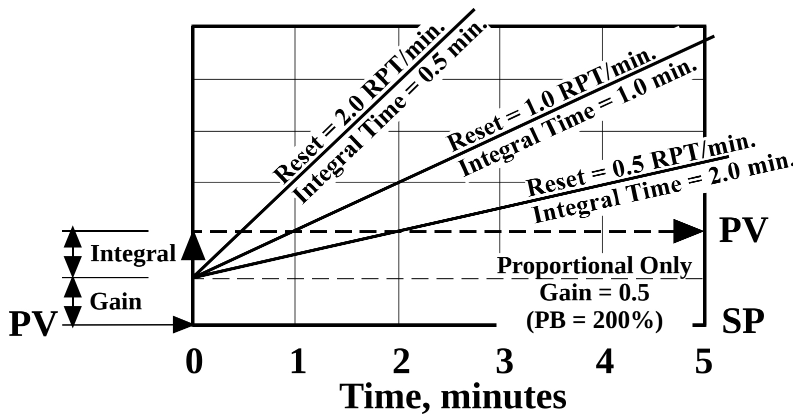
What is the effective output of a PI controller? An open loop response will give an indication of the integral action contribution. With the process lined-out and the controller in manual (PV, SP, and OP at steady state), a load change occurs at t=0. Assuming that the load change causes a step change in PV equivalent to 1 unit and the GAIN of the controller is 2 (proportional action), the proportional action of the controller will change the output of the controller by 2 units at t=0. With the integral time set to infinity, the output of the controller would remain constant for t greater than 0. Notice that as time progresses the model shows both the PV and SP are constant; therefore the error signal in the controller is constant.

If the integral time is set to 1 minute/repeat, then by the definition of integral action the controller output will change an additional 2 units every minute. If the integral time is set to 4 minutes/repeat, the controller output will change an additional 2 units every 4 minutes.

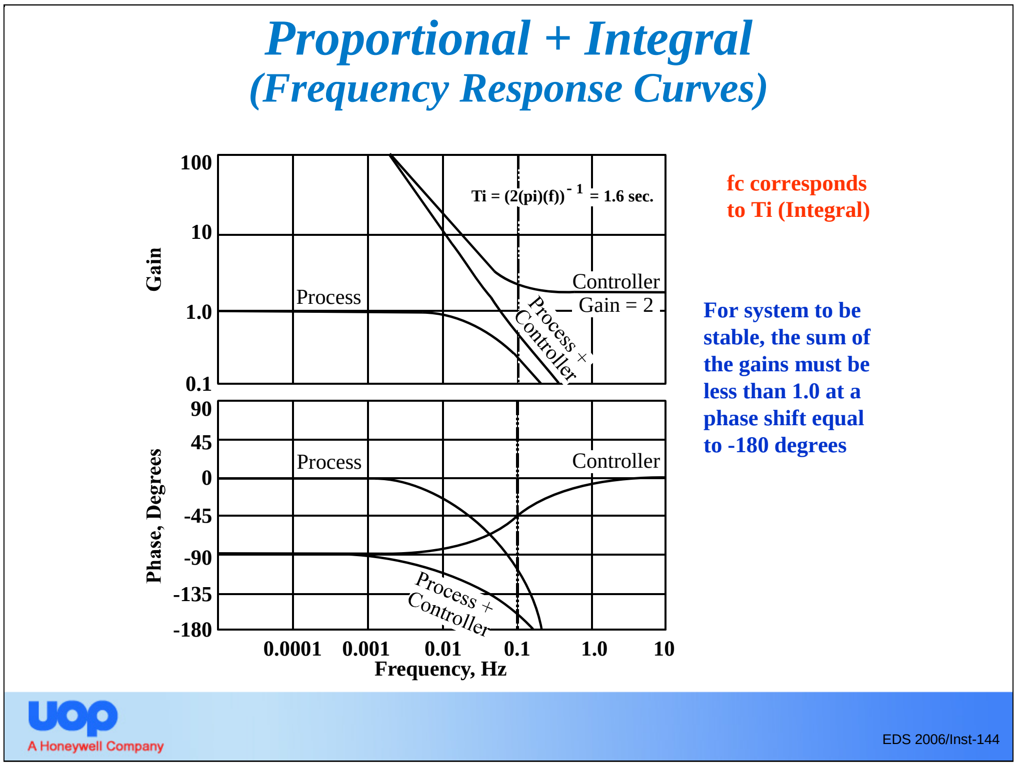
The Integral contribution as shown in the equation is the time integral of the error function of the controller. Therefore when the controller is in automatic and the controller output is throttling the manipulated variable, the error signal should be diminishing. How fast this occurs depends upon the controller settings for the PI controller, that is what is the gain setting of the controller and what is the Integral setting of the controller. There is an infinite number of combinations: some good and some bad.

Open Loop Integral Response

(Gain = 0.5)



This is another example of the open loop response of a PI controller. The gain of the controller is set at 0.5. Again for a 1 unit load disturbance in the PV, the output of the controller is immediately change by 0.5 units due to the proportional action of the controller. Depending upon the Integral time, the controller output integrates the error signal with respect to time.



Similar to the pneumatic tubing examples shown previously, we can investigate the frequency response for the PI controller. Individual components of the loop can be shown on the frequency response curve. The unique feature of the frequency response curve is that individual components of the loop are additive on the log scale. The process is shown as a single time constant process similar to the pneumatic tubing case investigated previously. (Flow loops and pressure loops are good examples.) The open loop response for the controller is shown with the Proportional action set at GAIN=2 and Integral action set at T_i=1.6 seconds/repeat. For high frequencies the proportional action of the controller is the dominant action. For low frequencies the Integral action of the controller will have significant affect on the controller output. In terms of the "corner frequency", the Integral time is equal to the corner frequency. Therefore by adjusting either the Integral time, the Gain, or both, the controller and sum of the controller plus process will be altered. However experimentally this is difficult to collect this type of data for any real process.

When the combined open loop response is plotted, Process plus Controller, the stability of the loop can be checked. For the system to be stable, the sum of the gains must be less than 1 at a phase shift equal to -180 degrees. There are many settings that will provide stable control and many that are unstable.

P + I + Derivative Controller

$$O_c = G[e + (1/T_i) \int_0^t e\,dt + T_d(de/dt)]$$

where:

O_c = Controller Output
G = Gain e = Error
T_i = Integral Time, minutes per repeat
T_d = Derivative Time, minutes

$$e = (PV-SP)$$

where:

PV = Process Variable
SP = Set Point

Derivative Time, (T_d) = minutes
The derivative contribution is proportional to the
Rate of Change of the error signal

A Honeywell Company

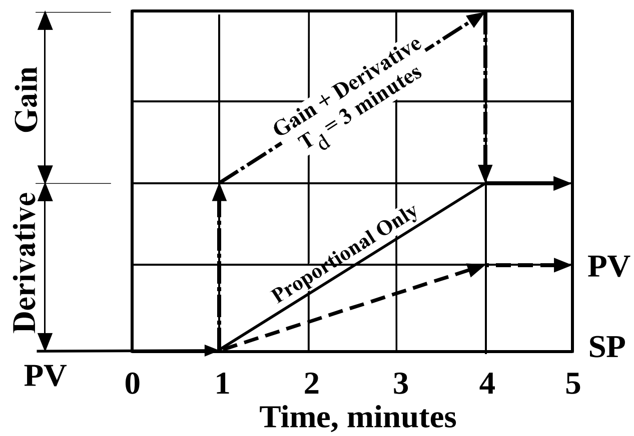
EDS 2006/Inst-145

PI control is typical for flow and pressure loops. These are fast-acting loops which have little, if any, lag in the process and small time constants. The amount of energy stored in the process is small. For a liquid flow loop as soon as the valve moves, the flow measurement changes almost instantaneously. For vapor flow or pressure loops, a small amount of energy is stored in the process due to the compressibility of the vapor. However the process is still first order with a slightly larger time constant.

For systems with a large amount of stored energy, such as temperature loops, lag in the process is significant. This type of process can not be modeled with the simplified first order process with a single time constant. Temperature loops also inherently possess a significant amount of dead time. A decrease in heater outlet temperature, causes an increase in burner pressure increasing the heat release. This must first heat up the tubes, then the process, and eventually get to the temperature measuring device on the heater outlet. This process has significant lag and dead time.

To be effective and catch up with the process change, the controller needs a high gain at the higher frequencies where the changes occur the fastest. This is accomplished with Derivative action. Derivative action takes into account how fast the error is changing and is proportional to the rate of change of the error signal with respect to time. Setting T_d to zero, defaults to PI control.

Open Loop Derivative Response
(Gain = 2.0)

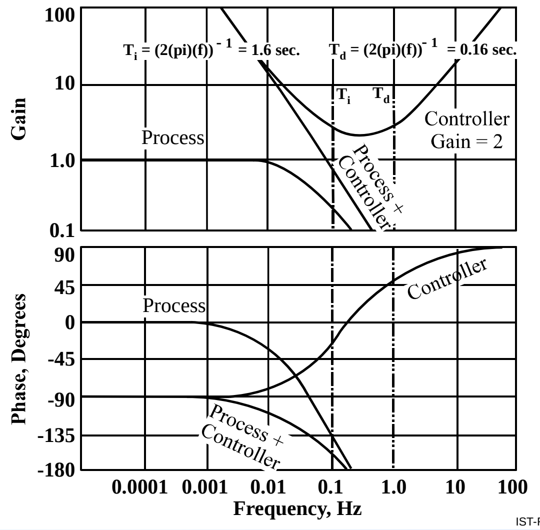


uop
A Honeywell Company

EDS 2006/Inst-146

What is the effective output of a PD controller? An open loop response will give an indication of the derivative action contribution. With the process lined-out and the controller in manual (PV, SP, and OP at steady state), a load change occurs at t=1. Unlike the open loop response for the integral action (a step change in PV was initiated), the load change causes a constant rate change in PV equivalent to 0.3333 units per minute. Assuming the GAIN of the controller is 2 (proportional action), the proportional action of the controller will change the output of the controller proportional to the rate of change of the PV. If T_d is set to zero, that is proportional only control, the controller output would ramp up twice as fast at the change in PV. With the rate of change at 0.3333 units per minute, Td set at 3 minutes, and GAIN set at 2, the Derivative action of the controller is equal to the product of these three settings (0.3333 x 3 x 2 = 2.0 units). Therefore as long as the rate of change is constant, the Derivative action is constant at 2 units. With the proportional action set with a gain of 2 this must be added to the derivative response, which is accumulative because of the change in error signal. At t=1 the load change appears and at t=4 the load change disappears with the error being constant at one unit. At t=4 the Derivative contribution is zero because the rate of change of error with time is zero. However there is a constant error of 1 unit and the proportional action remains with a change in controller output equivalent to 2 units.

Proportional + Integral + Derivative
(Frequency Response Curves)



IST-R00-186A

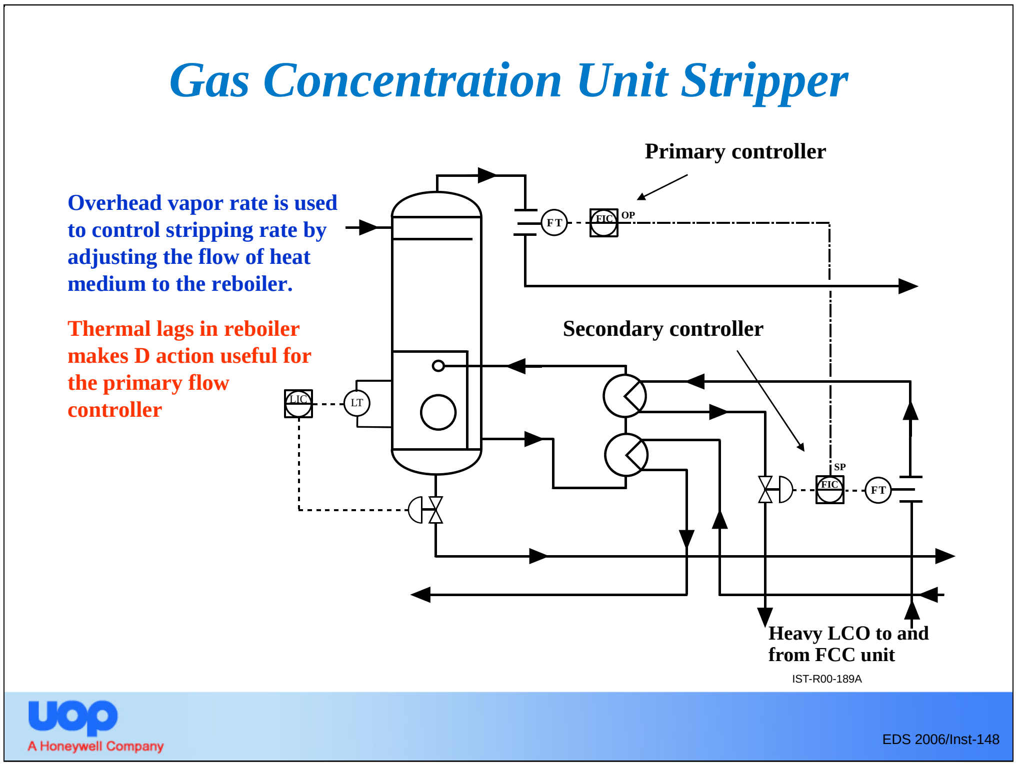
For system to be stable, the sum of the gains must be less than 1.0 at a phase shift equal to -180 degrees

A Honeywell Company

EDS 2006/Inst-147

Similar to the PI frequency response shown previously, we can investigate the frequency response for the PID controller. Again the process is shown as a single time constant process. The open loop response for the controller is shown with the Proportional action set at GAIN=2, Integral action set at T_i=1.6 seconds/repeat, and now T_d set at 0.16 seconds. For high frequencies the derivative action of the controller is the dominant action, and as before for low frequencies the Integral action of the controller will have significant affect on the controller output. In terms of the “corner frequency”, the Integral time is equal to the corner frequency for that portion of the controller response and likewise the Derivative time is equal to the corner frequency as determined for the other portion of the controller response. Therefore by adjusting either the Integral time, the Gain, Derivative time, or any combination of the three modes, the controller and sum of the controller plus process will be altered. Again however experimentally this is difficult to collect this type of data for any real process.

When the combined open loop response is plotted, Process plus Controller, the stability of the loop can also be checked. For the system to be stable, the sum of the gains must be less than 1 at a phase shift equal to -180 degrees. There are many settings that will provide stable control and many that are unstable.



PID control is typically reserved for temperature control where thermal lags in the reboiler (either exchangers or fired heaters) and transport lags internally in the column makes control more difficult. Because it takes longer for a change in control valve position to have an effect on temperature, the derivative action is a means of offsetting that delay. When the temperature begins to change, a controller with derivative action responds immediately anticipating or predicting the temperature in advance based on the rate of change.

For the gas concentration unit stripper flow is cascaded with flow to the reboiler heat input. The system deals with again similar thermal and transport lags in the reboiler circuit and derivative action is useful for the secondary flow controller.

However because the secondary controller involves a flow measurement, dampening of the flow transmitter is required to eliminate any high frequency noise. Remember derivative action is based on rate of change and a noisy process variable will have a negative impact on good control.

Controller Tuning

- **Frequency Response Plots**
  - **Difficult to obtain data**
  - **Excessive amount of time gathering data**
  - **Unit continuously upset**
- **Real Time Methods**
  - **Closed Loop Method**
    - **Ziegler-Nichols Method**
  - **Open Loop Method**
    - **Process Reaction Curve**

uop
A Honeywell Company

EDS 2006/Inst-149

As we reviewed the Bode Plots (frequency response curves), one feature was quite evident. Tuning parameters, Gain, T_i, and T_d, were randomly selected to illustrate the slope and magnitude of their contribution to the controller frequency response. To determine the frequency response of the controller, the process variable would need to be varied in a sinusoidal fashion and the respective output of the controller would be recorded for various combinations of the tuning parameters. On real processes this is very impractical. The bode plots lend themselves to a good understanding of control theory, but in the real world are not practical for loop tuning.

There are various methods available for loop tuning. Two of the more common methods are the Ziegler-Nichols closed loop method and the Process Reaction Curve open loop method. Both have their advantages and disadvantages. However both will provide adequate tuning parameters that will result in stable control, and in both cases, additional fine tuning of the control loop will provide adequate closed loop control.

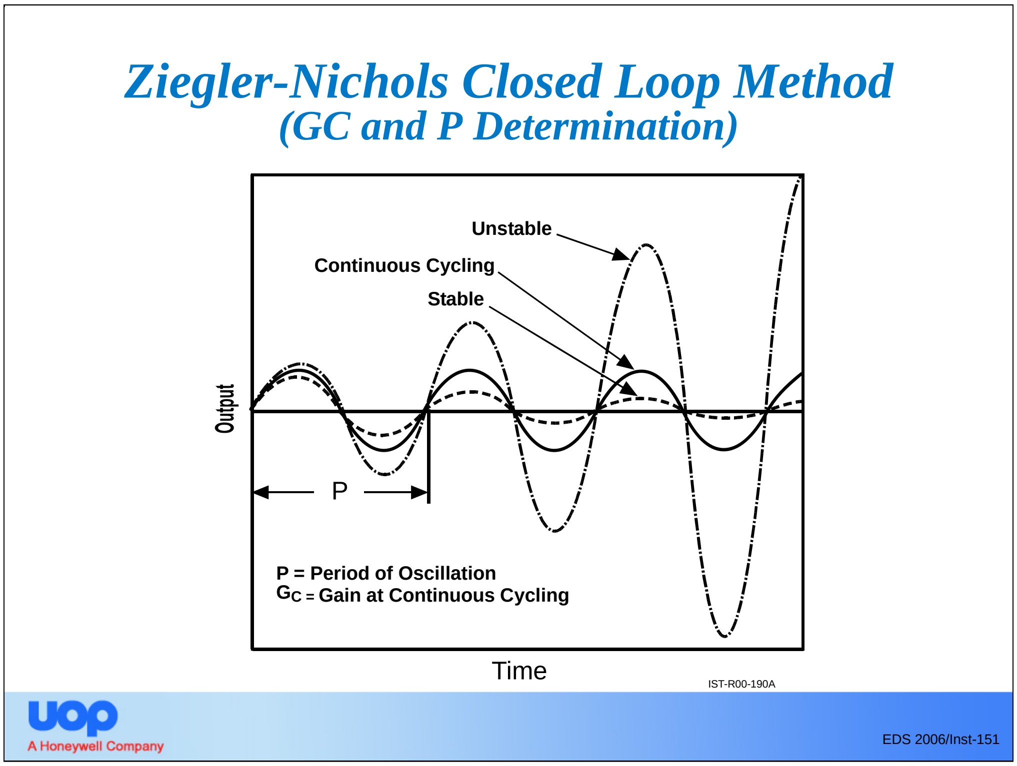
Ziegler-Nichols Closed Loop Method

- **On-Line Closed Loop**
- **Determine Period of Oscillation for P-only**
- **Procedure**
  - **In manual line-out P-only controller**
  - **Set Gain; switch to automatic mode**
  - **Change Set Point; 2-3% Step**
  - **Observe controller output oscillation**
  - **adjust Gain accordingly, repeating procedure until continuous oscillation (G_C) is found**

A Honeywell Company

EDS 2006/Inst-150

The Ziegler-Nichols closed loop method determines the gain (proportional action) and period of Oscillation for continuous cycling with a Proportional only controller. At this gain the total gain of the loop is 1.0. For higher controller gains the total gain of the loop is greater than 1.0, and the loop will be unstable (the magnitude of each additional period is growing larger). For lower controller gains the total gain of the loop is less than 1.0; however, the loop will be stable (the magnitude of each additional period is growing smaller)and will eventually line-out at some steady state value.



By following the procedure, a trial and error approach is conducted until the gain of the controller produces continuous oscillation. Continuous oscillation occurs when the magnitude of each cycle is the same. If the gain of the controller is too large then the process is unstable and the resultant oscillation increases in magnitude.

Once continuous oscillation is found, the gain of the controller, Gc, and period of oscillation, P, are noted.

Ziegler-Nichols Closed Loop Method
(Calculating P, I, and D Tuning Constants)

- **Proportional only**
  - **$G = 0.50 \times G_C$**
- **Proportional plus Integral**
  - **$G = 0.45 \times G_C$**
  - **$T_i = 0.85 \times P$**
- **Proportional plus Integral plus Derivative**
  - **$G = 0.60 \times G_C$**
  - **$T_i = 0.50 \times P$**
  - **$T_d = 0.12 \times P$**

UOP
A Honeywell Company

EDS 2006/Inst-152

For level control UOP will typically recommend proportional only control or PI control with a large Integral time to eventually bring the level back to its set point (surge drum level control). For most flow and pressure control loops, PI control is sufficient. PID control is usually limited to temperature control loops; and in some instances, PID control can be used on analyzer loops.

Once the controller modes are selected, whether it be P, PI, or PID control, the tuning parameters are estimated as shown above based on the Gain and Period determined from continuous oscillation.

Ziegler-Nichols Closed Loop Method
(Determining P, I, and D constants for a PID Controller)

- **For a Temperature Controller**
  - **it was determined that G_C = 10 and P = 5 min**
  - **select PID controller**
- **Calculate Tuning Parameters**
  - **G = 0.60 x G_C = 0.60 x 10 = 6.0**
  - **I = 0.50 x P = 0.50 x 5 = 2.5 minutes**
  - **D = 0.12 x P = 0.12 x 5 = 0.6 minutes**

A Honeywell Company

EDS 2006/Inst-153

As an example the Ziegler-Nichols closed loop method was applied to a temperature control loop. The closed loop method yielded continuous oscillation with the gain of the controller set at 10. The Period of Oscillation was measured to be 5 minutes.

PID controller is selected for the temperature loops and the collected data yields the G equal to 6, Ti equal to 2.5 minutes, and T_d equal to 0.6 minutes.

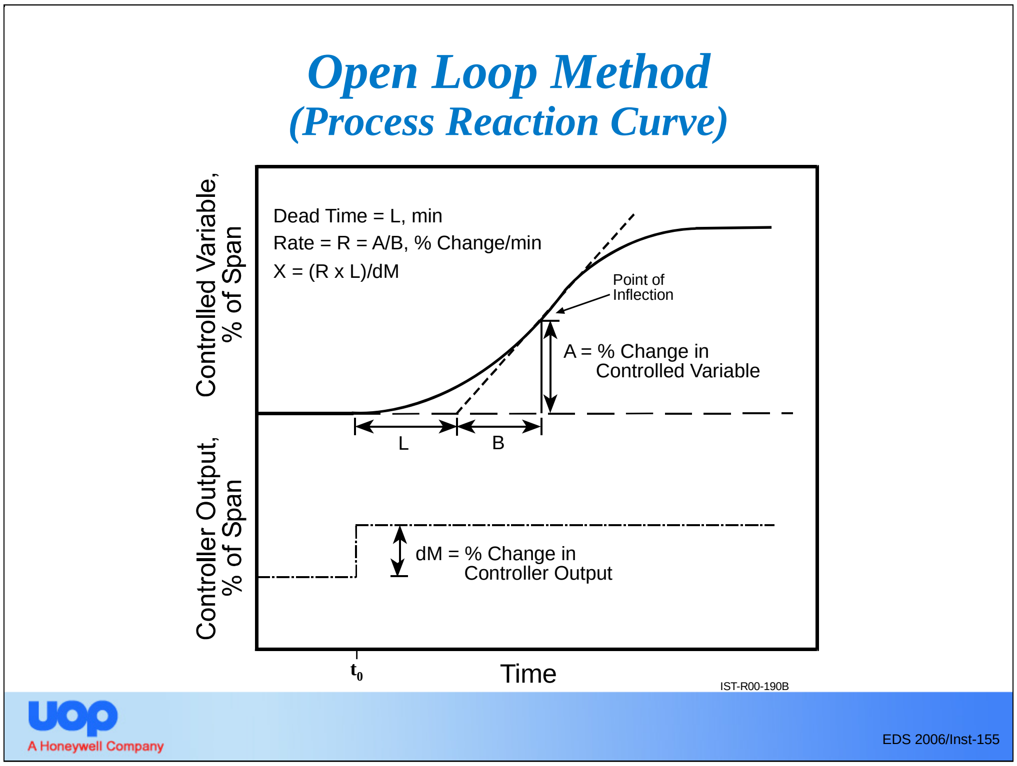
Open Loop Method

- **On-Line Open Loop**
- **Determine Dead Time and Reaction Rate**
- **Procedure for Process Reaction Curve**
  - **Line out at steady state with manual control**
  - **Induce step change (dM) in controller output**
  - **Observe the Process Reaction Curve**
  - **Obtain Dead Time (L) and Reaction Rate (R)**
  - **Calculate Parameter (X) = (L x R)/(dM)**

UOP A Honeywell Company

EDS 2006/Inst-154

The Process Reaction Curve open loop method determines the dead time and reaction rate of the process. Once the process dead time and reaction rate are determined, then the tuning parameters can be determined. The procedure is straight forward and easy to implement.



The Procedure for Process Reaction Curve is as follows:

1) With the controller in manual, line out at steady state
2) Induce a step change (dM) in controller output (2 - 5% change)
3) Observe the Process Reaction Curve
4) Draw a straight line through the inflection point on the curve
5) Estimate Dead Time (L) and Reaction Rate (R)
6) Calculate Parameter (X) = (L x R)/(dM)

Step 4 is the most difficult step with this graphical method. L and R are based on the inflection point and slope of the line drawn.

Open Loop Method
(Calculating P, I, and D Tuning Constants)

- **Process Reaction Curve Yields**
  - **Step (dM) % Change in Controller Output**
  - **Dead Time (L)**
  - **Rate (R) % Change in Controlled Variable/min**
  - **Parameter (X)**
- **Controller Tuning Constants**
  - **$G = 0.375 \times (X)$ (Range 0.25 to 0.5)**
  - **$T_i = 2.5 \times (L)$ (Range 2.0 to 3.0)**
  - **$T_d = 0.5 \times (L)$ (when Derivative is used)**

UOP A Honeywell Company EDS 2006/Inst-156

Once the parameters are extracted from the Process Reaction Curve the controller tuning parameters can be determined. Due to the uncertainty in the Process Reaction Curve the tuning parameters have bracketed ranges and can be adjusted with in these ranges for overall improvement in the control loop response.

Open Loop vs. Closed Loop Method

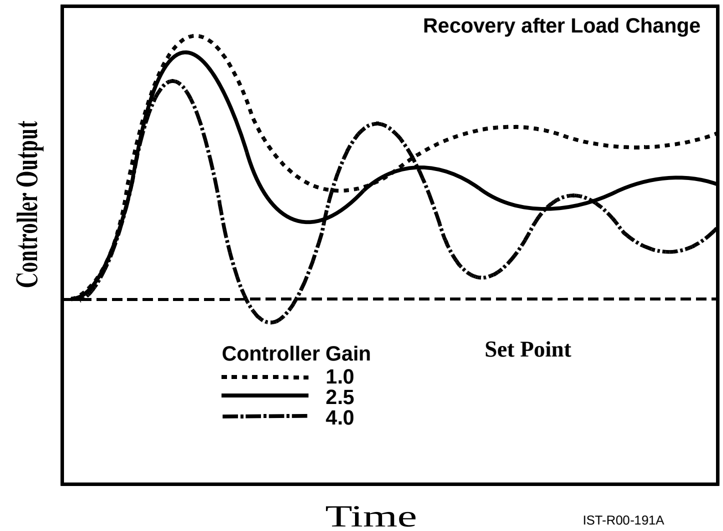
- **Advantages**
  - **Tuning Constants obtained in minimal time**
  - **Unpredictable amplitude oscillations avoided**
  - **Small Step Change avoids Process Upsets**
- **Disadvantages**
  - **Controller and Valve Dynamics not included**
  - **Noise interferes with Graphical Parameters**
  - **Tuning Constants not as accurate**

UOP A Honeywell Company EDS 2006/Inst-157

These are some of the advantages and disadvantages of the open loop method compared to the closed loop method. The open loop method minimizes the time to obtain the process reaction curve, avoid unpredictable amplitude oscillations, and can be implemented with little likelihood of a process upset.

However as pointed out previously, this graphical method interferes with a noisy process and the tuning constants are not as accurate partly due to the fact that the straight line drawn through the inflection point will alter the results if not drawn correctly. Another disadvantage of this method, is the valve dynamics and controller dynamics are not included because it is an open loop test method.

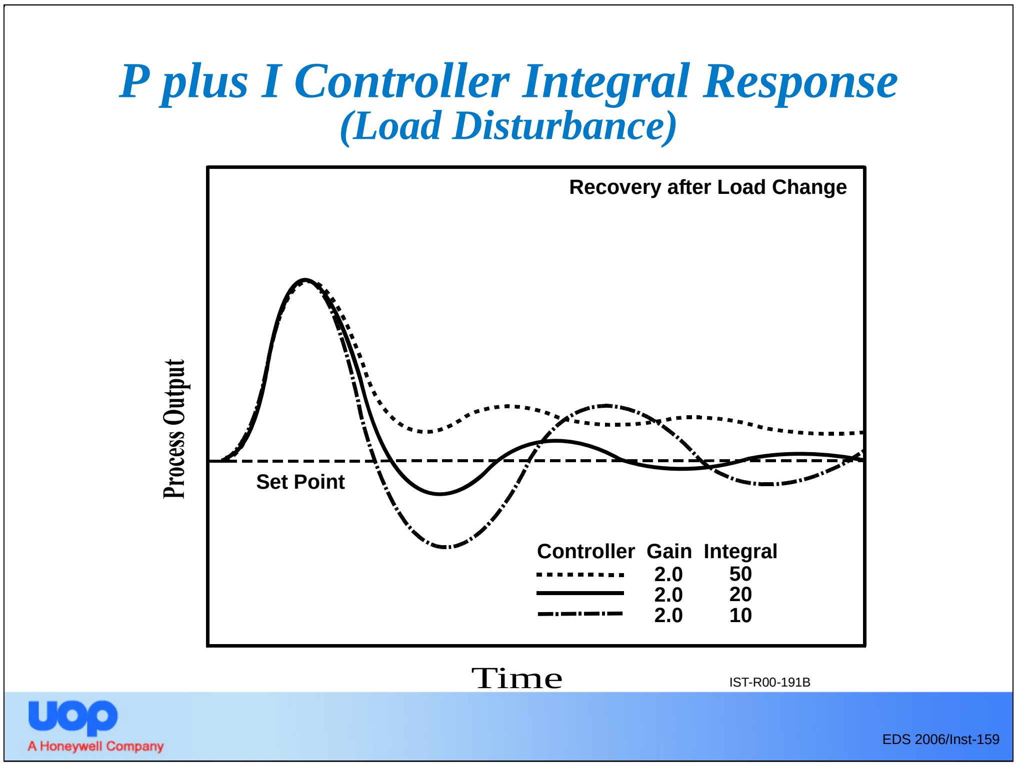
Proportional Only Controller Response



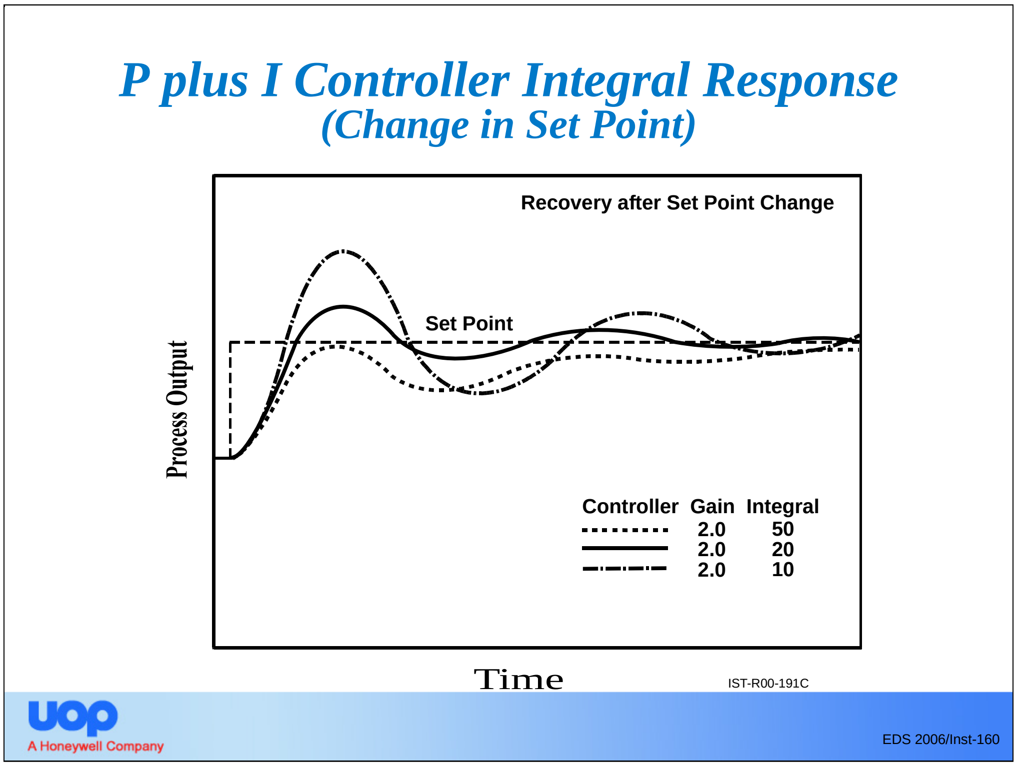
A Honeywell Company

EDS 2006/Inst-158

For the example provided the proportional only controller is at steady state prior to a load change. The gain of the controller was increased as indicated from an initial setting of 1.0 to 4.0. As the gain is increased the offset is smaller, but the period of oscillation is decreasing, the frequency of oscillation is increasing, and the control loop is becoming less stable. For a proportional only controller with level control, setting the gain at around 2.5 would be adequate.

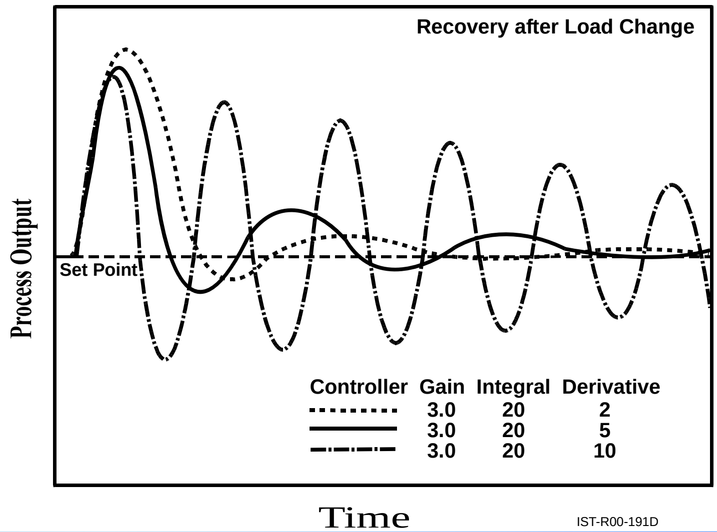


For PI control the gain is fixed at 2.0 in this example and the Integral action is varied from 50 seconds/repeat down to 10 seconds/repeat. With the Integral time at 50, the controller appears to respond as a proportional only controller. Eventually with the small amount of Integral action the process variable will line-out at it's set point. As the Integral action is increase the amplitude of oscillation tends to increase, the period of oscillation increases, and as a result overshoot also increases. As can be seen above too much integral action causes excessive gain in the controller and a significant amount of oscillation around the set point. For this example Integral time somewhere around 20 second/repeat provides adequate control for a load change on the process.



This example explores the same process as previously described, but instead of a load variable disturbance we have a change in set point that disturbs the system. With the controller gain remaining at 2.0 and the Integral time varying between 50 and 10, a similar response is seen for the variation in Integral time. Again with very little Integral time (50 seconds/repeat) the controller responds slowly to the change in set point but eventually will line out at the new set point. As the Integral time is decreased (more Integral action) the magnitude of the oscillations increase and more overshoot is experienced. Again an Integral time at 20 provides adequate control.

PID Controller Derivative Response
(Load Disturbance)



IST-R00-191D

A Honeywell Company

EDS 2006/Inst-161

Addition of the Derivative mode has a tendency of reducing the period of oscillation. Unlike Integral action, which accounts for a lag on the system, Derivative action anticipates how fast the error is changing and accounts for a lead on the system. This actually counterbalances the lag generated by the Integral action. As can be seen in this example with gain and Integral time held constant, an increase in Derivative time increases the controller gain. As a result the magnitude of the oscillations is increased and will lead to instability. For this example a Derivative time of 2 minutes provides adequate control minimizing overshoot and drives the process variable back to it’s set point.

Initial Tuning Constants

	GAIN (PB%)	INTEGRAL (min/repeat)	DERIVATIVE (minutes)
LEVEL	1.1 (90%)	> 60	0
FLOW	0.4 (250%)	0.05 – 0.10	0
PRESSURE (gas)	2.0 (50%)	0.5	0
TEMPERATURE	2.5 (40%)	0.5	0.5 – 1.0

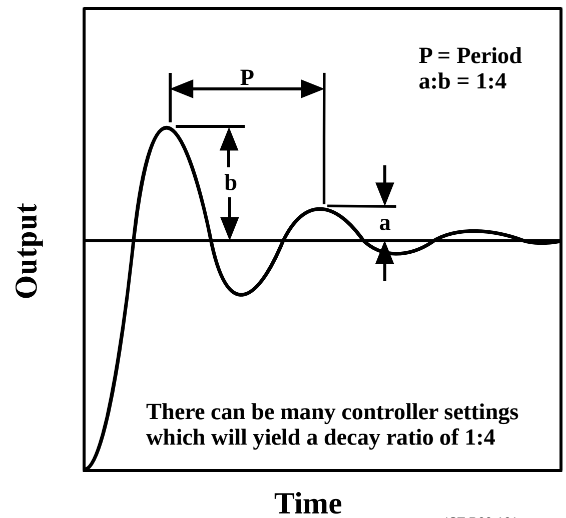
- Liquid pressure or liquid differential pressure loops are set similar to flow.
- Column tray temperature loops use higher integral and lower derivative times.
- Analyzer (composition) loops are set similar to tray temperature.
- UOP philosophy on level control allows vessel to act as surge volume allowing offsets to occur in the level for set point or load changes. The addition of a long integral time, will slowly return the process variable to its set point.

A Honeywell Company

EDS 2006/Inst-162

In general UOP recommends the above settings for initial tuning parameters at start-up of a new unit. These values will typically provide stable but sluggish control. The intent here is to insure that no one controller is unstable. Once the unit is up and running, loop tuning can be address on an individual basis.

Quarter Amplitude-Decay Ratio



A Honeywell Company

EDS 2006/Inst-163

What should one shoot for in terms of controller response? One method looks at quarter amplitude decay ratio. In this method the ratio of Magnitude on successive cycles is 4 to 1. Another method may be to minimize overshoot. By reducing either the gain or Integral time in the above example would reduce the controller gain; thus the overshoot would be reduced and the magnitude of the oscillations would be less (this would result in a higher ratio as compared to the Quarter Amplitude-Decay Ratio).

Care must be taken not to reduce the gain significantly otherwise the controller will become too sluggish and will not provide sufficient gain in the controller to drive the PV back to its set point in a reasonable amount of time.

Feedback Control Loop

- *Loop Components*
  - **Controller**
    - **P, PI, and PID Modes**
    - **Applications**
    - **Tuning Methods**

- **DCS System Requirements**
  - **Control System Performance**
  - **Loop Integrity**
  - **Data Storage and Retrieval**
  - **Analog Outputs**

A Honeywell Company

EDS 2006/Inst-164

This concludes our discussion on the feedback control loop and the individual components of the feedback control loop. Now we will look at some of the UOP requirements for distributive control systems (DCS). Performance, integrity, historical data storage, etc will be briefly discussed.

System Requirements
(Control System Performance)

- **Digital DCS Systems**
  - **Control algorithm execution rate shall be used as a measure of performance**
  - **Execution rate is defined as the total time the control system takes to read the process data from the input processor, perform the control algorithm, and change the controller output to the output processor**

A Honeywell Company

EDS 2006/Inst-165

As discussed previously UOP uses as a measure of performance the control algorithm execution rates. This is the time it takes for the input processor to read the PV, update the controller, perform the control algorithm, update the controller output, and for the output processor to update the output to the field.

This execution rate is important for fast acting loops such as flow and liquid pressure loops.

Control Algorithm Performance

Control Service	Update Rate	Execution Rate
Flow	4 updates per second	250 milliseconds
Liquid Pressure	4 updates per second	250 milliseconds
Gas Pressure	2 updates per second	500 milliseconds
Differential Pressure	4 updates per second	250 milliseconds
Temperature	2 updates per second	500 milliseconds
Level	1 update per second	1 second
All Others (Analysis, pH, etc)	2 updates per second	500 milliseconds

A Honeywell Company

EDS 2006/Inst-166

Again for processes that are instantaneous (flow, liquid pressure, differential pressure) the update rate is 4 times per second as a minimum. For process that have some capacity in the system (gas pressure, temperature, etc), the update rate is 2 times per second. For level measurement, as long as the vessel has been adequately sized, the update rate is 1 time per second as a minimum. If the particular DCS system has Update Rates that exceed these requirements, process control will be enhanced provided that the transmitters have a comparable update rate.

System Requirements
(Control System Performance, cont'd)

- **Analog Electronic Systems**
  - **Control loop frequency response shall be used as a measure of performance**
  - **Control loop frequency response is defined as the frequency response from the output connection at the transmitter to the input connection at the actuator of the final control element**
  - **This frequency response shall be at least 1 Hertz (1 cycle per second)**

uop A Honeywell Company

EDS 2006/Inst-167

When we looked at the pneumatic tubing case, it was mentioned that the frequency response of the pneumatic system should be at least 1 cycle per second. This requirement lead to most pneumatic controllers to be field mounted because of the distances involved with centralized control rooms. For distances in excess of 100 feet the time constant was too large to provide adequate control for most processes.

With the advent of electronic controllers, moving now from the speed of sound to the speed of light, transmission of the PV and OP signals are essentially instantaneous and do not inhibit the control system as compared to the pneumatic transmission of these signals. However control loop frequency response shall also be used to evaluate electronic control systems. For most systems this requirement is generally met without limitations.

System Requirements

- **Loop Integrity**
  - **A single component failure will not cause the loss of more than one control loop**
  - **Shared equipment shall be furnished with redundant or "backup" systems**
    - **Uninterruptible Power Supply**
    - **Redundant Equipment**
  - **Standby Manual Stations provide access to the final control elements**

UOP A Honeywell Company

EDS 2006/Inst-168

Analog electronic controllers and single loop digital controllers meet the Loop integrity requirement. However design of DCS systems have optional equipment that can be added to meet the loop integrity criteria. Most DCS systems can be purchased with redundant power supplies, redundant controllers, redundant I/O processors, etc. Also UPS systems can be provided and sized to ensure that the system is operational for a given amount of time in the event of a main power failure to the DCS equipment.

An optional piece of equipment is the Standby Manual Stations. These devices allow the operator or maintenance personnel to manually drive the final control elements. This equipment is external to the DCS operator console and allows the operator the interface to the final control element when access to the operator consoles is loss.

System Requirements

- **Data Storage and Retrieval**
  - **Plant material balance determination**
  - **Equipment performance trend analysis**
  - **Malfunction troubleshooting analysis**

Minimum Requirement

Specified process variables shall be sampled and the instantaneous value shall be stored at intervals of five seconds or less. Storage capacity shall be sufficient to store process variable data for at least a 7 day period.

A Honeywell Company

EDS 2006/Inst-169

UOP will typically specify the minimum amount of data to be stored long term for each process. This data is can be used for Plant material balance considerations, equipment performance trends, troubleshooting, etc. The UOP P&ID's will graphically represent the data to be recorded long term. If the client wishes to store more data, UOP will not object. UOP's requirements, as shown on the P&ID's, are the minimum requirement and should be stored for at least a 3 day period.

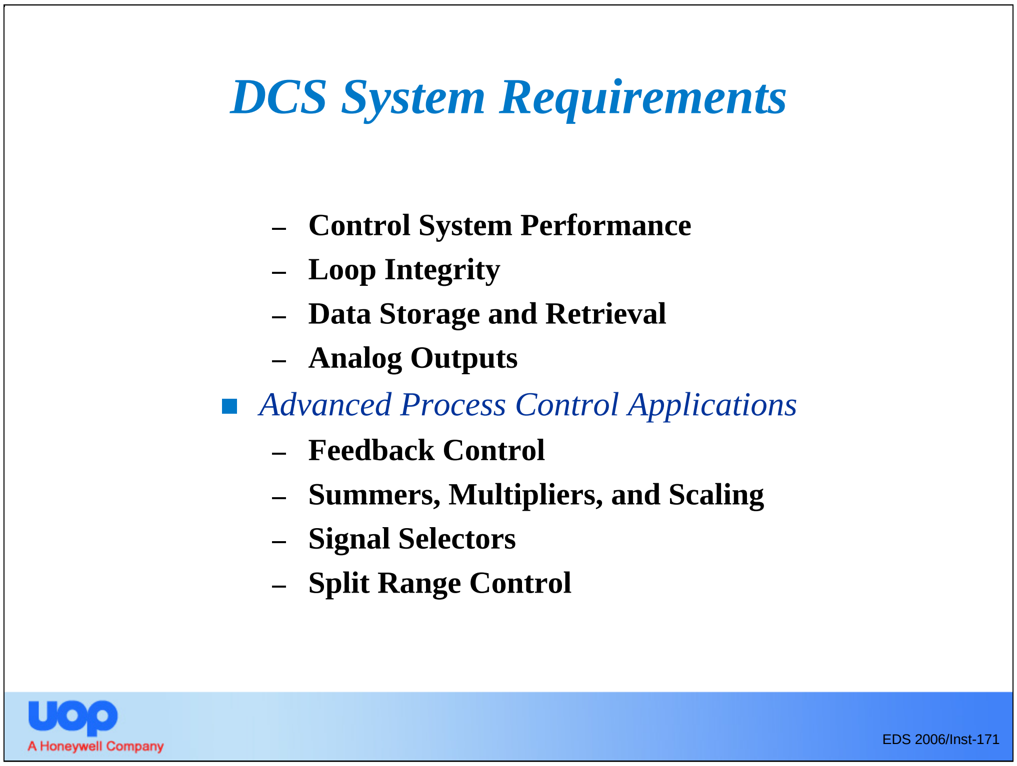
System Requirements

- **Analog Outputs**
  - **Controller and manual station outputs should be capable of operation from 25% below to 25% above the standard 0 to 100% signal range**
  - **The (+) or (-) 25% margin is used to provide additional force at the control valve actuator in order to insure proper control valve closure during normal operation**

uop
A Honeywell Company

EDS 2006/Inst-170

For analog systems, controllers and manual stations should be capable of operating at ±25% above/below the standard 0 to 100% signal. This is to ensure proper closure during normal operation of the control valves. However in DCS systems this number is more on the order of ±7 to ±10%. Actuator sizing and bench setting may be more important and more precise today than in years past to ensure that the valves seat properly when closed to prevent inadvertent leakage.



This was a brief review of the DCS system requirements. Now we will start looking into applications of the Feedback Control loop along with some of the APC applications.

Advanced Process Control Applications (continued)

– **Cascade Control**
– **Ratio Control**
– **Feedforward Control**

We will examine some simple APC along with some of the more difficult APC approaches. Feedforward and dynamic compensation are complex functions and rely heavily on process modeling. However if the process model is simulated well then their contribution greatly enhances the overall control system.

Feedback Control Loop
(Controller Action)

- **Definition: Controller Action is the change in controller output in response to an increase in process variable above its set point**
  - **Direct Acting Controller: For an increase in process variable above its set point, output of the controller increases**
  - **Reverse Acting Controller: For an increase in process variable above its set point, output of the controller decreases**

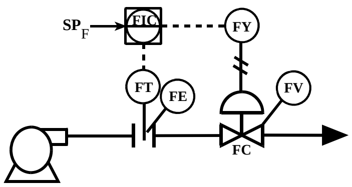
A Honeywell Company

EDS 2006/Inst-173

Earlier when we reviewed the various equations for P, PI, & PID control, error was defined as PV-SP. With this as the definition of error, if the PV goes above the SP then by definition OP of the controller increases. If we apply this to a simple flow control loop and the control valve is air to open (fail closed actuator), then for the case of an increase in flow, the output of the controller would increase. What happens to the control valve in this situation? The valve opens? What happens then to the flow rate when the valve opens? The flow rate goes up. Is this what we want? NO!!!!!!! What if the error was defined as SP- PV? Then for an increase in PV above the SP, the error would be negative and the OP of the controller would decrease. And if the valve is air to open, then the valve would start to close, and the flow rate would be reduce moving towards the SP.

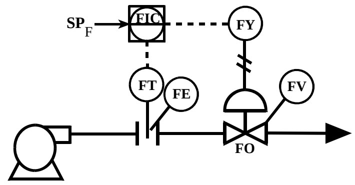
As a result any controller has the option of being either a direct acting or a reverse acting controller. The evaluation of whether the controller should be direct acting or reverse acting is done on a loop by loop basis. During configuration and commissioning the instrument personnel will configure/set the controller action appropriately.

Feedback Control Loop
(Controller Action)



Element	Action
FE (Process)	Direct
FT	Direct
FIC	unknown
FY	Direct
FV	Direct

Controller FIC needs to be Reverse Action



Element	Action
FE (Process)	Direct
FT	Direct
FIC	unknown
FY	Direct
FV	Reverse

Controller FIC needs to be Direct Action

EDS 2006/Inst-174

In order for any feedback control loop to function properly, the loop must have negative feedback (in terms of frequency response this negative feedback corresponds to a 180 degree phase shift somewhere in the loop).

One can logically step through the process to determine the appropriate controller action required for any feedback loop. For the example on the left with an air to open valve, if something happens that causes the flow to increase above the set point, the flow transmitter output to the controller also increases. In order for the flow control loop to function properly, the control valve must close to reduce the flow and bring the flow back to the controller's set point. Therefore the output of the controller must provide the 180 degree phase shift in the loop; when the input signal to the controller is increasing, then the controller output is decreasing (a reverse acting controller). This phase shift takes place in the controller as a result of specifying a reverse acting controller.

Another approach in determining the appropriate controller action is to evaluate the action of each individual components in the loop. Each reverse acting component will provide a 180 degree phase shift. Therefore an odd number of reverse acting components are required in the loop in order to provide a net 180 degree phase shift. The example on the right has an air to close control valve (fail open valve), which is a reverse acting component. If the controller action was set to reverse acting, the net phase shift is not 180 degrees and the loop will not function properly.

Summers, Multipliers, and Scaling

- **Develop Engineering Equations**
  - **relevant terms must be included**
  - **mathematics precisely defined**
- **Hardware Considerations**
  - **analog and low level digital devices**
  - **digital control systems**
- **Normalize the Variables**
- **Substitute and Simplify**

UOP A Honeywell Company EDS 2006/Inst-175

Summers and Multipliers are a couple of functions frequently employed to mathematically manipulate either input or output signals. Depending upon the task at hand, relevant mathematical equations need to be determined, hardware selected, and instrument signals scaled.

These type of applications are as simple as split ranging a control signal to two or more control valves to the more complex feed forward mathematical modeling applications and beyond.

Summers, Multipliers, and Scaling
(Develop Engineering Equations)

- **Basic Flow Equations**
  - **Liquid:** $F = C_{op}[(h_w)/(Gr_f)]^{0.5}$
  - **Vapor:** $Q = C_{op}[(h_w)(P_f)/(T_f)]^{0.5}$
  - **Summing:** $F_T = F_1 + F_2$
- **Exchanger Duty**
  - **Heat Transfer:** $Q = (M)(Cp)(DT)$
- **Process Conditions**
  - **Pressure:** $P_a = P_g + 14.7$
  - **Temperature:** $T_R = T_F + 460$

uop A Honeywell Company

EDS 2006/Inst-176

The first step is to develop the engineering equations to represent the task at hand. As an example the basic flow equation for vapor service is shown above. If during normal operation the pressure and/or temperature differs from the design basis, the measured flow is in error. Obviously the greater the deviation the greater the error. One technique is to measure both pressure and temperature and mathematically compensate for deviation from the design conditions.

For any orifice plate in vapor service, the meter flow range, differential pressure range, a design temperature (T_f), and a design pressure P_f was chosen from the process data as a design basis for the manufacturing of the orifice plate. The engineering equation is shown above and the coefficient of the orifice plate, C_{op}, can be obtained from the design data supplied with the orifice plate from the manufacturer.

Other types of equations are represented above. In some instances signal conditioning may be as simple as converting a pressure or temperature measurement to an absolute scale.

Summers, Multipliers, and Scaling
(Hardware Considerations)

- **Analog and low level digital devices**
  - **Scaling typically required**
  - **Limited to simply math functions**
  - **May require more than 1 device**
  - **Changes require recalibration**
- **Digital control systems**
  - **Based on engineering units**
  - **Complex math functions**
  - **Application reconfigurable**

UOP A Honeywell Company EDS 2006/Inst-177

With the electronic analog systems, the instrument signals are typically 4 - 20 ma signals representing 0 to 100% of the span of the instrument in Engineering Units. Scaling of the instrument signal is required in order to manipulate these signals mathematically. Because of the complexity of these type of systems, more than 1 device may be required; and in most applications, respanning any transmitter input signal will require recalibration of the device(s).

With the advent of the digital control systems, a major portion of the complexity has been removed. Typically the DCS systems convert the input signals to engineering units; thus the scaling of the instrument signals is eliminated. Also complex mathematical functions can be utilized in the DCS systems allowing for a wide variety of applications.

Summers, Multipliers, and Scaling
(Normalize the Variables)

- **Scaling Instrument Signals**
  - **$E = Span(E^n) + Zero$**
    - **E = Engineering Units**
    - **Span = Upper Value - Lower Value**
    - **E^n = Instrument Signal (0 to 1.0 or 100%)**
    - **Zero = Lower Value**
- **Process Variable Examples**
  - **$T_F = 100(T^n) + 50$**
  - **$Pg = 50(P^n) + 0 = 50(P^n)$**
  - **$Gr = 0.5(Gr^n) + 0.5$**

A Honeywell Company

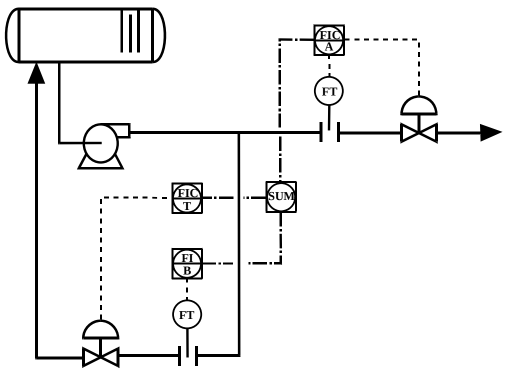
EDS 2006/Inst-178

For a given pressure transmitter that has been calibrated for 0 to 50 psig, the instrument signal (0 to 100%) represents 0 to 50 psig. The zero of the instrument is the Lower Value; and in this case, Lower Value is equal to 0. The span of the instrument is equal to the Upper Value minus the Lower Value. In this case the Upper Value is 50. Therefore the span is 50 minus 0, which is 50. If the pressure in Engineering Units is 25 psig, then the instrument signal in terms of % would be 50% (0.5). Mathematically then a signal of 50% would represent 25 psig in Engineering units based on this equation.

Scaling instrument signals is typically the first step in conditioning these signals for the various applications.

Summers, Multipliers, and Scaling
(Summing 2 Flows)

Minimum Flow Spillback



Design Considerations:

FIC_A: 0 - 100 gpm
FI_B: 0 - 40 gpm
Minimum Flow: 30 gpm

Engineering Equation:

$FIC_T = FIC_A + FI_B$

Scaled Equations:

$FIC_A = 100FIC^n_A + 0$
$FI_B = 40FI^n_B + 0$
$FIC_T = (100 + 40)FIC^n_T + 0$

IST-R00-193A

uop A Honeywell Company

EDS 2006/Inst-179

Looking at the summer in the above example the engineering equation used is the sum of the two flow meters. The first step is to scale each flow measurement based on the design conditions of each meter. Once each flow meter scaled equation is determined, the next step is to substitute into the engineering equation and simplify.

Summers, Multipliers, and Scaling
(Summing 2 Flows cont'd)

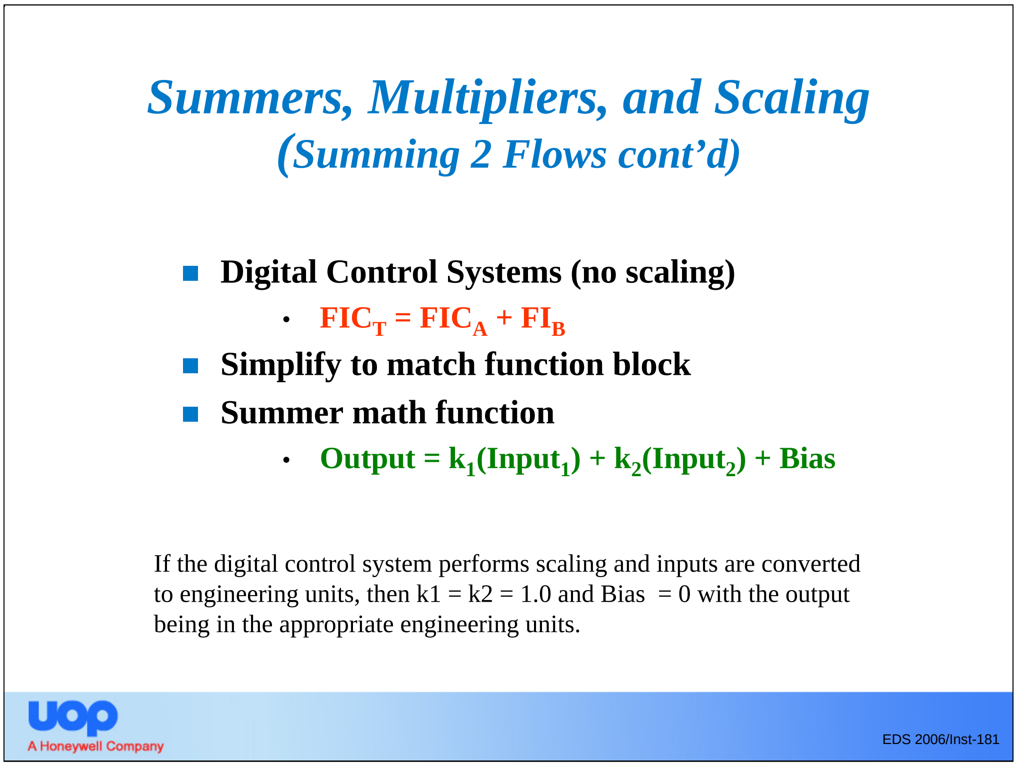
- **Engineering Equation**
  - $FIC_T = FIC_A + FI_B$
- **Substitute Scaled Equations**
  - $(140)FIC_T^n = (100)FIC_A^n + (40)FI_B^n$
- **Simplify to match hardware**
  - $FIC_T^n = (0.713)FIC_A^n + (0.287)FI_B^n$
- **Analog Summer (2 inputs/1 output)**
  - $Output = k_1(Input_1) + k_2(Input_2) + Bias$

Output is a normalized signal ranging from 0 to 1.0;
span has been chosen to be 0 - 140 gpm

UOP
A Honeywell Company

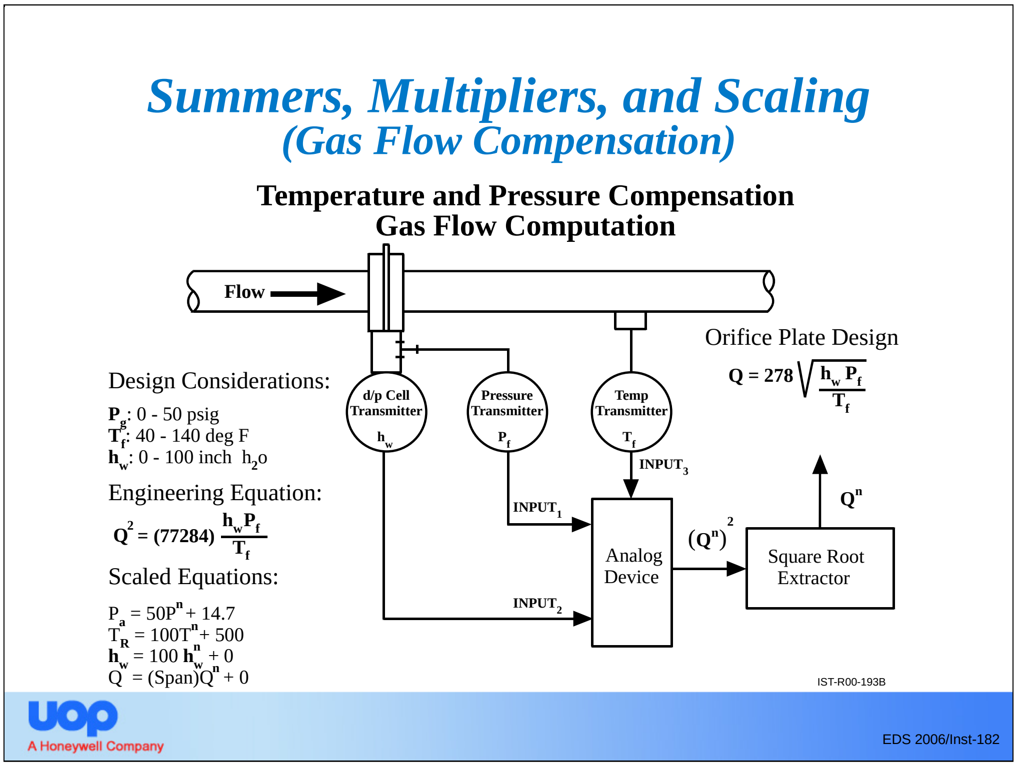
EDS 2006/Inst-180

After the equations have been scaled and simplified, specific hardware is selected to accomplish the task. In the above example an analog summer with w in puts and 1 output was selected to sum the two flow signals. If both signals are at 50%, then the total flow would be 50 plus 20 or 70 gpm. Analyzing the analog summer equation results in 0.713 x 0.5 + 0.287 x 0.5 + 0. This simplifies to 0.3565 + 0.1435 + 0 = 0.5. The output of the summer is 50%. The span is 140; therefore the analog summer also represents 70 gpm.



In a DCS system the input signals are converted to Engineering units. The function block equation is provided above and the constants are set to 1. Therefore the summer of the two is quite simple compared to the analog device previously illustrated.

If one or both of these transmitters are respanned to different engineering units, the analog device equation will change and recalibration of the device is mandatory. For the DCS system the equation stays the same. A simple change in the flow measurement span is required, but can be accomplished in a matter of seconds by the control engineer. Since the DCS is using Engineering Units a minimal amount of work is required to update the configuration of the loop in question.



Gas flow compensation is common within many process streams coming into or leaving a particular process unit as a means of material balance closure. Based on the engineering equation for vapor flow through an orifice plate, the relationship exists as shown previously. Assuming that the molecular weight of the gas is unchanged for this example, the engineering equation has three inputs: the differential pressure across the orifice plate, the flowing pressure and the flow temperature. Again the first step is to develop the scaled equations for the instrument signals, followed with substitution and simplification into the engineering equation. Due to the complexity of this task, two analog devices are required to complete the flow compensation for changes in operating temperature and/or pressure.

Summers, Multipliers, and Scaling
(Gas Flow Compensation cont'd)

- **Engineering Equation**
  - $Q = C_{op}[(h_w)(P_f)/(T_f)]^{0.5}$ where $C_{op} = 278$
- **Determine flow meter span**
  - $Qmax = 278[(100)(64.7)/(500)]^{0.5} = 1000$ scfh
  - $Q = 1000Q^n + 0$
- **Substitute Scaled Equations**
  - $1000Q^n = 278[(100h_w{}^n)(50P^n + 14.7)/(100T^n + 500)]^{0.5}$
  - $(Q^n)^2 = 0.077284[(100h_w{}^n)(50P^n + 14.7)/(100T^n + 500)]$
  - $(Q^n)^2 = 7.7284[(h_w{}^n)(50P^n + 14.7)/(100T^n + 500)]$

UOP A Honeywell Company EDS 2006/Inst-183

The maximum flow within the range of pressure and temperature measurements will occur when the pressure is at its 100% point (64.7 psia) and when the temperature is at is 0% point (500 °R). This value is then used to set up the scaled equation for the gas flow rate. Also note that the coefficient of the orifice plate was given from the manufacturer as a value of 278.

Summers, Multipliers, and Scaling

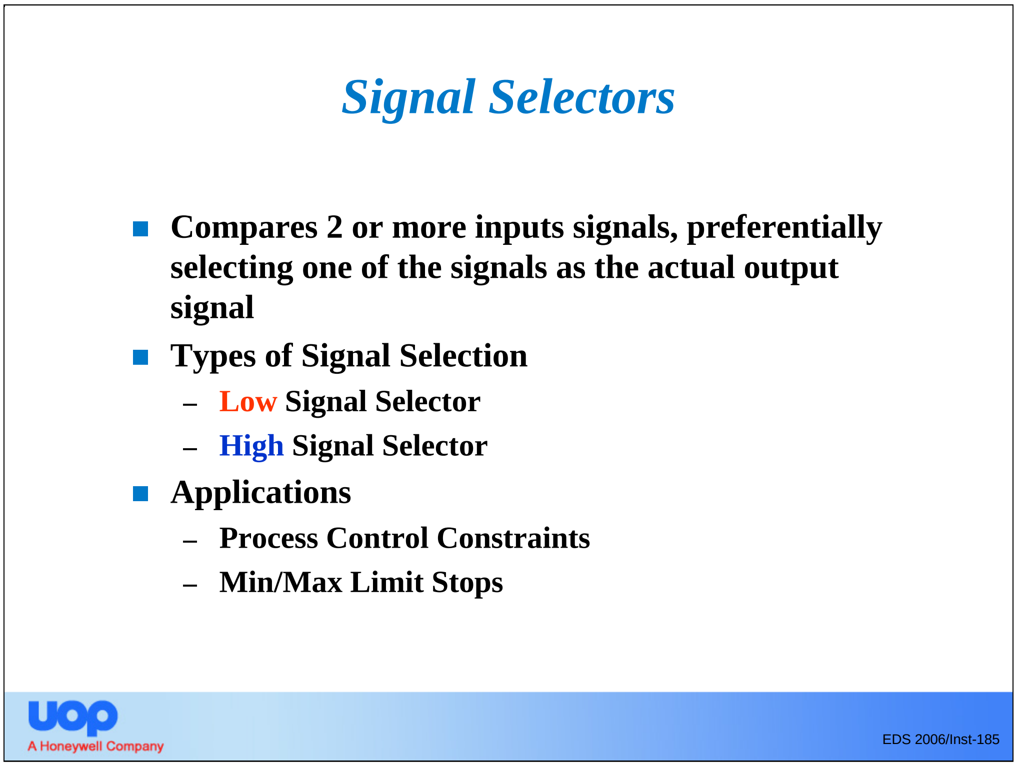
(Gas Flow Compensation cont'd)

- **Simplify to match hardware**
  - $(Q^n)^2 = 7.7284[(h_w)(50P^n + 14.7)/(100T^n + 500)]$
    - **Input signals must be factored not to exceed 1.0**
  - $(50P^n + 14.7) = 64.7[(50/64.7)(P^n) + (14.7/64.7)]$
  - $(50P^n + 14.7) = 64.7[0.77(P^n) + 0.23]$
  - $(100T^n + 500) = 600[0.17(T^n)+ 0.83]$
  - $(Q^n)^2 = 0.83(h_w{}^n)[(0.77P^n + 0.23)/(0.17T^n + 0.83)]$
- **Analog Device (3 inputs/1 output)**
  - $Output = [k_1 Input_1 + b_1][k_2 Input_2 + b_2]/ [k_3 Input_3 + b_3]$
  - **Square Root Extractor yields flow signal**

Output from analog device is proportional to flow^2; Output from Square Root Extractor is a normalized flow signal (0 - 1.0) with a span of 0 - 1000 SCFH.

As shown in the previous diagram the first analog device has 3 inputs and 1 output with the mathematical function as indicated above. The differential pressure measurement becomes Input 1, the pressure measurement becomes input 2, and the temperature measurement becomes input 3. The output of the device is proportional to the flow squared. The second device is a square root extractor, which effectively has one input and 1 output. The output of the square root extractor is a normalized flow signal representing 0 to 100% of the flow signal. As we determined earlier in terms of Engineering Units, 0 to 100% represents 0 to 1000SCFH.

Again in the analog devices if any of the instruments spans are changed, then the first analog device would have to be recalibrated also. However in a DCS system, where the input signals are converted to Engineering Units, the changes would be minimal.

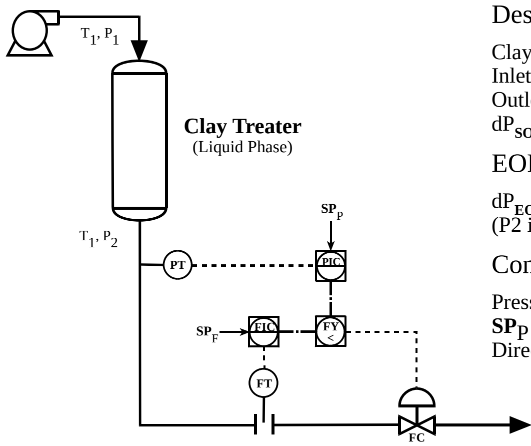


Signal selectors are a means of overriding control signals. These selectors compare 2 or more inputs and preferentially selects one of the control signals for use in the control loop while ignoring the other(s) control signal.

A couple of the more common applications include process constraint and limit stop. We shall look at how the signal selectors are applied.

Signal Selectors
(Process Constraint)

Pressure Override Control



Design Considerations:

Clay Treater operates at:
Inlet - T_1 and P_1
Outlet - T_1 and P_2
$dP_{SOR} = P_1 - P_2$

EOR Considerations:

$dP_{EOR} = P_1 - P_2 > dP_{SOR}$
(P2 is decreasing with time)

Controllers:

Pressure	Flow
SP$_P$ $> P_V$	**SP**$_F$ = Normal (100%)
Direct	Reverse

IST-R00-195A

UOP
A Honeywell Company

EDS 2006/Inst-186

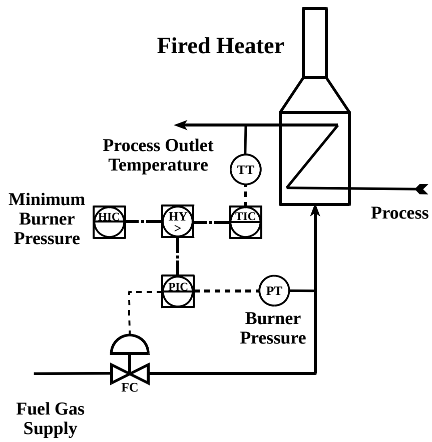
In the example above the liquid has been heated to T_1 and has a normal operating pressure of P_1 at the discharge of the pump for a flow at 100%. The clay treater is used to extract water from the incoming feed. The clay treater has been designed with a maximum differential pressure at end of run conditions. At this point Operations can plan a turn-a-round to replace or regenerate the clay material as needed. If P_2 drops below the vapor pressure of the fluid, the clay material in the clay treater will be damaged if vaporization occurs. We want to prevent this from happening at all cost.

Under normal operation P_2 is well above the vapor pressure of the fluid. As time progresses the delta P across the bed rises and P_2 decreases approaching the vapor pressure of the fluid. Feed to the downstream unit is on flow control. Given that the control valve is an air to open control valve, we see that the controller action needs to be reverse acting. In the event that the pressure at the outlet of the clay treater approaches the anticipated vapor pressure of the fluid (plus some margin), the operator could place the flow controller in manual and manually decrease the flow keeping P_2 arbitrarily above the vapor pressure.

Alternatively an override process constraint control can be added with the use of a low signal selector. A pressure transmitter is required on the outlet of the clay treater and a pressure controller is required. The low signal selector compares the output of each controller and passes the lower of the two signals to the control valve. The PIC controller action has to be direct acting.

Signal Selectors
(Minimum Limit Stop)

Minimum Burner Pressure Limit



Design Considerations:

Process Temperature Controller is cascaded with Burner Pressure Controller.
TIC output is PIC set point.
HIC output is minimum burner pressure limit for safe operation.

Controllers:

Temperature	Pressure
Reverse	Reverse

IST-R00-195B

uop A Honeywell Company

EDS 2006/Inst-187

The above control scheme has the primary temperature controller adjusting the fuel gas burner pressure to vary the heat release in the fired heater. If the process requires more heat release, the output of the temperature controller increases the burner pressure controller’s set point. If the process requires less heat release, the temperature controller decrease the set point.

During turndown operation the possibility exists of lowering the burner pressure to a point where there is flame-out at the burner tips. Needless to say this is a dangerous scenario and must be avoided. A high signal selector can be introduced into the system to limit just how low the burner pressure set point can be.

Once this constraint is reached and the master temperature controller still request a reduction in burner pressure, the operator can block in burners on a multi-burner heater. By doing so the burner pressure required for less burners in service will require a higher burner pressure and should eventually increase the pressure above the minimum.

Split Range Control

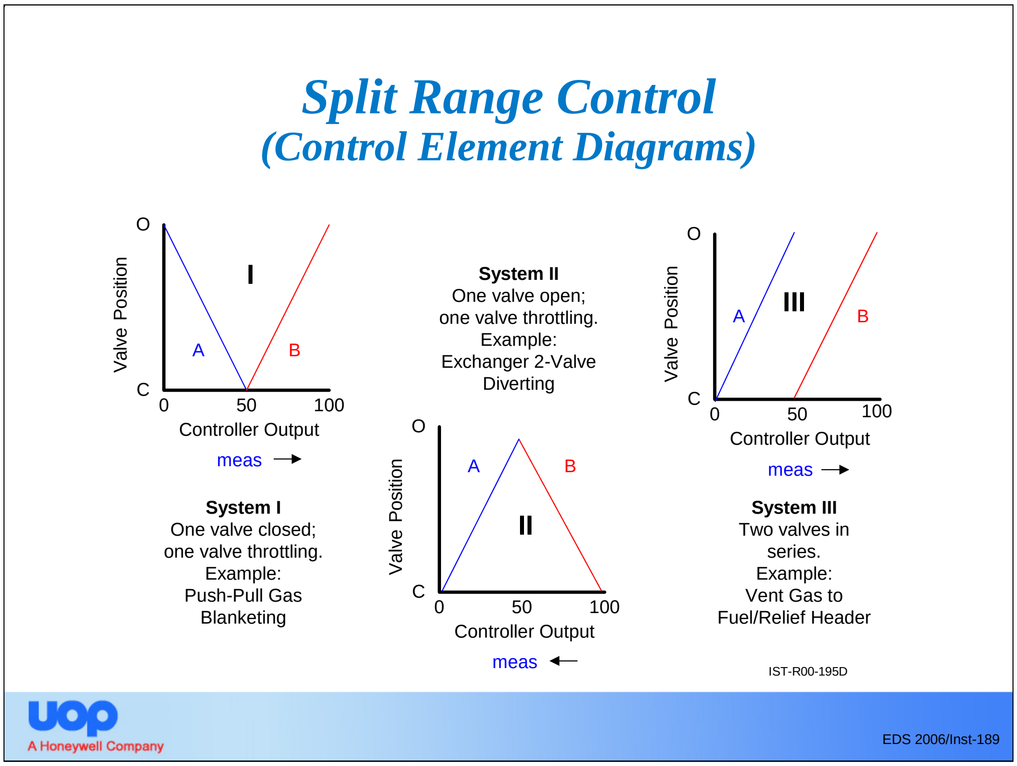
- **Output from Controller manipulates 2 or more control valves independently to control the process**
- **Control Element Diagram depicts operation of control valves over the full range of controller output**
- **Multipliers are incorporated to scale the output signal from 0 - 100% of the control valve stroke**

A Honeywell Company

EDS 2006/Inst-188

In some applications, the process controller can be used to control 2 or more control valves. This application is termed “split-ranging” the control valves. A control element diagram (CED) can be constructed to depict the operation of the controller and manipulation of the control valves (final control elements).

Mathematical multipliers are utilized to scale the output signal over the full range of the control valves.

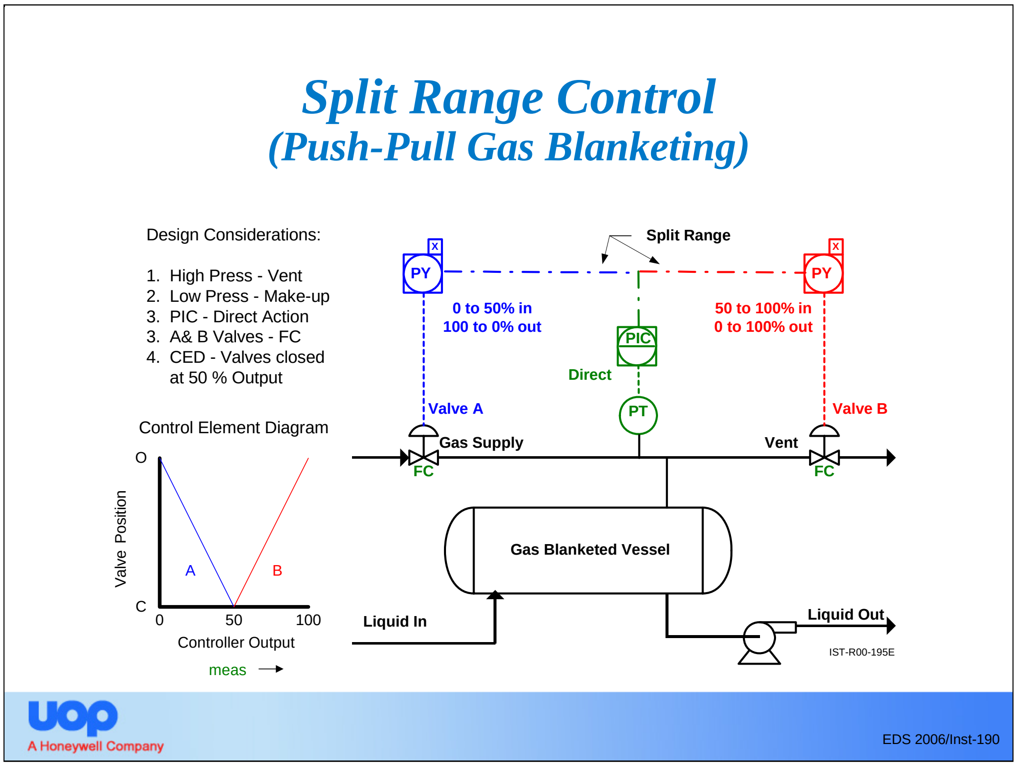


For system I, such as a push-pull gas blanketing system on a feed surge drum, both valves are closed when the controller output signal is 50%. When the controller output signal is greater than 50%, one valve (valve A) remains closed, and the second valve (valve B) is used to throttle one process stream. Likewise when the controller output signal is less than 50% just the opposite scenario exists, valve (valve B) remains closed, and the other valve (valve A) is used to throttle a second process stream.

For system II, such as an exchanger 2-valve diverting system, both valves are open when the controller output signal is 50%. As the controller output increases above 50%, valve B closes while valve A remains open. Just the opposite occurs when the controller decreases below 50%.

For system III, such as vent gas relief system, the valve are operated in series. As shown in the CED diagram, when the controller output is 0%, both valves are in the closed position. As the controller output increases from 0 to 50%, valve A moves from the closed position to the wide-open position while valve B remains closed. From 50 to 100% valve A remains wide-open and valve B is being manipulated from the closed position to its wide-open position.

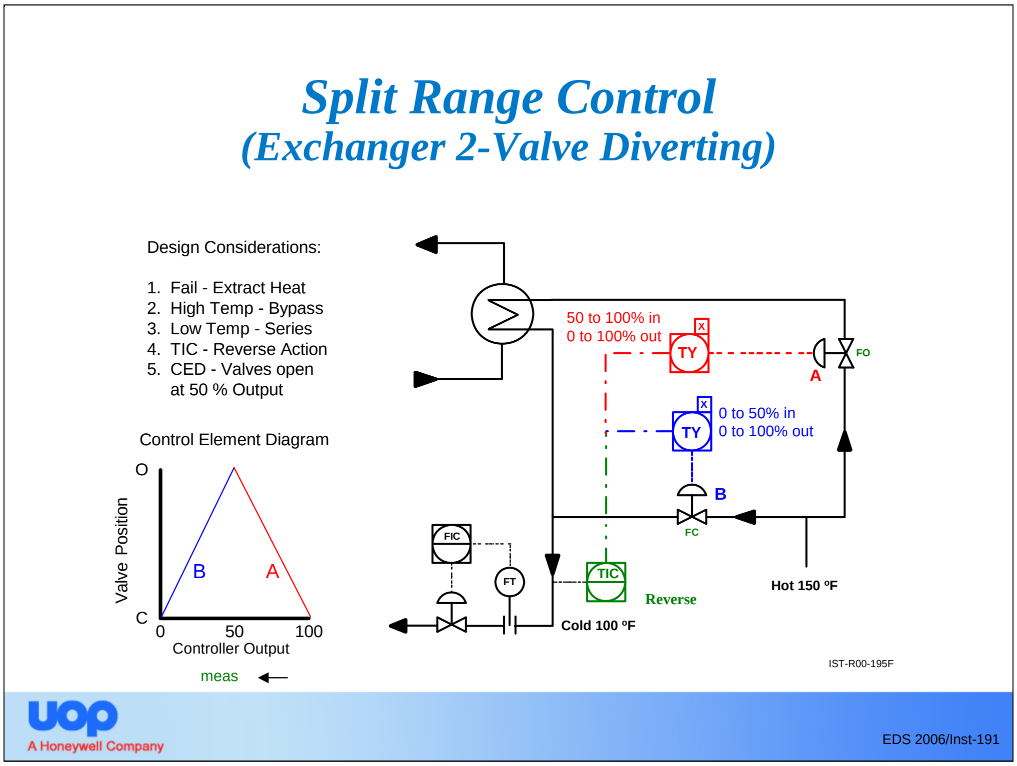
Other systems exist, some with three valves, but these three examples cover the majority of the cases.



This is an example of the Push-Pull Gas Blanketing System. Notice in this case as shown both valves are air to open (fail closed valves). Utilizing the CED for system I, both valves are closed at a controller output equal to 50%. If the pressure in the drum increases above the set point, then the vent valve must be manipulated in order to bring the pressure in the drum back down to the set point. We can accomplish this by making the controller action direct and establishing the fact that valve B will be the vent valve on the CED. Therefore the multiplier has an input of 50% to 100% with a corresponding output of 0 to 100%. Notice that this multiplier is direct acting, i.e. if the input increase then the output increases also, but at a ratio (slope) of 2 to 1.

With the controller action set as direct acting, if the pressure decreases then the output of the controller also decreases. Therefore we need to open the gas supply valve(while the vent valve is closed) to bring the pressure back up to its set point. However the gas supply valve is also an air to open valve. When the controller output is 50% this valve should also be in the closed position. In order to accomplish this task, the multiplier must be reverse acting. In other words when the controller output is 50%, the output of the multiplier needs to be 0% in order for the valve to be in the closed position. Then as the output of the controller decreases below 50%, the signal to the gas supply valve increases towards 100%. The ratio (slope discussed earlier) is -2 to 1.

What would happen if the controller action was set as reverse acting?



In the above Exchanger 2-Valve Diverting System, one valve is in series and one valve bypasses the heat exchanger. The failure mode of these two valves is contingent upon the process. In this case the process is heat extraction; therefore, the series and bypass valves are designed as air to close and air to open, respectively. Utilizing the CED for system II, both valves are open at a controller output equal to 50%. If the temperature of the controller increases above the set point, more heat must be extracted by forcing additional process fluid through the exchanger. This is accomplished by throttling the bypass valve with the series valve wide open.

Looking at the bypass valve (this is a fail close valve) with a controller output at 50%, the valve should be wide-open with a 100% air signal. If the controller is setup as a reverse acting controller, then the bypass valve becomes the B valve on the CED. Therefore the multiplier has an input of 0 to 50% with a corresponding output of 0 to 100% (slope 2 to 1).

If the temperature of the controller decreases below the set point, too much heat has been removed by the heat exchanger. Therefore less fluid should pass through the heat exchanger. This is accomplished by throttling the series valve while the bypass valve is wide-open. At a 50% controller output signal the A valve (series valve) should be wide open. This is accomplished with the output of the multiplier at 0% to a fail open valve. Therefore the multiplier has an input of 50 to 100% with an output of 0 to 100%.

Cascade Control

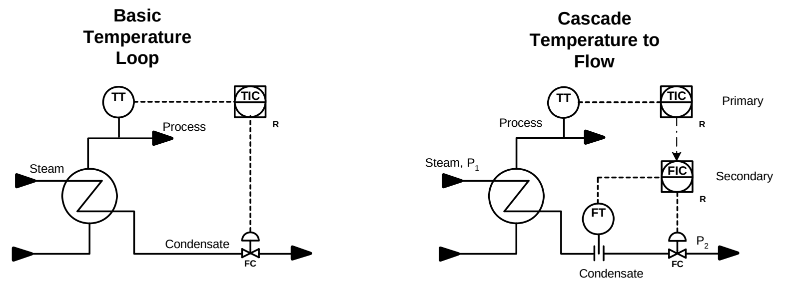
- **Consists of a primary controller and a secondary controller**
- **Output of the primary controller is the set point of the secondary controller**
- **Response of the secondary loop must be much more rapid than the primary loop**
- **Minimizes effects of load variables in the secondary loop before these variables influence the primary loop**

uop A Honeywell Company EDS 2006/Inst-192

In cascade control the output of one controller is used to manipulate the set point of another controller. The two controllers are then said to be cascaded, one upon the other. Each controller has its own process variable input, but only the primary controller has an independent set point and only the secondary controller has an output to the process. The manipulated variable, the secondary controller, an its process variable constitute a closed loop within the primary loop.

The control valve positioner was sited as a secondary loop with the manipulation of the final control element. Temperature, pressure , and level are often cascaded with flow loops to minimize load variables such as line pressure changes. The output of the primary controller is essentially a manipulation of the flow of mass or energy by the primary controller. Cascade control is of great value where high performance is mandatory in the face of random disturbances.

Cascade Control
(Temperature/Flow)



Design Considerations:

1. Basic Temperature Loop will adjust condensate flow for deviation in process temperature
2. Load Variables P_1 and P_2 influence the pressure drop (flow) across the control valve
3. Cascading TIC to FIC minimizes effect of variation in the Load Variables
3. Output from TIC is Set Point for FIC
4. FIC configured for Local/Remote Set Point capability

IST-ROO-197A

uop
A Honeywell Company

EDS 2006/Inst-193

Steam is often used throughout the Process Industry as a heating medium. As shown in the Basic Temperature Loop, the process outlet temperature measurement is the PV to the temperature controller. The manipulated variable is the condensate flow rate through the exchanger. As the flow of condensate varies, the amount of ‘condensate backup’ in the exchanger varies exposing more/less tube heat transfer area. The output of the controller is a function of temperature. If the PV is low more heat is required opening the control valve and visa versa. However what effect does an increase in the steam header pressure have on the process temperature?

If the steam or condensate header pressures vary, then the driving force (valve differential pressure) across the control valve varies; and as a result, the flow of condensate across the valve will change as the differential pressure fluctuates. Therefore the heat transfer surface area is varying the amount of heat transfer and the process outlet temperature may not be controlled very well.

In the cascade loop a flow measurement and flow controller have been added to control the flow of condensate from the exchanger. The output of the primary controller, now the set point to the secondary controller, is a function of condensate flow instead of process outlet temperature. At steady state the temperature controller requires ‘X’ gpm of condensate. If the condensate rate changes, due to a header pressure disturbance, the secondary controller will immediately throttle the valve to maintain the flow of condensate at ‘X’ gpm.

Ratio Control

- **Manipulates material balance to control the ratio of one variable to another to satisfy a defined objective**
- **Ratio Flow Control - Multiplier Function**
  - **Input Variable: Flow Measurement (X)**
  - **Gain: Ratio of Flows - Y/X**
  - **Output = Input times Gain = (X)(Y/X) = Y**
- **Output is required flow of stream Y and is cascaded as set point to flow controller Y**

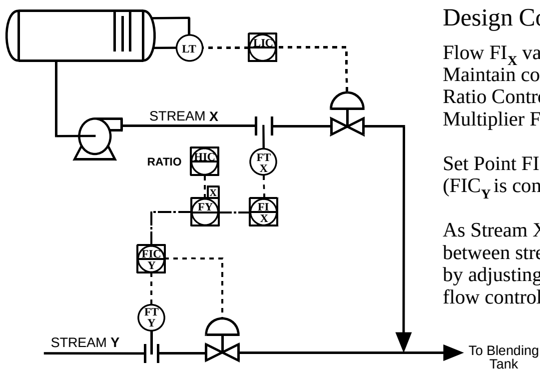
uop
A Honeywell Company

EDS 2006/Inst-194

Ratio control is a feed forward system wherein one process variable is manipulated in a predetermined ratio to another process variable to satisfy some higher-level objective. In blending systems the higher-level objective may be composition, Octane Number, or whatever; but this higher-level objective cannot always be measured, in which case the real objective cannot be controlled by a feedback loop.

The real controlled variable in a ratio control system is the ratio of two measured process variables. The typical system manipulates one valve controlling one of the process variables, while the other is a "wild" uncontrolled process variable. As the uncontrolled variable changes, then the system will manipulate the controlled variable in direct proportion as dictated by the predetermined ratio.

Ratio Control
(Blending Application)



Design Considerations:

Flow FI_X varies as level changes
Maintain constant blend of Streams X & Y
Ratio Control Station HIC adjustable Y/X
Multiplier FY: Input X times Y/X = Y

Set Point FIC_Y is Multiplier FY output
(FIC_Y is configured for local/remote SP)

As Stream X flowrate changes the ratio between streams X and Y is maintained by adjusting the set point to Stream Y's flow controller

IST-R00-197B

A Honeywell Company

EDS 2006/Inst-195

The "wild" process variable is stream X. Stream X's flow rate is contingent upon the level in the drum. The objective of the control scheme is to blend stream X and stream Y in a pre-determined ratio of Y/X. This ratio then is the gain of the FY multiplier block.

The "wild" flow is measured and the PV is an input to the FY multiplier block. With a gain of "Y/X" times the PV "X", the output of the multiplier block is the required set point of stream Y to maintain the ratio of stream Y to stream X.

As the level varies in the drum the level controller will adjust the outflow from the drum (stream X), and the control scheme will continuously control stream Y in direct proportion to the changes induced in stream X by the level controller.

Ratio Control
(Forced Draft Fired Heater)

- **Forced Draft System – air flow to heater is controlled variable**
- **Transient states are always fuel lean/air rich**
- **For an increase in fuel demand, control system will first increase air flow followed by an increase in fuel supply**
- **For a decrease in fuel demand, control system will first decrease fuel supply followed by a decrease in air flow**

A Honeywell Company

EDS 2006/Inst-196

One way to improve the overall efficiency of a fired heater is to design the heater with a forced draft system where the air to the heater is preheated by the hot flue gas exiting the heater. In order to accomplish this air preheat the air flow to the heater is a controlled variable.

As a result of this added complexity, the control system for a natural draft fired heater has been modified, to prevent starving the fire box of oxygen. The control scheme will essentially insure that excess air is available and that transients states are always fuel lean and air rich.

Ratio Control
(Air/Fuel Ratio)

- **Air/Fuel stoichiometric quantities**
  - **Methane - $CH_4 + 2O_2 = 2H_2O + CO_2$**
  - **Butane - $C_4H_{10} + 6.5O_2 = 5H_2O + 4CO_2$**
- **Fired Heater designed for 15% excess air**
- **$Ratio_{WT} = [(1.15M_{O_2}/0.21)MW_{AIR}]/[(M_{FUEL})MW_{FUEL}]$**

FUEL	AIR/FUEL $Ratio_{WT}$
Methane	19.9
Ethane	18.5
Propane	18.0
Butane	17.8

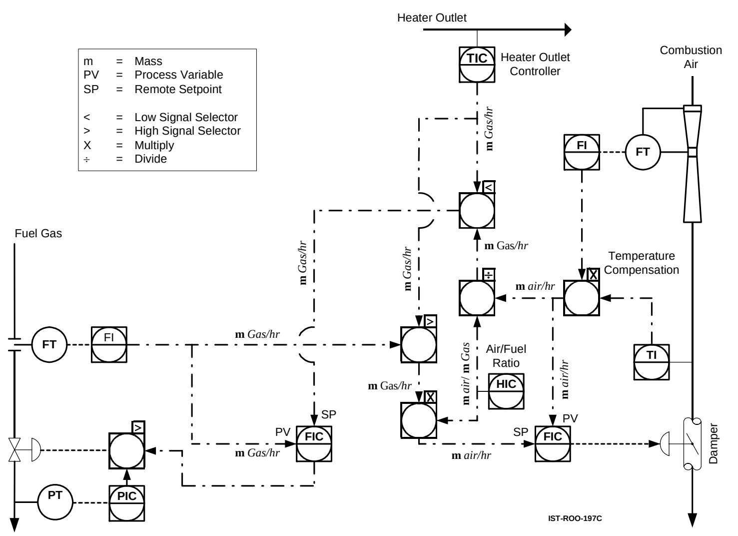
uop A Honeywell Company

EDS 2006/Inst-197

Ratio control is implemented in the control scheme on a weight bases of Air to Fuel. One common fuel is refinery off-gases from the various process units. These fuel gases are mixtures of anything from hydrogen to C4's and C5's. The above examples illustrate the stoichiometric equations for complete combustion of a particular fuel. Based on 15% excess air, the air to fuel ratio varies between 17 and 20 for typical refinery fuel gas. The heavier the fuel the lower the ratio becomes. For liquid fuels this ratio approaches 17.

Ratio Control

(Fired Heater Fuel/Air Cross-Limiting Control System)



EDS 2006/Inst-198

With experience the operator will be able to observe the burner flame pattern and for a typical fuel gas adjust the air to fuel ratio to obtain optimum performance in the fired heater. The above control scheme incorporates a ratio control station with an adjustable air to fuel ratio between 17 and 20, a multiplier block, a divider block, and high/low signal selectors. The engineering units for both the fuel gas and air flow rates are determined on a mass basis.

The primary controller is the TIC on the heater outlet. If the process temperature decreases below the set point, then the TIC will demand an increase in fuel firing. However this demand first passes through a low signal selector. This selector compares the calculated fuel rate at steady state (based on current air flow) with the transient demand fuel rate (based on the TIC output). Initially the low signal selector ignores the increase in demand from the primary controller. The demand signal from the TIC is also an input to a high signal selector. This selector compares the transient demand fuel rate with the actual fuel consumption at steady state (based on actual fuel as measured). The high signal selector passes the fuel demand signal onto the multiplier which in turn increases the air set point.

As illustrated a demand in fuel firing will first increase air and as the air rate increases the fuel rate will follow. Likewise on a decrease in demand from the the TIC, fuel will be reduced first followed by a decrease in the air rate.

Feedforward Control

- **Based on an inferential or empirical process relationships, the manipulated variable is adjusted directly as load variables change**
- **Basic control loop components**
  - **Feedforward model**
    - **steady state model of the process using material and energy balance equations and process measurements**
    - **model predicts changes in manipulated variable as a function of load variable changes**
    - **set point adjustments based on model prediction**

uop
A Honeywell Company
EDS 2006/Inst-199

Processes which cannot be controlled well because of their difficult nature are very susceptible to disturbances from load or set-point changes. A means of solving this type of control problem directly is called feedforward control. The principal factors affecting the process are measured and are used to determine the appropriate change to meet current conditions. One important feature is that the controlled variable is not used by the system. The controlled variable is used as a feedback trim to account for inaccuracies in the model.

When a disturbance is initiated, control action starts immediately to compensate for the disturbance before it changes the controlled variable. Feedforward action theoretically is capable of exact control, its performance being limited only by the accuracy of the measurements and engineering calculations.

In general the feedforward control system continuously balances mass and/or energy delivered to the process against the demands of the load variables. As a result the engineering calculations made by the control system are mass and energy balances around the process, and the manipulated variables must be accurately regulated flows. Therefore the feedforward system will predict transient state responses to the manipulated variables and make these adjustments (as SP changes) before ever seeing any changes in the controlled variable. If the calculations are in error then the predicted state will be in error and a deviation will exist. The better the model, the better the prediction.

Feedforward Control
(continued)

- **Basic control loop components**
  - **Feedback (trim) controller**
    - **corrects for errors in the steady state model and for errors in process measurements**
    - **steady state model usually based on major load variables and does not account for all load variables**
  - **Controlled variable controller**
    - **based on steady state model prediction and correction from feedback trim controller, adjustments to controlled variable set point are implemented**

A Honeywell Company

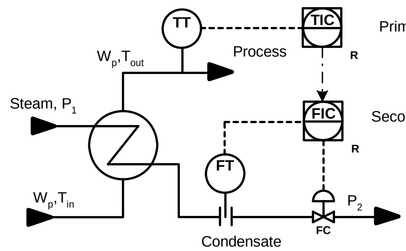
EDS 2006/Inst-200

The feedback trim controller is an integral part of the feed forward model. It's contribution if done properly is minor but will eliminate any offset. The feedback trim controller will correct for errors in the steady state prediction and for any inherent error in the process measurements. Also most feedforward systems will incorporate only the major load variables. Therefore changes in some of the minor load variables will not be accounted for in the feedforward scheme and the steady state model will not be exact.

Combining the steady state model prediction with the feedback trim controller, adjustment are made to the set point of the controlled variable. The feedback trim controller should be slow responding and makeup only the difference that the model has not accounted for.

Feedforward Control

(continued)



Feedforward Model

$HEAT_{in} = HEAT_{out}$

$W_{cond}(H_{vap}) = W_p(C_p)(T_{out}-T_{in})$

$W_{cond} = W_p(C_p/H_{vap})(T_{out}-T_{in})$

Let $K = C_p/H_{vap}$

and T^*_{out} = desired temp

then

$W_{cond} = K(W_p)(T^*_{out}-T_{in})$

Design Considerations:

1. Cascade control improved response of the control system for changes in steam conditions.
2. What happens for major Load Variables on process side? Flowrate or inlet temperature?
3. Steady state model represented by:
   $Heat_{in} = W_{cond}(H_{vap})$ and $Heat_{out} = W_p(C_p)(T_{out}-T_{in})$
   where H_{vap} = Heat of Vaporization for steam
   and C_p = Heat Capacity of Process Material
4. Additional Instruments required to measure process flow and inlet temperature.

IST-ROO-197D

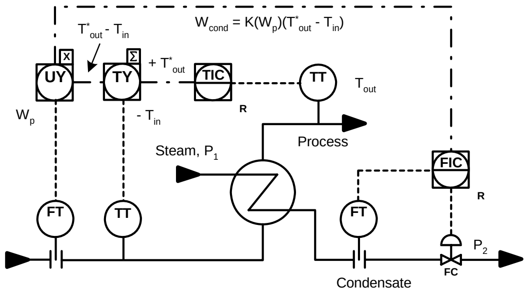
uop
A Honeywell Company

EDS 2006/Inst-201

We implemented the cascade control system above to compensate for load changes on the heating medium side of the exchanger. But what about the load variables on the process side? What happens to the process stream outlet temperature if the process flow rate changes, or the process inlet temperature changes, or stream composition changes? In order to implement the feedforward control system, one needs to identify the major load variables and then implement process measurement of those load variables. In the above example it was identified that the process flow rate and temperature were the major load variables. Therefore additional instrumentation is required to measure both the process flow rate and the process inlet temperature.

The next step is to develop the engineering equations to represent the heat input at steady state. On the steam side, the heat of vaporization needs to be an assumed value as well as the heat capacity of the process fluid. Knowing the heat capacity of the fluid, the process flow rate and the expected temperature rise across the heat exchanger, the anticipated duty can be calculated. Likewise knowing the steam heat of vaporization and this anticipated exchanger duty, the quantity of steam can be calculated. Therefore an orifice plate and temperature element are required on the inlet side of the shell in order to measure these process variables.

Feedforward Control
(continued)



Feedforward Model

$HEAT_{in} = HEAT_{out}$
$W_{cond}(H_{vap}) = W_p(C_p)(T_{out}-T_{in})$
$W_{cond} = W_p(C_p/H_{vap})(T_{out}-T_{in})$
Let $K = C_p/H_{vap}$
and T^*_{out} = desired temp
then
$W_{cond} = K(W_p)(T^*_{out}-T_{in})$

Design Considerations:

1. FIC represents the controlled variable controller with remote set point.
2. With appropriate Eng Units, TY and UY represent the feedforward model at steady state.
3. TIC represents the feedback (trim) controller, compensating for deviations from set point.
4. The required condensate flow is calculated to compensate for changes in the major load variables of process flow and inlet temperature.
5. For inefficiencies in the feedforward model, process measurements, etc, the feedback (trim) controller corrects the desired outlet temperature.

IST-ROO-197E

UOP A Honeywell Company

EDS 2006/Inst-202

With the implementation of the process flow rate and temperature measurements, the exchanger duty can be predicted on the process side and the quantity of steam required can be determined with the assumed heat of vaporization. This value then is the steady state prediction and becomes the set point to the controlled variable controller. However due to temperature or flow measurement errors or errors in the assumed heat of vaporization or heat capacity, the steady state prediction may be in error. Thus the feedback trim controller (TIC) is used to tweak the process and compensate for this error. The output of the feedforward calculation is the set point to the flow controller (controlled variable).

Advanced Process Control Applications

- Feedback Control
- Summers, Multipliers, and Scaling
- Signal Selectors
- Split Range Control
- Cascade Control
- Ratio Control
- Feedforward Control

■ *Distillation Controls*

A Honeywell Company

EDS 2006/Inst-203

This concludes our presentation on advanced process control. This is only a scratch on the surface, but it should provide some insight into the field of Advanced Process Control and how the computer can be utilized for difficult to control processes.

Next we will look at some of the more common distillation control schemes.

Distillation Controls
(continued)

- **Strategies for constant Heat Input Control**
- **Review of Gibbs Phase Rule**
- **Design basis for Column Pressure Control**
- **Alternate designs for Composition Control**
- **Recommended Material Balance Control**

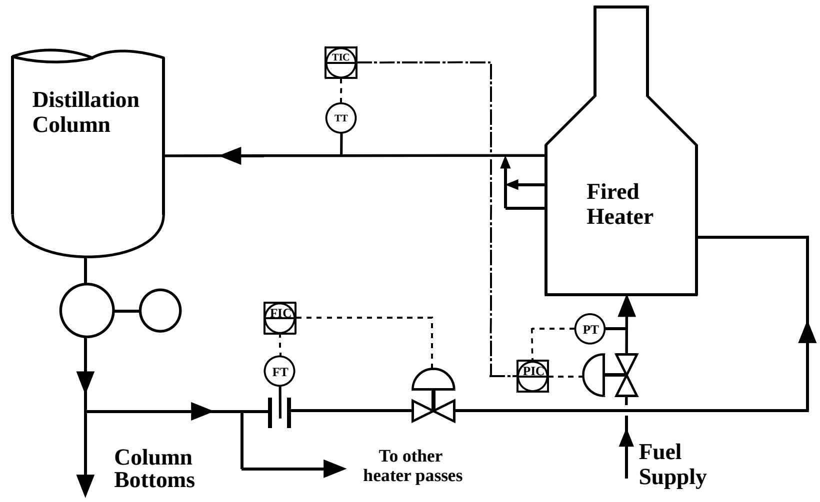
A Honeywell Company

EDS 2006/Inst-204

Initially we will look at a couple of control schemes around the bottom of the column. UOP philosophy centers around constant heat input, but cascaded control schemes or dual composition control schemes are also a viable alternative. Next we will look at Gibbs Phase Rule and apply this principle to multi-component distillation.

After review of the Gibbs Phase Rule we will see that pressure control is one independent parameter that we will fix in order to achieve the desired split. We will also review two alternative control schemes for composition control based on temperature control in the rectifying section of the distillation column. We will also see some of the selection criteria for picking one control scheme over the other.

Distillation Controls – Heat Input
(Fired Reboiler TIC Controller)



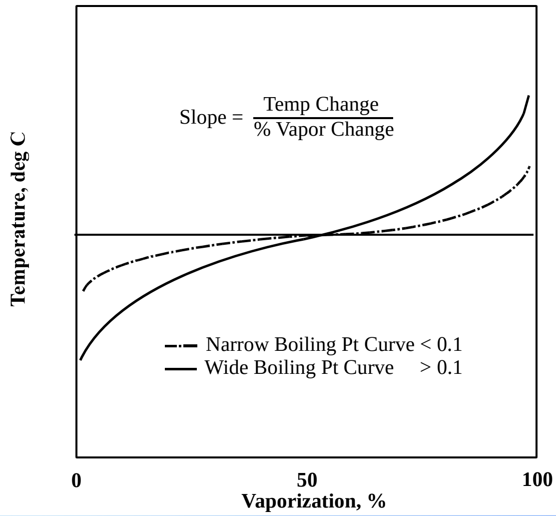
EDS 2006/Inst-205

For the most part UOP will design heat input control schemes for constant heat input. The control scheme listed above is for a multi-pass fired heater. Each individual heater pass will have its own flow control loop to balance the flows through each heater pass. Therefore the process load to the fired heater is constant.

The primary controller, TIC control on the combined heater outlet, is cascaded to the secondary controller, PIC burner pressure controller. The primary controller will adjust the burner pressure up or down to satisfy the temperature set point of the controller.

During normal operation if the column bottom material contains too much of the light key, then the operator will increase the temperatureset point a couple of degrees. This set point change would increase the pressure set point on the PIC, thus increasing the heater firing rate.

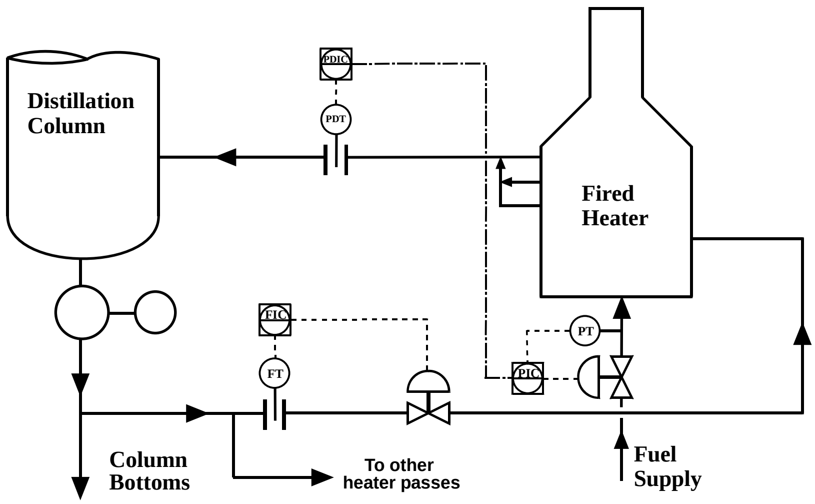
Distillation Controls – Heat Input
(Boiling Point Curve)



What would happen to the temperature profile through the fired heater if the column bottoms material was a pure component or a narrow boiling point material? For a binary distillation, such as a benzene-toluene column in an aromatics complex, the bottom of the column is almost pure toluene and the temperature curve is nearly a fixed temperature as determined by the operating pressure of the column. For a narrow boiling point material, such as the xylene isomers (boiling point range of 277 to 291 °F), the Temperature vs. % Vaporization curve would be a very shallow (flat) curve. If the temperature curve is a relatively flat curve in the design range of the fired heater, then TIC control can not be implemented

For UOP designs, the typical fired heater will be sized to provide somewhere between 40 to 60% vaporization at the heater outlet. Therefore UOP will look at the slope of the boiling point curve in this range. If the slope is greater the 0.1 °C per % vaporization, then TIC control on the heater outlet can be implemented. However if the slope is less than 0.1 °C per % vaporization, then TIC control will not work. Some other alternative must be implemented in order to control the heat input to the distillation column.

Distillation Controls – Heat Input
(Fired Reboiler PDIC Controller)



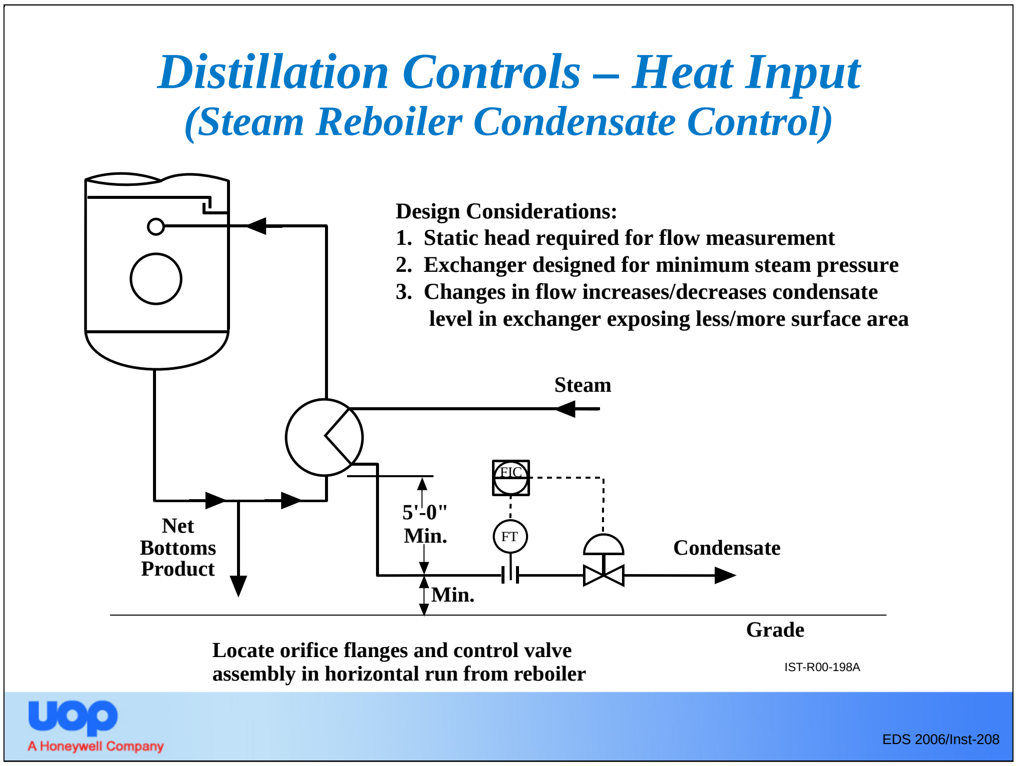
A Honeywell Company

EDS 2006/Inst-207

In the above control scheme the primary TIC control loop has been replaced with a PDIC control loop. This is a pressure differential controller with a pressure differential transmitter (PDT) and pressure differential element (PDE). The PDE is an eccentric orifice plate that must be installed in a horizontal line in order to allow for the passage of the two phase fluid through the orifice plate. An eccentric plate is chosen to ensure that the liquid passes through with the least amount of effort and minimal hold-up.

The theory behind this application involves the relative pressure differential between liquid and vapors generated across the orifice plate. For a fixed liquid flow to the heater (notice that we still have constant heater pass flow rates), the PDIC (ranged 0 to 100% heat input) is a measure of the vapor rate generation in the fired heater. In other words the pressure differential is a measure of the heat input to the distillation column (% vaporization).

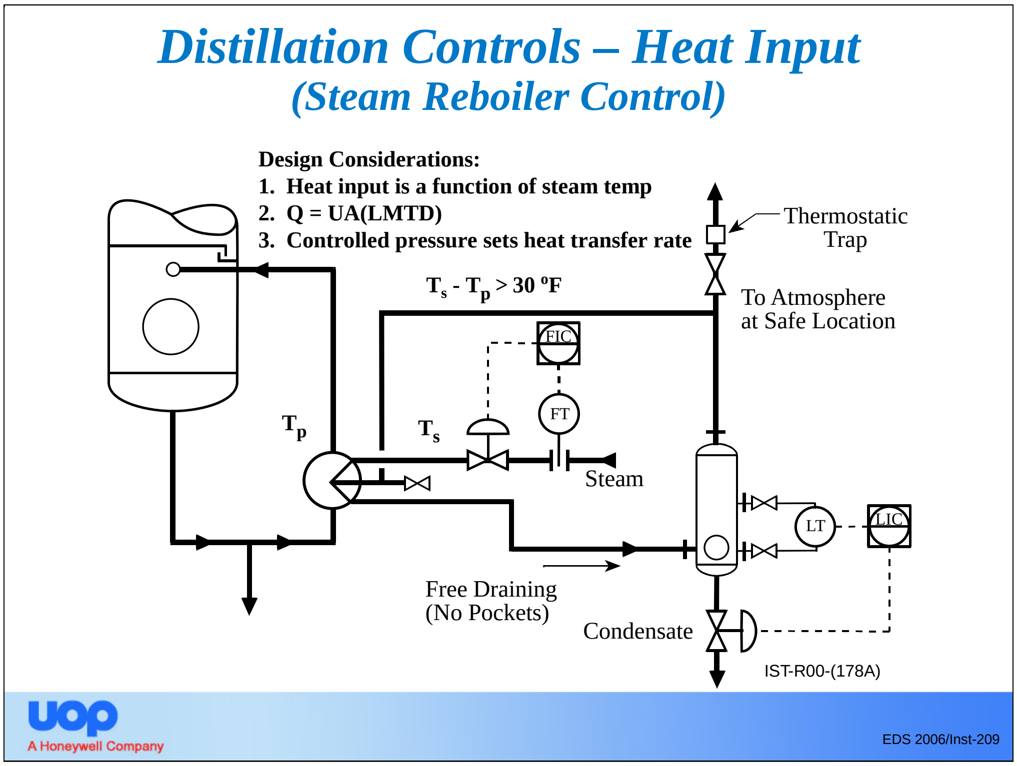
During normal operation if the column bottom material contains too much of the light key, then the operator will increase the set point say from 50 to 55%. Likewise this set point change would increase the pressure set point on the PIC, thus increasing the heater firing rate.



Steam is often used in conjunction with a thermo-siphon reboiler as the heating medium. Two potential control schemes exist to control constant heat input from the reboiler. The first control scheme (shown above) uses a flow control loop on the condensate from the reboiler.

Inside the reboiler a vapor/liquid equilibrium surface layer is established at steady state. The liquid at this interface is at its bubble point as well as the vapor is at its dew point. Care must be taken in the design of the flow element in order to ensure that flashing does not occur across the orifice plate; otherwise, the flow measurement is meaningless. UOP will require a minimum static head upstream of the orifice plate along with using a maximum pressure differential span of 50 inches water column or less. For low or medium pressure steam headers the minimum head requirement is 5 feet.

During normal operation if the column bottom material contains too much of the light key, then the operator will increase the set point to the flow controller. As the condensate flow rate is increased, the vapor/liquid interface layer is lowered in the exchanger exposing more surface area of the tubes for condensation (heat transfer) to occur. Thus the heat input to the reboiler is increased.



The second control system utilizing steam as a heating medium is shown as above. This control scheme employs steam flow control in place of the condensate flow control. Additional equipment is used because the reboiler essentially operates dry utilizing 100% of the exchanger heat transfer surface area and the condensate drains into a collection drum with level control on the drum. The driving force for heat transfer is the log mean temperature difference (LMTD), and the steam temperature is manipulated by controlling the steam pressure in the reboiler.

In order for this type of heat transfer to function properly the steam saturation temperature (Ts) must be greater than the process outlet temperature plus 30 °F.

During normal operation if the column bottom material contains too much of the light key, then the operator will increase the set point to the flow controller. As the steam flow rate is increased, the control valve differential pressure is being reduce; thus, the downstream pressure increases (and as a consequence the saturation temperature in the exchanger rises). The LMTD is increased and with a larger steam flow rate the heat input to the reboiler is increased.

Distillation Controls
(Gibbs Phase Rule)

- **Provides basis for Composition Control**
- **F = C - P + 2**
  - **F = Number of Degrees of Freedom**
  - **C = Number of Components**
  - **P = Number of Equilibrium Phases**
- **Independent variables are temperature, pressure, and composition**
- **For binary system, F = 2 - 2 + 2 = 2**
- **Approximation for multi-components**

uop
A Honeywell Company

EDS 2006/Inst-210

The Gibbs Phase Rules states that the total number of Degrees of Freedom is equal to the total Number of Components in the system minus the Number of Equilibrium Phases plus 2. For the typical distillation column, the independent variables within the boundary of the the column are temperature, pressure, and composition. For a binary system the Number of Components is 2 and the Number of Equilibrium Phases is 2; therefore the number of Degrees of Freedom is 2.

Gibbs Phase Rule states that if two of these independent variables are fixed (controlled), the third independent variable will also be fixed. Therefore if pressure and temperature, pressure and composition, or temperature and composition are controlled, then the third remaining independent variable is fixed.

For most applications the column pressure and a column tray temperature are the two independent variables that will be controlled. Fixing these two variables will fix the composition of the overhead product.

Distillation Controls
(Pressure Control)

- **Gibbs Phase Rule: F = 2**
  - **Controlling pressure fixes one degree of freedom**
  - **One degree of freedom remains**
- **Design Options for OVHD pressure control**
  - **Cooling Medium: Air or Water**
  - **Overhead Composition: Total or Partial Condensation**
  - **Actual magnitude of column pressure**

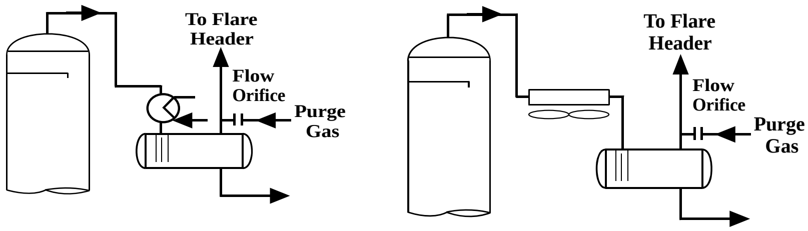
uop
A Honeywell Company

EDS 2006/Inst-211

For multi-component distillation the Gibbs Phase Rule is an approximation and the number of Degrees of Freedom is 2. Pressure control is relatively simple, but is contingent upon several design factors. These design factors include the type of cooling medium, composition of the column overhead material, and the design operating pressure of the column.

These factors will be reviewed and we will investigate UOP's philosophy in reviewing various column designs.

Distillation Controls – Pressure Control
(Total Condensation, P < 5 psig)



Cooling Water Condenser **Air Fan Condenser**

Design Considerations:

1. Receiver designed to float on flare header
2. Condenser located above receiver in either design
3. Purge gas sweeps line to prevent backfow of contaminants
4. Purge gas velocity designed for 0.5 - 1.0 feet/second

IST-R00-199A

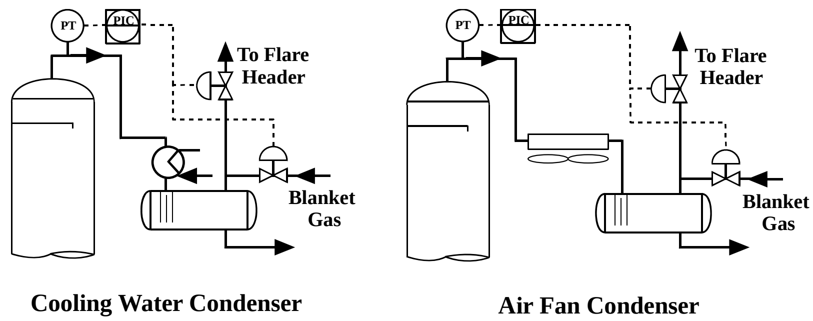
EDS 2006/Inst-212

The first system, illustrated above, has an operating pressure less than 5 psig without any non-condensables in the column overhead material. The receiver/column is designed to float on the flare header; thus operating the column at or near atmospheric pressure. The size of the vent line to the flare header is contingent upon the design of the receiver. A nitrogen purge (via pressure regulator and restriction orifice) is added to the receiver to sweep the vent line and prevent back-flow contamination of material from the flare header into the receiver during transient states throughout the refinery. Once the vent line size is determined, the nitrogen flow rate can be determined based on a design velocity between 0.5 to 1.0 feet/second. The design basis downstream of the restriction orifice is 1 psig.

If a water cooled condenser is in the system design, the exchanger should be elevated above and free draining to the receiver. The non-condensable purge gas must be able to get back to the exchanger; otherwise the exchanger will tend to flood and column pressure will not change until the flooded area changes (causing an increase in rate of condensation).

Distillation Controls – Pressure Control

(Total Condensation, 5 psig < P < 25 psig)



Design Considerations:

1. Condenser located above receiver in either design
2. Split ranged gas blanket/vent system employed to control pressure
3. Maximum blanket gas rate based on rated capacity of overhead pumps
4. Maximum vent gas rate assumed equal to maximum blanket gas rate

IST-R00-200A

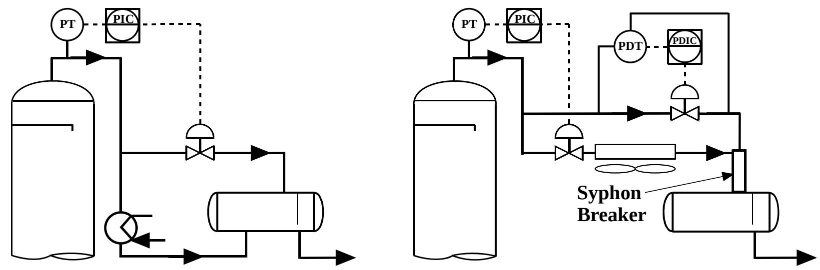
uop A Honeywell Company

EDS 2006/Inst-213

The next system illustrated above is also for a totally condensing system, but with an operating pressure between 5 and 25 psig. A "push-pull" gas blanketing system is employed for column pressures operating within this range. This system (as discussed previously) is a split range system which opens a valve to vent gas to the flare header if the pressure rises above the pressure controller's set point, and opens another valve to bring in make-up gas when the pressure falls below the pressure controller's set point. For instances when the measured pressure and the pressure controller's set point are equal, both the vent valve and the make-up valve will be in the closed position.

The condensers, whether water or air cooled, are located above the receiver for optimal pressure control as discussed in the previous example.

Distillation Controls – Pressure Control
(Total Condensation, P > 25 psig)



Cooling Water Condenser **Air Fan Condenser**

Design Considerations:

1. Condenser located below receiver for cooling water design
2. Increasing vapor rate raises liquid level in exchanger - less heat transfer
3. Pressure valve control pressure on column for both designs
4. Pressure differential valve maintains drum press/temp for air system

IST-R00-201A

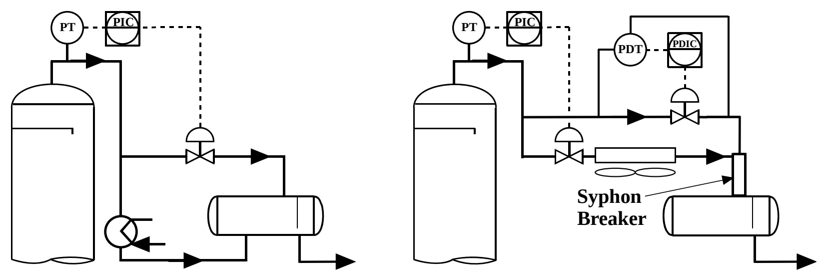
uop A Honeywell Company EDS 2006/Inst-214

For totally condensing systems with operating pressures above 25 psig, the system will differ depending upon whether water-cooled or air-cooled condensers are built into the design.

For a water-cooled system, the design is based on a flooded condenser with the condenser being located below the receiver. The outlet piping from the condenser is routed to the bottom of the receiver and therefore the water condenser is "sealed" in the liquid. The pressure is controlled by bypassing hot vapors around the condenser directly to the overhead receiver. The pressure differential across the control valve is equal to the liquid static head from the receiver vapor/liquid interface to the exchanger interface plus the frictional pressure drop. The receiver and condenser form a "U" tube arrangement. As the hot vapor bypass valve opens the pressure in the receiver will increase. The increase in receiver pressure forces liquid out of the receiver back into the condenser flooding more tubes. With additional tubes flooded, less surface area is available for condensation; therefore, the pressure in the column will rise. Alternately with the hot vapor bypass control valve closing, the column pressure pushes liquid out of the exchanger into the receiver exposing more surface area for condensation to occur. Therefore the column pressure will drop.

The hot vapor bypass valve is sized for about 6% of the overhead vapor rate (as a maximum)and specified with an equal percentage characteristic trim.

Distillation Controls – Pressure Control
(Total Condensation, P > 25 psig continued)



Cooling Water Condenser

Air Fan Condenser

Design Considerations:

1. Condenser located below receiver for cooling water design
2. Increasing vapor rate raises liquid level in exchanger - less heat transfer
3. Pressure valve control pressure on column for both designs
4. Pressure differential valve maintains drum press/temp for air system

IST-R00-201A

UOP
A Honeywell Company

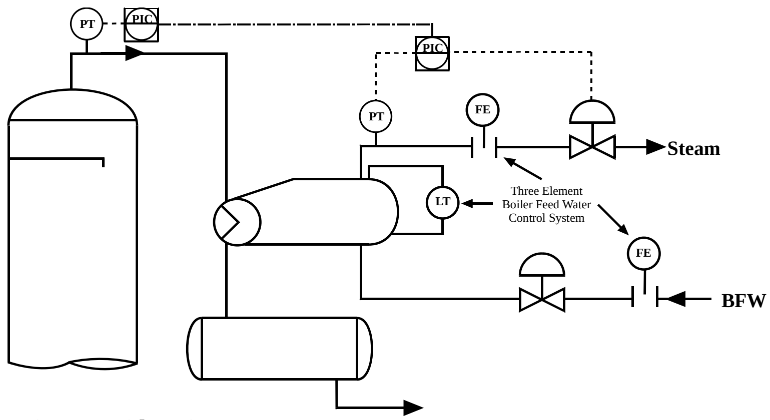
EDS 2006/Inst-215

For an air-cooled system installing the air condenser below the receiver is not practical and; and therefore the condenser is not operated as explained for the water-cooled system. Column pressure is maintained by installing a pressure control valve upstream of the condenser to throttle the overhead vapor and control the column pressure directly.

However with this in mind the receiver's pressure/temperature relationship is a function of the available cooling in the condensers at any given time. Since the pressure in the receiver is essentially the vapor pressure of the condensing fluid (total condensation system); and as the amount of heat extraction varies, the reflux temperature and pump suction pressure will vary. This will lead to very unstable control of overhead composition as ambient conditions change between winter/summer, night/day, sunny/stormy conditions, etc. Not only is the reflux temperature affected, changes in receiver pressure affect the pressure differential across the control valve. As the temperature drops (and consequently the pressure drops), the discharge pressure on the pumps for a given flow drops. Thus the control valve inlet pressure and differential pressure are lower and for a given reflux rate the valve must respond by opening more and more to compensate for the loss in driving force.

Installing the hot vapor bypass control valve will maintain the heat balance in in the system and maintain the receiver at a constant pressure and temperature. The hot vapor bypass valve is sized for about 6% of the overhead vapor rate (as a maximum)and specified with an equal percentage characteristic trim.

Distillation Controls – Pressure Control
(Total Condensation, Steam Generation)



Design Considerations:

1. OVHD product must have sufficiently high boiling point to produce steam
2. Cascade OVHD pressure controller to steam generator pressure controller
3. OVHD pressure controller should be reverse acting controller
4. Increasing the steam generator pressure controller set point, decreases heat transfer rate

IST-R00-202A

A Honeywell Company

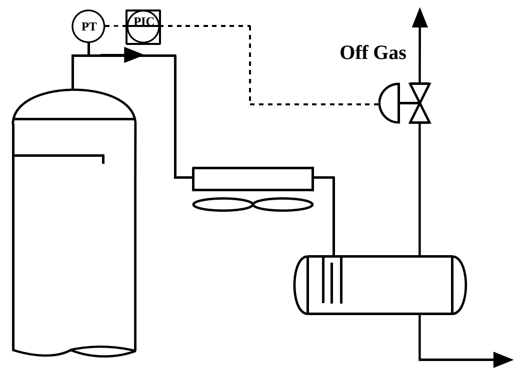
EDS 2006/Inst-216

For some distillation columns heat integration with the refinery-wide steam system may be a feasible alternative to generate low to medium pressure steam. The boiling point of the overhead material must be at a sufficiently high enough temperature (usually 300 °F or higher) to generate low or medium pressure steam.

A kettle type condenser is used in the overhead of the column and a pressure to pressure cascade control scheme is utilized to control the pressure in the column. The column pressure controller is the primary controller in the cascade, while the steam pressure controller is the secondary controller.

If pressure in the column drops, indicating a reduction in overhead vapor rate, the set point to the steam pressure controller increases. An increase in the pressure in the steam drum increases the steam temperature. This in effect reduces the LMDT across the exchanger reducing the heat transfer rate; and therefore reduces the condensation rate on the process side leading to an increase in column pressure.

Distillation Controls – Pressure Control
(Partial Condensation)



Design Considerations:

1. For partial condensation, locate pressure control valve in off-gas line
2. Locate transmitter near column top when using composition control
3. Transmitter may be located on receiver if stripping column
4. Water condensers may be located above or below receiver

IST-R00-203A

A Honeywell Company

EDS 2006/Inst-217

For distillation processes that have non-condensables in the overhead, a vent gas line is required to route the non-condensables out of the system. The destination can be the refinery fuel gas header, flare system, storage, etc, depending upon the quality and value of the product. The control valve is installed in the vent gas line and the pressure transmitter can be installed either at the top of the column or on the overhead receiver. If composition control is a requirement, then the pressure transmitter should be installed near the top of the column for optimum control of the column products. However if the column design is for stripping then the transmitter can be installed at or near the receiver for easy accessibility.

The overhead condenser can be either water cooled or air cooled. If water cooled, the condenser can be located either above or below the receiver.

Distillation Controls
(Composition Control)

- **With F = 2, pressure control fixes 1 degree of freedom; 1 degree of freedom remains**
- **Remaining independent variable are Temperature (T) and Composition (C)**
- **Fixing either T or C fixes the last variable**
- **For most applications temperature control is adequate for control of product composition**
- **Alternatively on-line analyzers could be used for composition control directly**

A Honeywell Company

EDS 2006/Inst-218

Pressure control is relatively easy to implement on the overhead of a column. We have reviewed several applications over a wide range of operating conditions and included total or partial condensation. In terms of Gibbs Phase Rule (and assuming the approximation of a binary system for most multi-component applications) fixing the column pressure defines one of the two degrees of freedom.

Temperature and composition are the remaining degrees of freedom. Therefore if we can control temperature, then composition will be fixed; or if we can control composition then the temperature will be fixed.

For most applications temperature control is adequate for control of product composition. Several variations of temperature control can be implemented depending upon the difficulty of the separation. However in some instances on-line gas/liquid chromatographs could be used for composition control directly.

These are some of the more common means of temperature control from the simplest to some more complicated systems. The higher the degree of separation difficulty, the more sophisticated the system.

For stripping columns a simple end point determination of the overhead product can be achieved by locating the temperature measurement in the overhead piping.

For the more traditional composition control, the temperature measurement is located on one of the column trays above the feed tray. The temperature measurement is then utilized to control the split between the light and heavy key components in the top of the column.

For the truly binary separation processes, especially when column design pressures are near atmospheric, a temperature control scheme can be implemented as a means for pressure compensation.

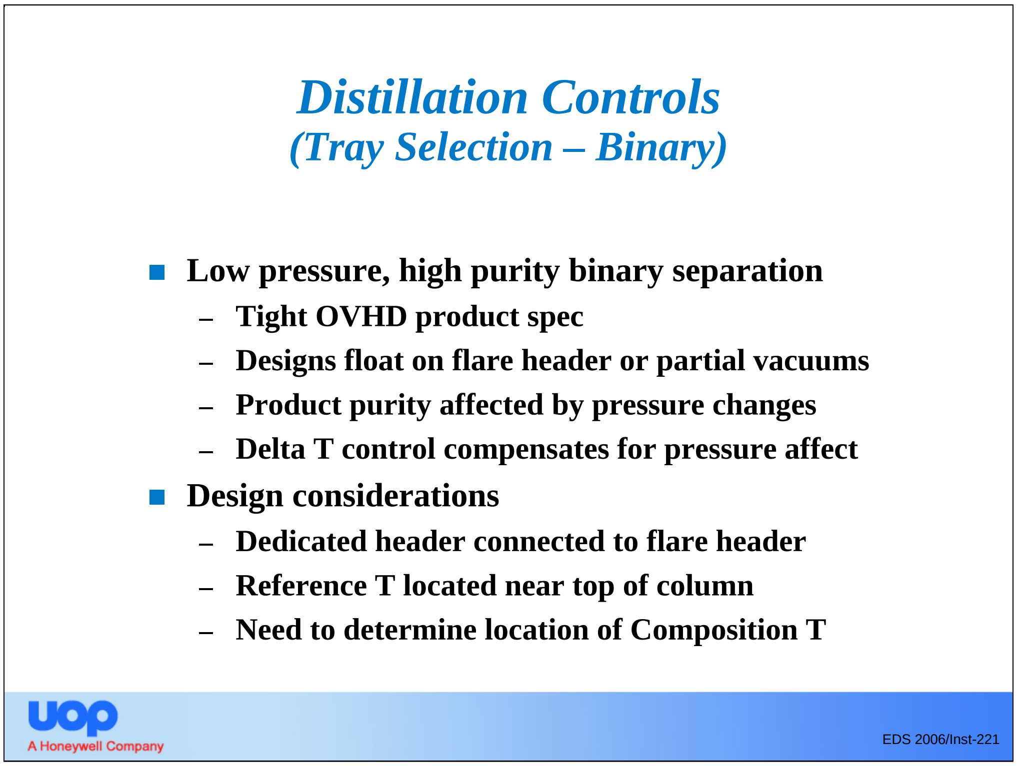
We can go one step further and model the composition and temperature as a function of pressure and implement a control system that utilizes temperature control with pressure compensation as a function of composition.

As the control schemes become more complicated, better VLE data and computer simulations play an even more important part in developing the control scheme.

Distillation Controls
(Tray Selection – multi-component)

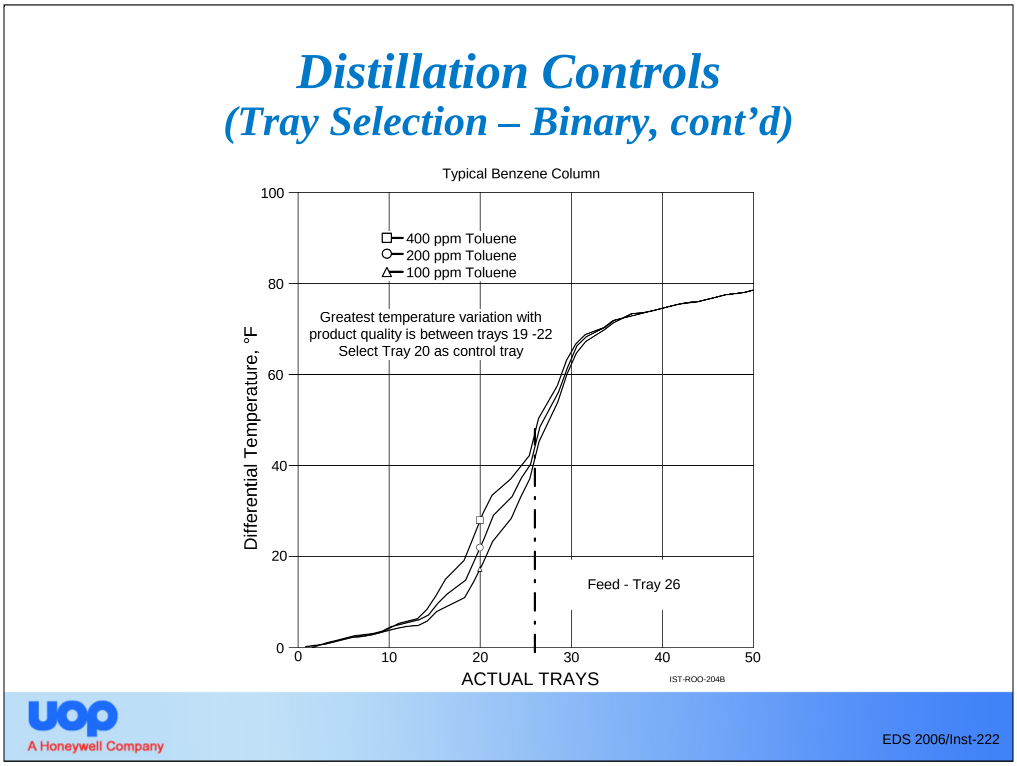
- **Acceptable to use tray half way between feed tray and reflux tray**
- **Computer tray to tray simulations**
  - **Product Quality vs. Tray Temperature**
  - **Affects of Feed Rate or Feed Composition**
  - **Tray with least variation in product quality at constant temperature**
- **Installation of spare thermowells above and below control tray selected (optional)**

Tray to Tray simulations will provide a means of predicting product quality versus tray temperature as a function of various process variables such as column feed rate and/or feed composition. The tray which shows the least variation in product quality at a constant temperature is determined. If uncertainty still exists, then spare thermowells can be located above and/or below the control tray selected.



In an Aromatics Complex with BTX operation (Benzene, Toluene, and Xylene recovery), benzene/toluene separation is a common example of binary distillation at low pressure. The benzene spec is typically tight allowing for only PPM levels of toluene in the overhead product. UOP typically designs the overhead system to float on a dedicated header to the flare knockout drum.

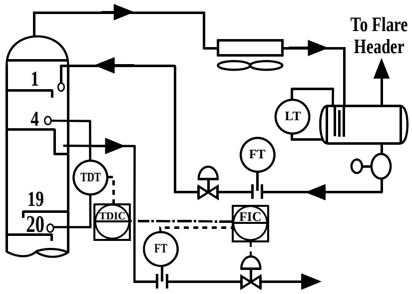
Slight variations in header pressure affect tray temperatures and ultimately product purity. A differential temperature control scheme is used to control overhead composition. One temperature measurement, often referred to as the reference temperature measurement, is located near the top tray of the column. The second temperature measurement is located 'x' number of trays down the column and is often referred to as the composition temperature measurement. Since the system is basically a binary system the top of the column is primarily pure benzene. Further down the column on tray 'x', is a mixture of benzene and toluene. As the composition changes at constant pressure the temperature at the top of the column is unaffected, but the temperature at the composition tray will vary as the composition varies. Therefore if the composition is changing at constant pressure the differential temperature is changing in proportion to the composition changes. However if the pressure is not constant, the reference temperature measurement will change with respect to benzene's vapor pressure. Likewise on the composition tray, the temperature will change similarly. Therefore the differential temperature will not change with respect to changes in pressure at constant composition.

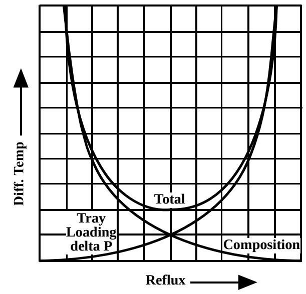


In the above example computer simulations were run for the benzene/toluene system and the differential temperature between the reference tray and various trays in the column was plotted as a function of the actual tray. The family of curves represent simulations at various toluene impurity levels.

From the simulations the greatest temperature variation with product quality is seen between trays 19 and 22. Therefore for the binary separation, tray 20 was selected as the composition temperature measurement for this design.

Distillation Controls
(Tray Selection – Binary, cont'd)





Design Considerations:

1. Reference point located in top of column (pure component)
2. Control point located where greatest T variation with product quality
3. Changes in T due to pressure swings cancel with delta T control
4. Increasing reflux changes composition decreasing tray temperature; but increases tray loading/tray pressure drop with increasing tray temperature
5. Low loadings - Composition governs; high loadings tray delta P governs
6. Tray temp is sum of two effects with minimum delta T attainable

IST-R00-205A

uop
A Honeywell Company

EDS 2006/Inst-223

The differential temperature control scheme is shown above with the reference tray at tray 4 and the composition tray at tray 20. The above plot illustrates that at low tray loadings (small reflux rates) the differential temperature is at its greatest. As reflux rates are increased less and less toluene will exist in the upper section of the column and the differential temperature should diminish. However with an increase in reflux rate, tray loading is increased (an increase in tray pressure drop) and ultimately affects tray temperature with the increase in pressure on any given tray.

Therefore at low tray loading, composition changes govern the differential temperature; but at high loading tray, tray differential pressure governs. As can be seen in the above plot, the differential pressure swings through a minimum with increasing reflux rate. Normal operation must be to the left of this minimum. Feedback control will not work to the right of the curve due to the change in slope of temperature differential vs. reflux flow.

Distillation Controls
(Traditional Composition Control)

- **Traditional Composition Control**
  - **Temperature control adjusts reflux rate**
  - **Receiver level control adjusts distillate rate**
    - **Preferred if Reflux is smaller than Distillate**
    - **Preferred if Reflux is 10 times greater than Distillate**
    - **Preferred if receiver used as surge for downstream units**
  - **Good response for changes in feed composition**
  - **Poor response to external disturbances to heat balance (i. e., rain storm or reboiler upsets)**

UOP A Honeywell Company

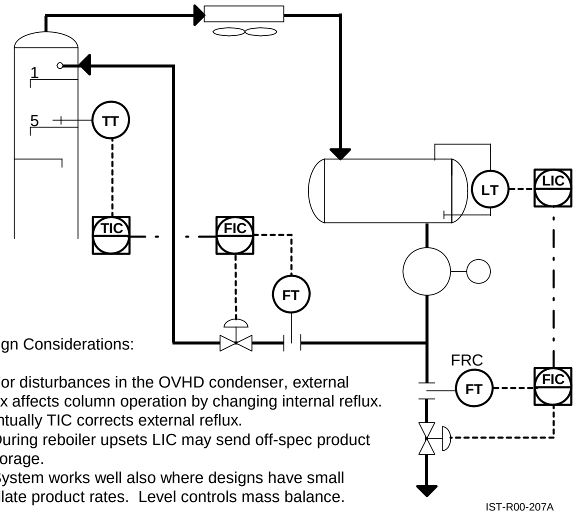
EDS 2006/Inst-224

UOP uses two distinct control schemes for composition control. The first is called Traditional Composition Control. For the traditional composition control, the temperature controller will be aligned with the reflux rate and the receiver level control will be aligned with the distillate draw-off rate.

This control scheme is preferred if the reflux to distillate rate is less than 1 or greater than 10. This is also preferred when the receiver is used as a surge drum for downstream units.

In the event of frequent feed composition changes this control scheme will respond well. However the control scheme responds poorly to external disturbances to the heat balance.

Distillation Controls
(Traditional Composition Control, cont'd)



Design Considerations:

1. For disturbances in the OVHD condenser, external reflux affects column operation by changing internal reflux. Eventually TIC corrects external reflux.
2. During reboiler upsets LIC may send off-spec product to storage.
3. System works well also where designs have small distillate product rates. Level controls mass balance.

UOP A Honeywell Company

EDS 2006/Inst-225

For a disturbance in reflux temperature, such as a sudden rain storm when air-cooled condensers are used, the decrease in external reflux temperature will immediately change the column's internal reflux rate. The external reflux rate will remain constant until the control tray temperature detects a change in composition, but by this time the column is out of heat balance.

For a disturbance in the reboiler, such as an increase in steam pressure, the vapor rate up the column will increase ultimately increasing the level in the receiver. The traditional control scheme will send this increase in inventory to storage (this additional material may well be an impurity in the product and the potential for off-spec product is high).

Distillation Controls
(Traditional Composition Control, cont'd)

- **Additional instrumentation and calculations required for internal reflux control**
  - **OVHD Temperature**
  - **Reflux Temperature**
- **Calculate internal reflux from heat and mass balance around top tray**
- **Calculated internal reflux becomes the corrected process variable and external reflux is adjusted accordingly**

UOP A Honeywell Company EDS 2006/Inst-226

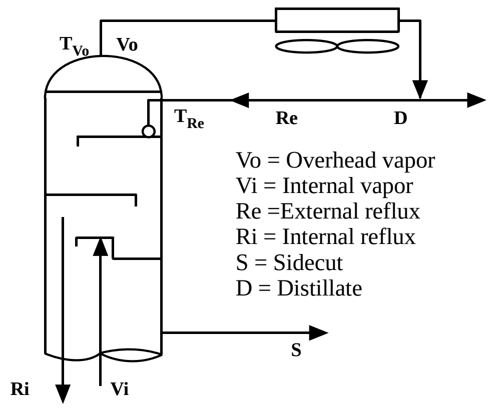
To avoid these types of upsets with the traditional system, the internal reflux must be calculated from a mass and heat balance around the top tray of the column. The calculated internal reflux becomes the corrected process variable and the external reflux is adjusted to manipulated the internal reflux. In order to calculated the internal reflux, additional instrumentation is required to measure both the temperature of the overhead vapor and the temperature of the external reflux stream.

Distillation Controls
(Traditional Composition Control, cont'd)

Internal Reflux Calculations

Material Balance:
Vi + Re = Vo + Ri + S
Energy Balance:
(Hvi)Vi +(Hre)Re =
(Hvo)Vo +(Hri)(Ri + S)

Solve for Ri:
Ri = (1 + KdT) -S
where:
K = (Cp/dHvap)
$dT = T_{Re} - T_{Vo}$



uop
A Honeywell Company
EDS 2006/Inst-227

To initiate internal reflux control additional instruments are required. Both the external reflux temperature and the overhead vapor temperature is required along with a measurement of the external reflux rate. For the top tray, a material and energy balance leads to calculating the internal reflux of the column. This is used as the process variable to an internal reflux controller. The internal reflux controller manipulates the external reflux flow, thus maintaining the internal reflux flow at its set point.

As described earlier in our discussion on feed-forward applications, the calculations above depend upon the assumed value of K. This is the ratio of the heat capacity of the external reflux and the heat of vaporization of the liquid on the first tray. The better the approximation the better the modeling will be for the internal reflux calculation.

Distillation Controls
(Material Balance Control)

- **Material Balance Control**
  - **Temperature control adjusts distillate rate**
  - **Receiver level control adjusts reflux rate**
    - **Preferred if Reflux is larger than Distillate and less than 10 times Distillate**
    - **Not preferred if receiver is used as surge for downstream units**
  - **Level is a measure of Reflux and Distillate rates**
  - **Good response for changes in feed composition and external disturbances to heat balance**

uop
A Honeywell Company

EDS 2006/Inst-228

The second composition control scheme is known as Material Balance Control. In this control scheme temperature control is aligned with the distillate draw-off rate and the receiver level is aligned with the reflux rate. This control scheme is preferred when the Reflux to Distillate rate is greater than 1 and less than 10.

In this control scheme the reflux rate is not measured directly. Instead the level measurement is a measure of both the reflux and distillate rates. Therefore when the composition controller makes adjustments to the distillate rate, the incremental change is immediately compensated for by making a corresponding adjustment in the reflux flow.

As we saw in the traditional control scheme, the material balance control scheme responds well to changes in feed composition and responds exceptionally well to external disturbances in the column heat balance.

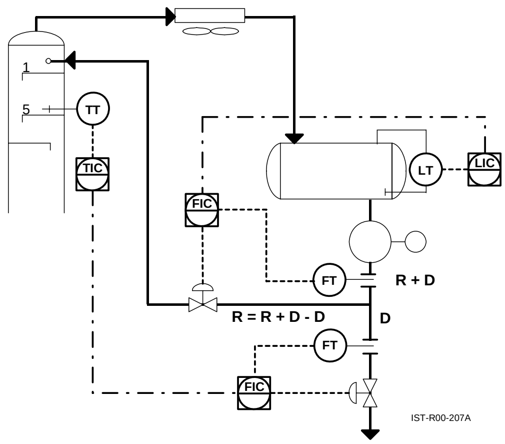
Distillation Controls
(Material Balance Control, cont'd)

Design Considerations:

1. Material Balance control (as it is named) because the temperature controller controls the on-spec distillate in the OVHD and by material balance out the bottom of the column on level control.

2. Better response to both feed composition changes and heat balance changes.

3. Composition controller changes product flow: and because the reflux controller is a function of (R +D), the reflux flow is afftected. As soon as the distillate flow changes, the total OVHD liquid flow changes and the reflux flow is changed simutaneously. This preserves the column heat balance and thus is self-correcting.



A Honeywell Company

EDS 2006/Inst-229

For the traditional control scheme we saw that for an upset in the reboiler, the increase in vapor rate in the column translated into an increase in receiver level and ultimately an increase in product draw-off (good or bad product). However in the material balance control scheme an increase in level is returned to the column as reflux. If the upset is severe and the column is severely upset the composition tray temperature will increase and ultimately the composition controller will adjust the distillate rate downward. In the extreme the column will automatically be controlled on total reflux until the composition profile throughout the column is reestablished.

Distillation Controls

- Review of Gibbs Phase Rule
- Strategies for constant Heat Input Control
- Design basis for Column Pressure Control
- Alternate designs for Composition Control
- Recommended Material Balance Control

A Honeywell Company

EDS 2006/Inst-230

This concludes the review of distillation control and the various control schemes utilized by UOP in our typical designs.

If there are any questions about my presentation, I have a few minutes to field a couple of them from the audience.