PIDStandardNotation

Title: Piping and Instrumentation Diagram Standard Notation

Authors: Halley Crast, Andrew Laskowski, Maurice Telesford, Emily Yatch

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Stewards: Ji Sun Sunny Choi, Sang Lee, Jennifer Gehle, Brian Murray, Razili Stanke-Koch

Last Updated: December 14, 2009

By: Stephanus Oscar, Kaitlin Harrington , Suhendra Lie

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Introduction

Piping and Instrumentation Diagrams (P&IDs) use specific symbols to show the connectivity of equipment, sensors, and valves in a control system. These symbols can represent actuators, sensors, and controllers and may be apparent in most, if not all, system diagrams. P&IDs provide more detail than a process flow diagram with the exception of the parameters, i.e. temperature, pressure, and flow values. "Process equipment, valves, instruments and pipe lines are tagged with unique identification codes, set up according to their size, material fluid contents, method of connection (screwed, flanged, etc.) and the status (Valves - Normally Closed, Normally Open)."[1] These two diagrams can be used to connect the parameters with the control system to develop a complete working process. The standard notation, varying from letters to figures, is important for engineers to understand because it a common language used for discussing plants in the industrial world.

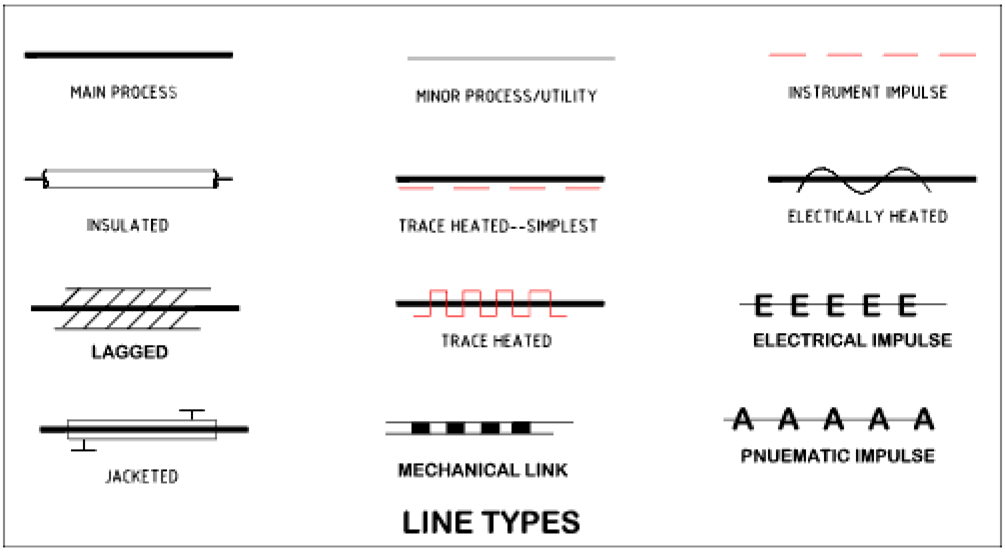
P&IDs can be created by hand or computer. Common programs, for both PC and Mac, that create P&IDs include Microsoft Visio (PC) and OmniGraffle (Mac). As with other P&IDs, these programs do not show the actual size and position of the equipment, sensors and valves, but rather provide a relative positions. These programs are beneficial to produce clean and neat P&IDs that can be stored and viewed electronically. See below for P&ID templates for these programs.

This section covers four main types of nomenclature. The first section describes the use of lines to describe process connectivity. The second section describes letters used to identify control devices in a process. The third section describes actuators, which are devices that directly control the process. The final section describes the sensors/transmitters that measure parameters in a system.

Line Symbols

Line symbols are used to describe connectivity between different units in a controlled system. The table describes the most common lines.

Table 1: Line Symbols



In Table 1, the "main process" refers to a pipe carrying a chemical. "Insulated" is straightforward, showing that the pipe has insulation. "Trace heated" shows that the pipe has wiring wrapped around it to keep the contents heated. "Lagged" indicates on a

P&ID that the pipe is wrapped in a cloth or fiberglass wrap as an alternative to painting to improve the appearance of the pipe see here for more information. The last column in Table 1 shows pipes that are controlled by a controller. "Electrical impulse" shows that the manner in which information is sent from the controller to the the pipe is by an electrical signal, whereas "pneumatic impulse" indicates information sent by a gas.

In addition to line symbols, there are also line labels that are short codes that convey further properties of that line. These short codes consist of: diameter of pipe, service, material, and insulation. The diameter of the pipe is presented in inches. The service is what is being carried in the pipe, and is usually the major component in the stream. The material tells you what the that section of pipe is made out of. Examples are CS for carbon steel or SS for stainless steel. Finally a 'Y' designates a line with insulation and an 'N' designates one without it. Examples of line short codes on a P&ID are found below in Figure A.

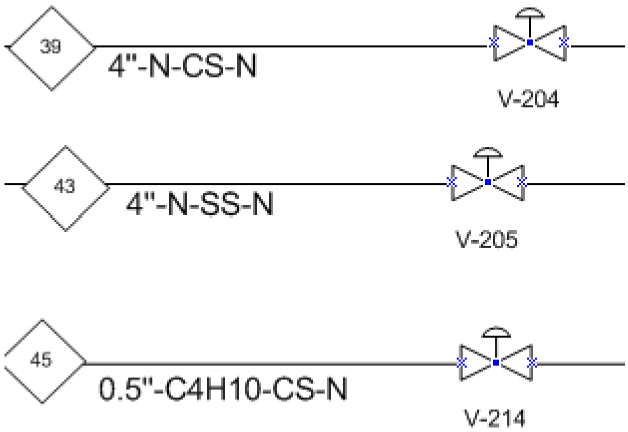


Figure A: Line Labels

This is useful for providing you more practical information on a given pipe segment.

For example in stream 39 in Figure A, the pipe has a 4" diameter, services/carries the chemical denoted 'N', is made of carbon steel, and has no insulation.

Identification Letters

The following letters are used to describe the control devices involved in a process. Each device is labeled with two letters. The first letter describes the parameter the device is intended to control. The second letter describes the type of control device.

Table 2: First Identification Letter

First Letter	Parameter Controlled
A	Analysis
C	Conductivity
D	Density
E	Voltage
F	Flow rate
I	Current
L	Level
M	Moisture (Humidity)
P	Pressure or Vacuum
T	Temperature
V	Viscosity

Table 3: Second Identification Letter

Second Letter	Type of Control Device
A	Alarm
C	Control
I	Indicate
T	Transmit
V	Valve

For example, the symbol "PI," is a "pressure indicator."

Valve Symbols

The following symbols are used to represent valves and valve actuators in a chemical engineering process. Actuators are the mechanisms that activate process control equipment.

Table 4: Valve Symbols

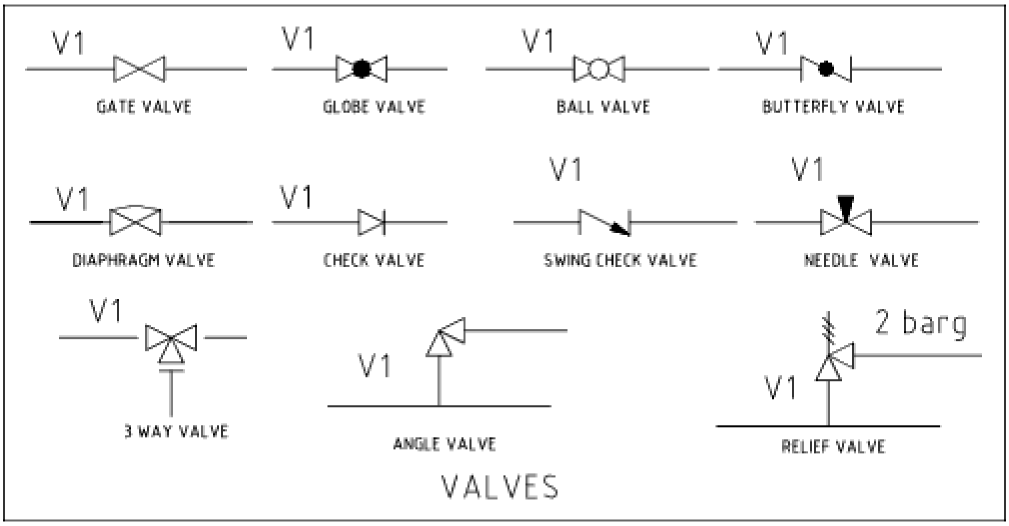
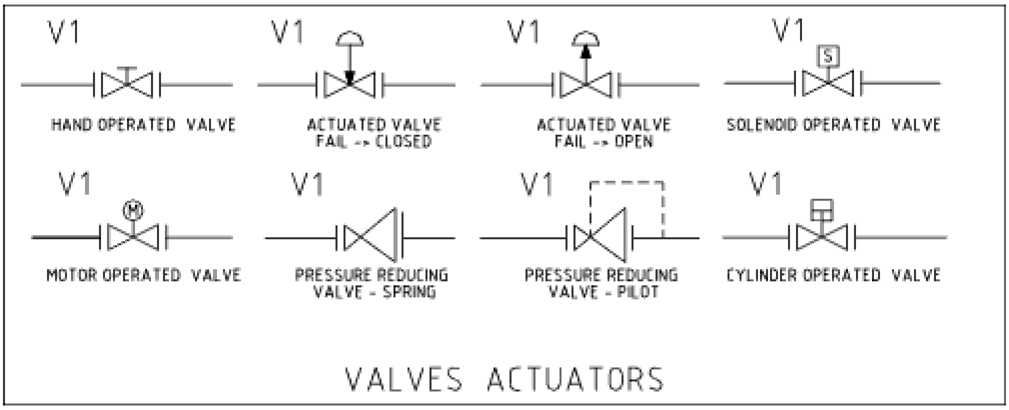


Table 5: Valve Actuator Symbols



General Instrument or Function Symbols

Instruments can have various locations, accessibilities, and functionalities in the field for certain processes. It is important to describe this clearly in a P&ID. Below is a table of these symbols commonly used in P&IDs.

General instrument or function symbols			
	Primary location accessible to operator	Field mounted	Auxiliary location accessible to operator
Discrete instruments	1	2	3
Shared display, shared control	4	5	6
Computer function	7	8	9
Programmable logic control	10	11	12

1. Symbol size may vary according to the user's needs and the type of document.
2. Abbreviations of the user's choice may be used when necessary to specify location.
3. Inaccessible (behind the panel) devices may be depicted using the same symbol but with a dashed horizontal bar.

Source: Control Engineering with data from ISA S5.1 standard

Discrete instruments are instruments separate or detached from other instruments in a process. Shared display, shared control instruments share functions with other instruments. Instruments that are controlled by computers are under the "computer function" category. Instruments that compute, relay, or convert information from data gathered from other instruments are under the "Programmable logic control" section.

For example, a discrete instrument for a certain process measures the flow through a pipe. The discrete instrument, a flow transmitter, transmits the flow to a shared display shared control instrument that indicates the flow to the operator. A computer function instrument would tell the valve to close or open depending on the flow. An instrument under the "Programmable logic control" category would control the valve in the field if it was pneumatically controlled, for instance. The instrument would gather information from discrete instruments measuring the position of the actuator on the valve, and would then adjust the valve accordingly.

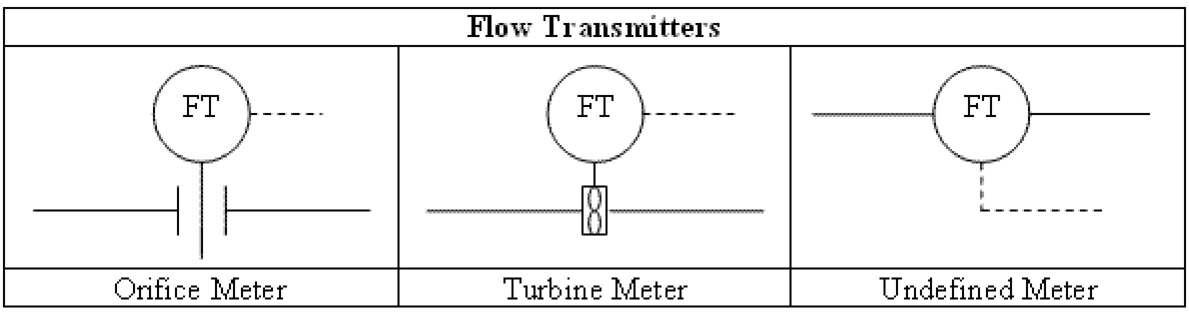
In the chart above, it is necessary to know where the instrument is located and its function in order to draw it correctly on a P&ID. A primary instrument is an instrument that functions by itself and doesn't depend on another instrument. A field mounted instrument is an instrument that is physically in the field, or the plant. Field mounted instruments are not accessible to an operator in a control room. An auxiliary instrument is an instrument that aids another primary or auxiliary instrument. Primary and auxiliary instruments are accessible to operators in a control room.

Transmitter Symbols

Transmitters play an important role in P&IDs by allowing the control objectives to be accomplished in a process. The following are commonly used symbols to represent transmitters.

Below are three examples of flow transmitters. The first is using an orifice meter, the second is using a turbine meter, and the third is using an undefined type of meter.

Table 6: Transmitter Symbols

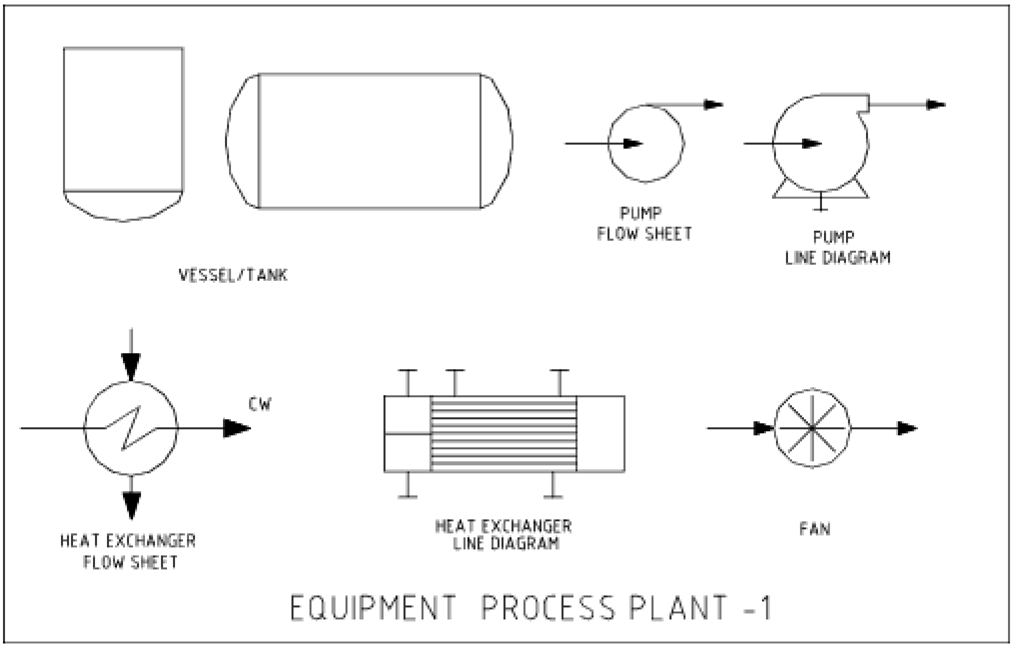


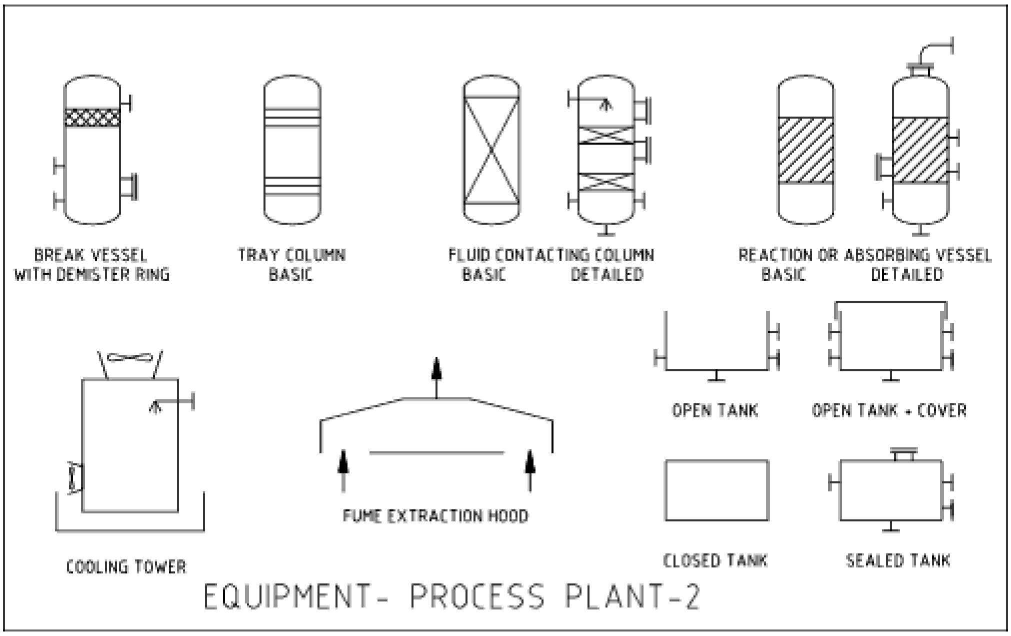
The location of the transmitter depends on the application. The level transmitter in a storage tank is a good example. For instance, if a company is interested in when a tank is full, it would be important for the level transmitter to be placed at the top of the tank rather than the middle. If the transmitter was misplaced in the middle because a P&ID was misinterpreted then the tank would not be properly filled. If it is necessary for the transmitter to be in a specific location, then it will be clearly labeled.

Miscellaneous Symbols

The following symbols are used to represent other miscellaneous pieces of process and piping equipment.

Table 7: Process Equipment





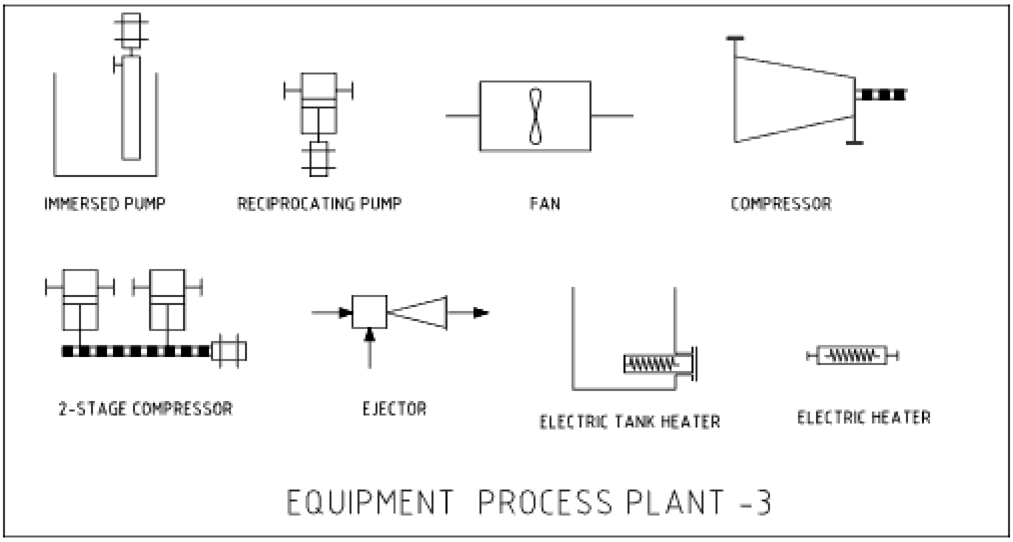


Table 8: Line Fittings

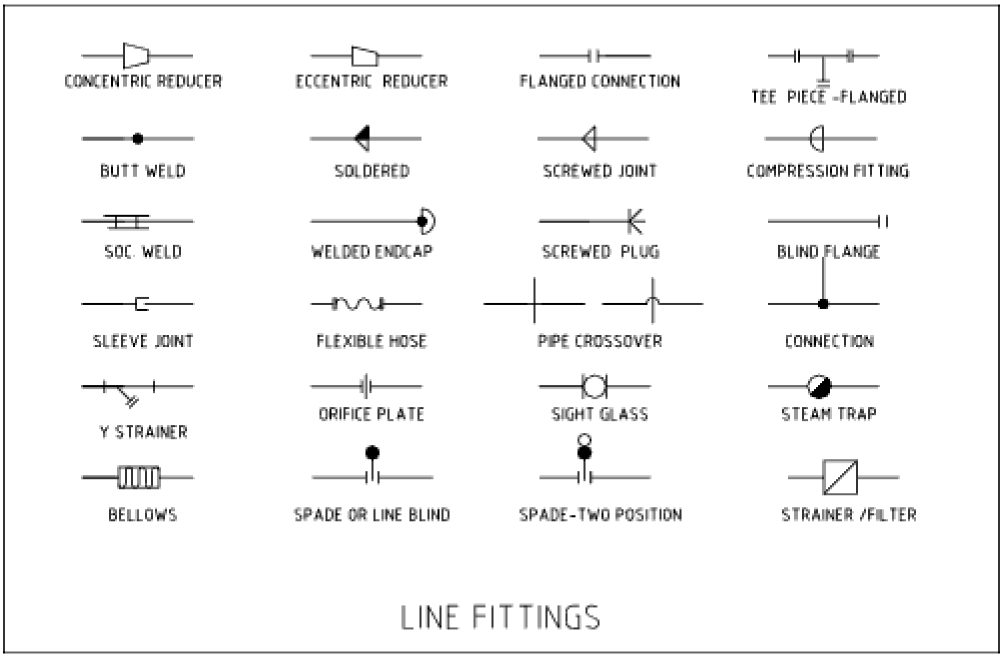
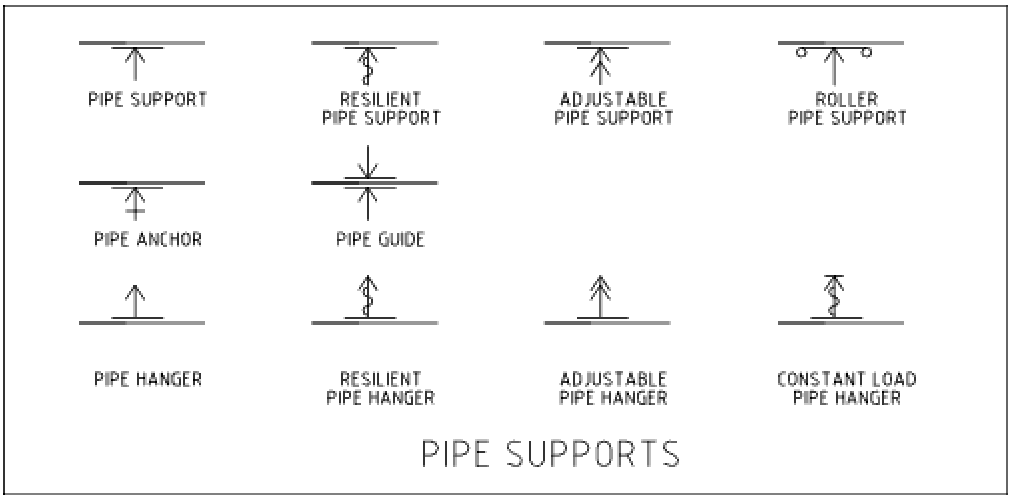


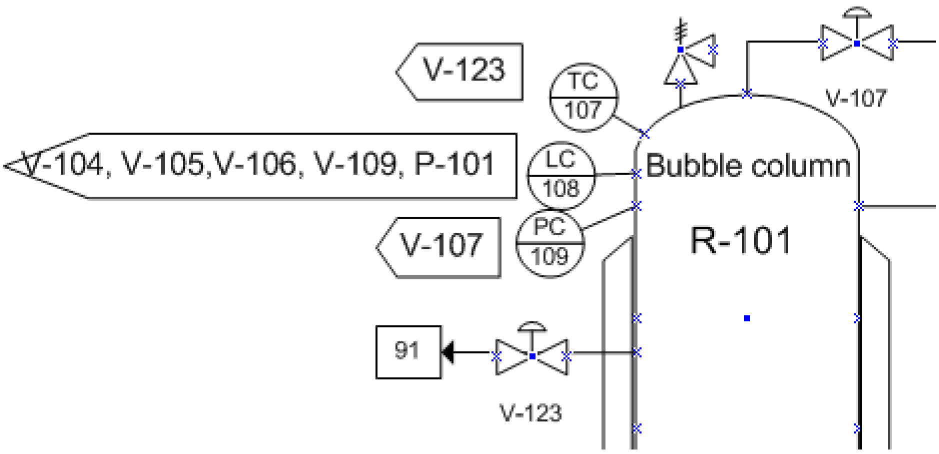
Table 9: Pipe Supports



Crafting a P&ID

In order to greatly simplify P&ID diagrams for the purposes of this class, a standard convention must be employed. This convention simplifies the many control devices that need to be used. For the sake of brevity, sensors, transmitters, indicators, and controllers will all be labeled on a P&ID as a controller. The type of controller specified (i.e. temperature or level) will depend on the variable one wished to control and not on the action needed to control it.

For instance, consider if one must control the temperature of fluid leaving a heat exchanger by changing the flow rate of cooling water. The actual variable to be controlled in this case is temperature, and the action taken to control this variable is changing a flow rate. In this case, a temperature controller will be represented schematically on the P&ID, not a flow controller. Adding this temperature controller to the P&ID also assumes that there is a temperature sensor, transmitter, and indicator also included in the process.



As you can see on the P&ID above, these controllers are represented as circles. Furthermore, each controller is defined by what it controls, which is listed within arrow boxes next to each controller. This simplifies the P&ID by allowing everyone the ability to interpret what each controller affects. Such P&IDs can be constructed in Microsoft Office Visio.

Sample Diagram

Below is a sample P&ID Diagram that is actually used in an industrial application. It is clearly more complicated than what has been detailed above, however, the symbols used throughout remain the same.

Table 10: Sample P&ID Diagram

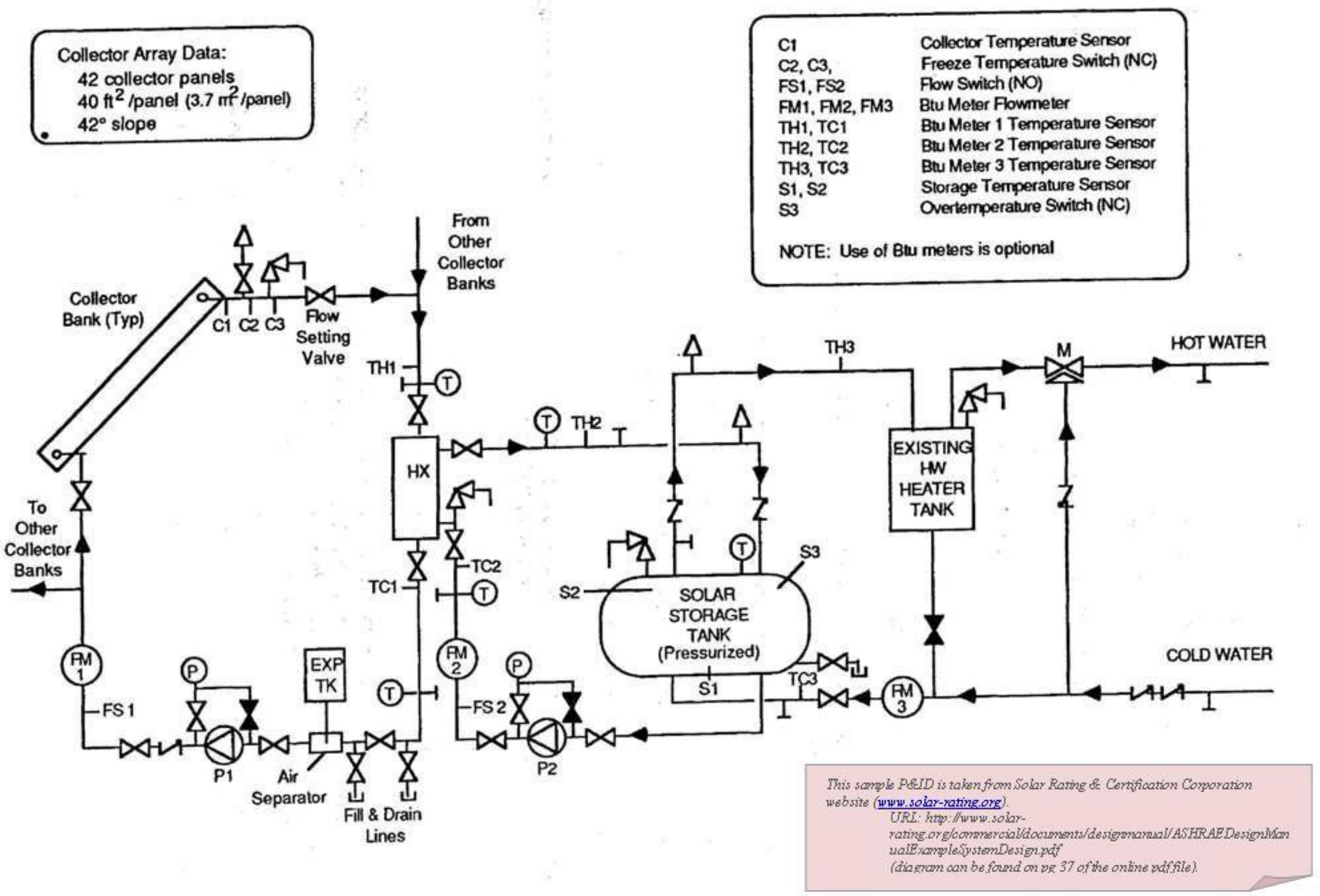
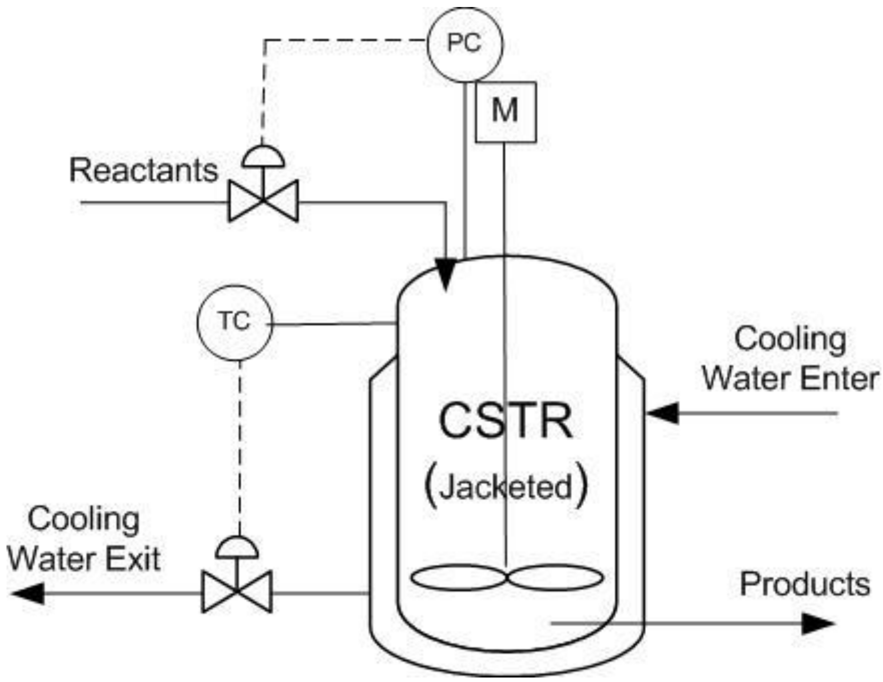


Figure B. Example System P&ID.

Example 1

Describe the following controlled process in words:

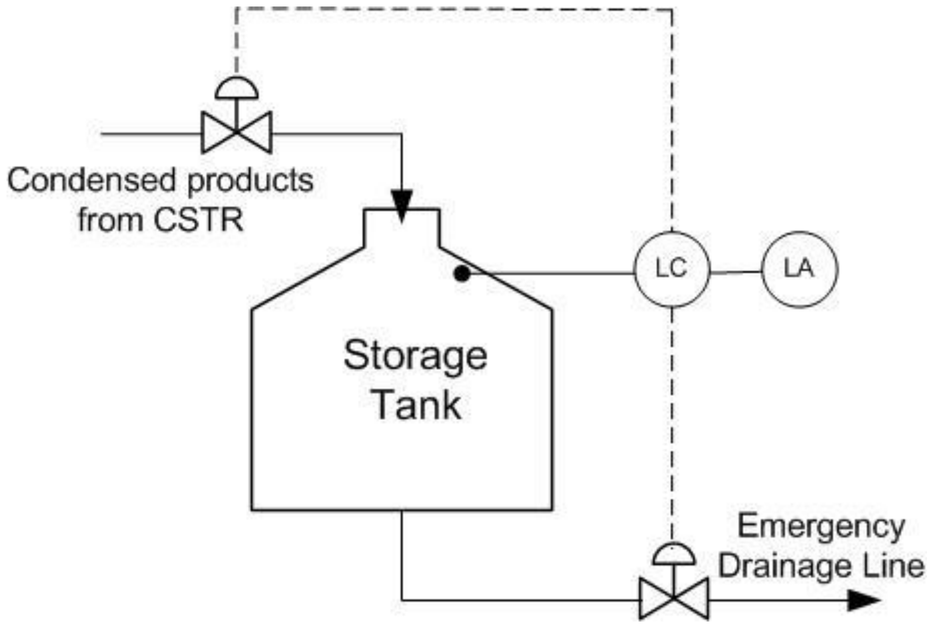


Answer: Reactants enter a jacketed CSTR where a reaction takes place and the products exit. The reactor is cooled via a coolant water stream. The temperature inside the reactor vessel is monitored with a temperature controller (also contained in the controller is a sensor, indicator, and transmitter), which electrically controls a valve. The valve can alter the flowrate of the coolant water stream, thereby controlling the temperature inside the reactor. A pressure controller is also present which feeds back to an inlet valve. Therefore, we can deduce that this reaction is most likely gas phase and if the CSTR becomes too full (high pressure) the inlet valve will close.

Example 2

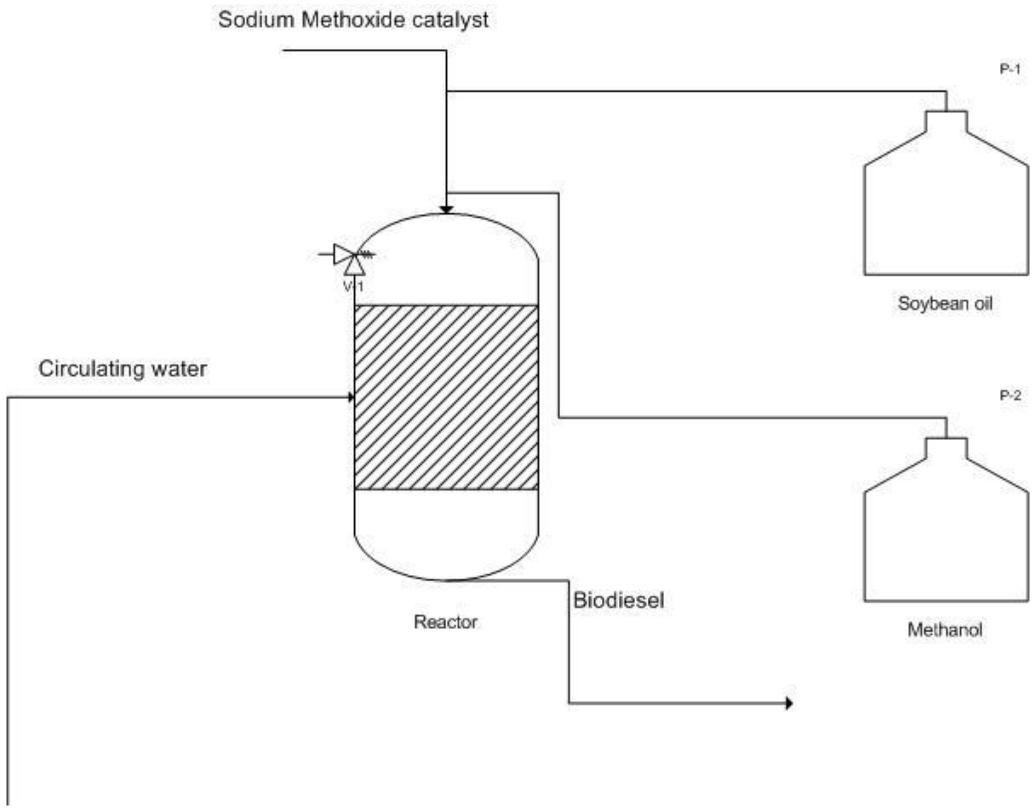
Draw a proper P&ID diagram of the following process:

A storage tank is filled with condensed products formed via the CSTR in Example 1. The tank contains a level controller at a set point on the top of the tank. If this tank were to fill, materials would get clogged up in the reactor. Therefore, if the tank reaches 90% of its total capacity, the level controller will send an electric signal, which opens an emergency drainage line located at the bottom of the tank. The level controller will also activate an alarm alerting plant engineers that there is a problem with the storage tank. Finally, the level controller will also close the inlet valve to the storage tank.

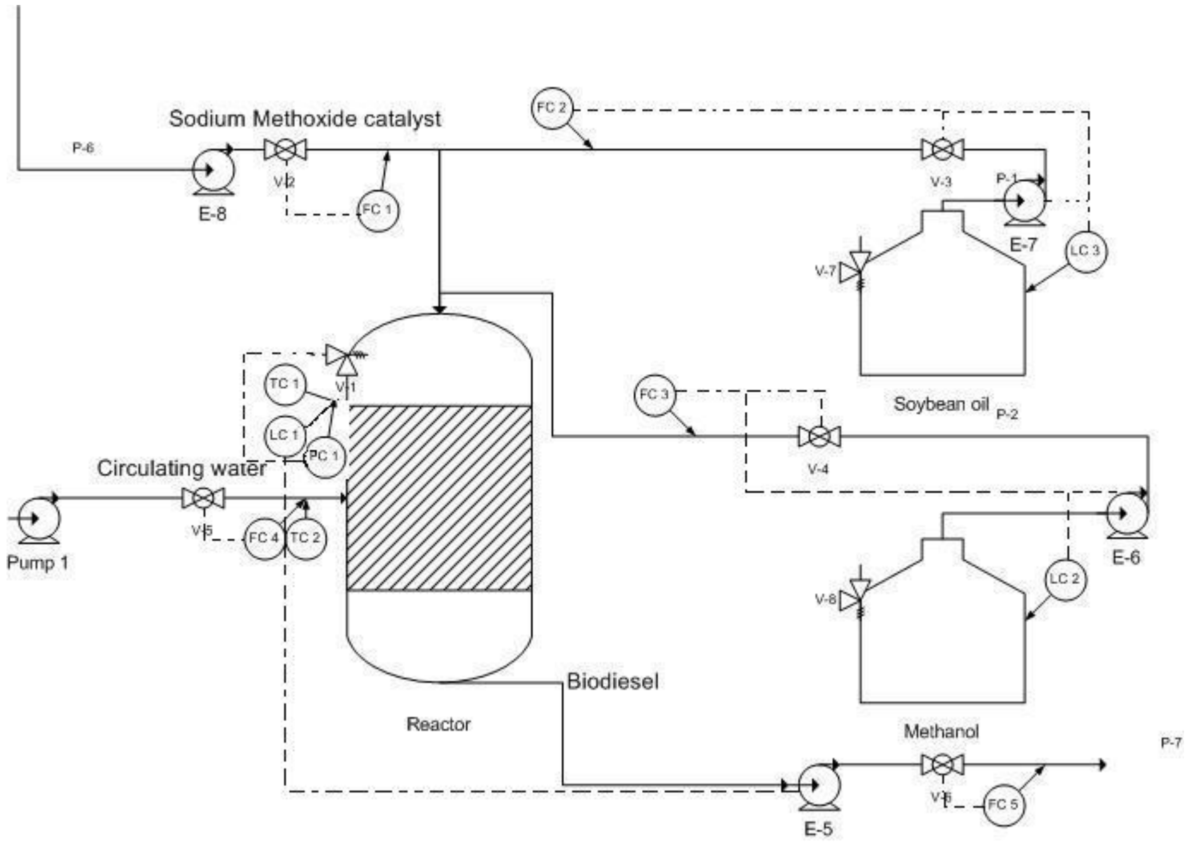


Example 3

Below is a P&ID diagram of the transesterification process to produce biodiesel. Soybean oil, methanol, and the sodium methoxide catalyst are pumped in to the reactor. The temperature of the reactor is regulated by the circulation water. The resulting biodiesel is then pumped out of the reactor and goes on to other processes so that it can be sold. Below is a P&ID of the process that is missing the valves, pumps, and sensors. Add the pumps, sensors, and valves that are needed to successfully control the process.



Solution:



Example 4

Below is a example problem of a typical P&ID problem. A is a liquid at Tamp but boils at Trx. B and P are high boiling point liquids and C is a solid. The reaction for the process is 2A+B+C-->P at Trx. Ais fed in excess.

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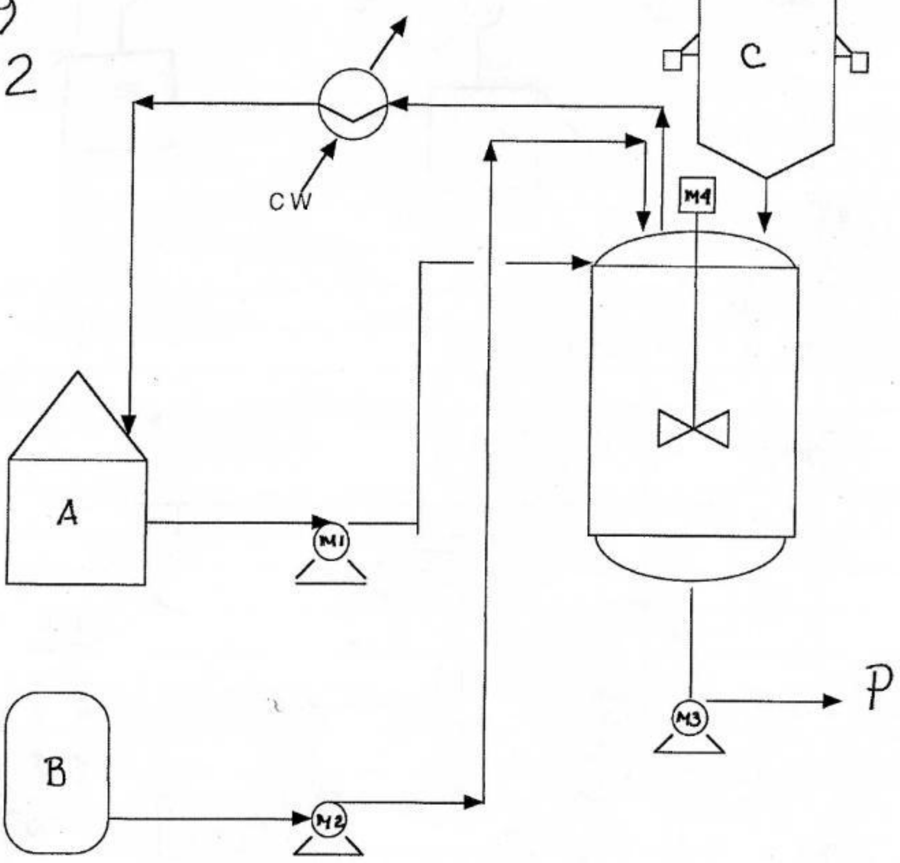


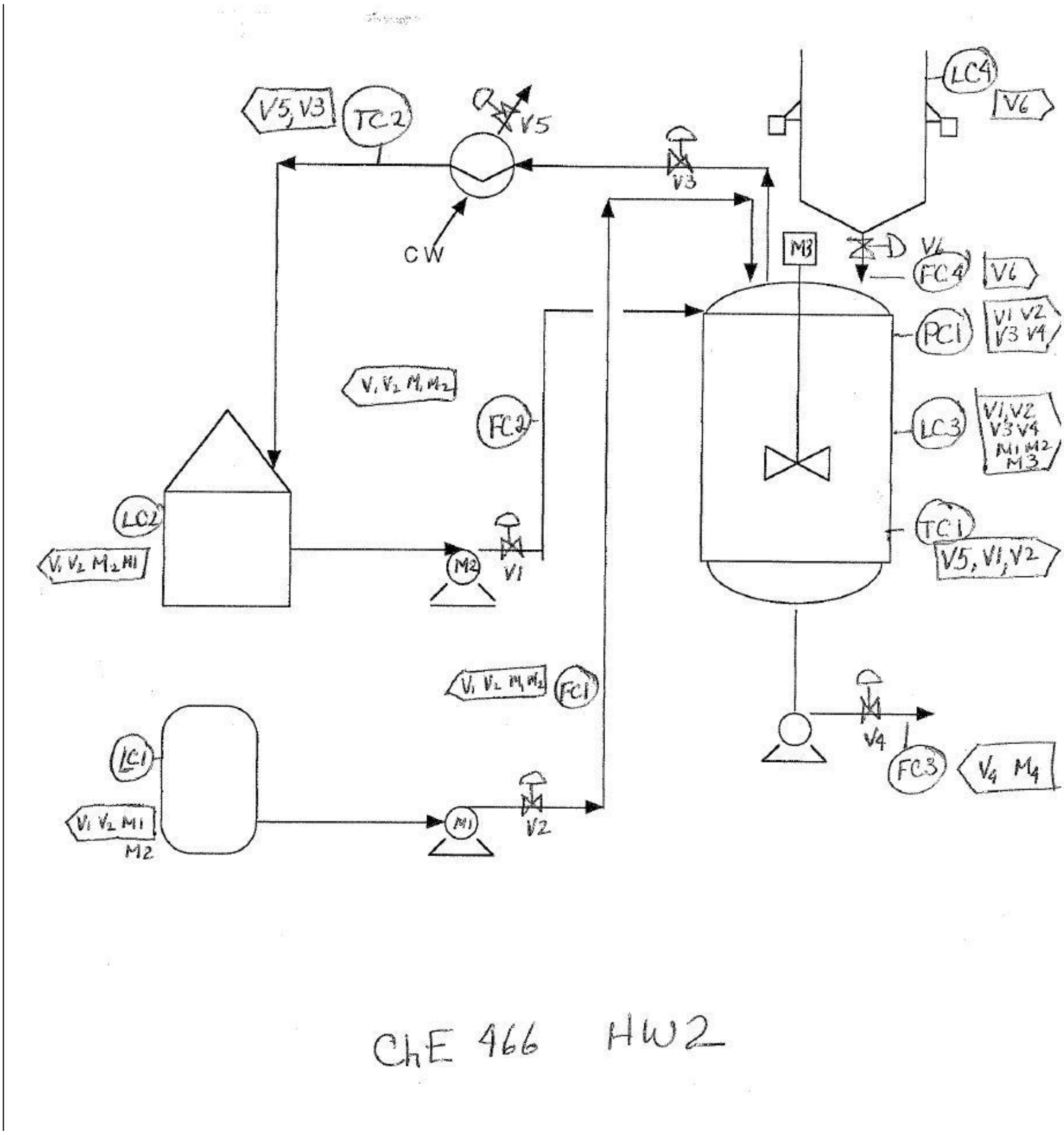
A is a liquid at T_{amp} but boils at T_{Rx}
B and P are high boiling liquids
C is a solid

$$2A + B + C \rightarrow P \;@\; T_{Rx}$$

A is fed in excess

1) Instrument the process ie add as needed
2) Show control devices ie or $\boxed{M}$
3) Show "what" controlers control ie M_1, M_2

Below is the solution to the problem above.



PIDStandardStructure

Title: Piping and Instrumentation Diagrams: Location of Controls and Standard Control Structures

Note: Video lecture available for this section!

Authors: Christopher Bauman, Jennifer DeHeck, Merrick Miranda, Evan Leonard

Stewards: Robert Chockley, Chloe Sweet, Irene Brockman, Scott Dombrowski

Anchors: Megan Boekeloo, Stacy Young

Date Presented: September 21, 2006 / **Date Revised:** September 7, 2007

- First round reviews for this page
- Rebuttal for this page

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Introduction

A Piping & Instrumentation Diagram (P&ID) is a schematic layout of a plant that displays the units to be used, the pipes connecting these units, and the sensors and control valves. Standard structures located on a P&ID include storage tanks, surge tanks, pumps, heat exchangers, reactors, and distillation columns. The latter three comprise most of the chemical process industry single-unit control problems.

P&IDs have a number of important uses in the design and successful operation of chemical process plants. Once a process flow diagram is made, P&IDs help engineers develop control strategies that ensure production targets are met while meeting all safety and environmental standards. These diagrams are also very important for locating valves and process components during maintenance and troubleshooting.

Each page of a P&ID should be easy to read and correspond to a specific action of the plant. Also, symbols used in the P&ID are uniform throughout. Diagrams for heat exchangers, continuously stirred tank reactors (CSTRs), and distillation columns shown in subsequent sections of this wiki represent a small sample of standard symbols for these structures. For a list and explanation of standard controls-related symbols, please refer to PID Standard Notation.

General Strategies for Implementation of Good Control Systems

There are several guidelines to follow when designing P&IDs for a plant. The first consideration is stable operation of the plant so that all safety and environmental standards are met. Maintaining product quality should also be a primary design objective. Additionally, systems should be designed to respond quickly to rapid changes in rate and product quality. Usually control systems should be run in an automatic mode so the system will correct itself, as opposed to a manual mode, which requires operator supervision.

When setting up a control system, it is useful to first focus on mass balance control, which can be monitored by level and pressure control loops that use gas or liquid flow rates. Additionally, a product control structure should be set up to ensure efficient process operation. To prevent controls from conflicting with one another, a final examination of the controlled streams should be performed. Further instrumentation can then be installed to prevent conflicts between controllers. Finally, appropriate tolerances should be established for controllers that directly affect the action of other controllers. Determination of the optimal control placement is essential for successful plant operation.

Standard Structures and Location of Control Features

As mentioned above, control of heat exchangers, reactors, and distillation columns represent the majority of single-unit control problems. The table below summarizes the control schemes derived in the following examples. The most common control schemes are listed in the table below.

Table 1: Summary of common P&ID structures

Structure	Control Variable	Manipulated Variable	Location of Control Valve	Comments	Typical Disturbances
Steam-heated Heat Exchanger	Temperature of outgoing stream	Steam pressure	Steam line or condensate stream	Placement of a control valve on the steam line gives tighter control, but this type of valve is bigger and more expensive than one placed on a condensate line	Fluctuation of the inlet flow rate
Liquid/Liquid Heat exchanger	Temperature of outgoing stream	Bypass flow rate	Bypass stream	This allows high coolant flow rates and decreases fouling. Changes in bypass stream's flow-rate result in temperature control loop disturbances that are easily regulated.	Fluctuations in inlet flow rates
Jacketed CSTR w/ Endothermic RXN	Product stream temperature	Steam flow rate or steam pressure in the jacket	Steam line	Changing steam flow rates allows direct measure of the amount of heat entering the system and therefore a measure of conversion. It also makes a reactor sensitive to heat load changes and enthalpy changes. Manipulating jacket pressure is better at absorbing reactor heat duty and steam enthalpy upsets but it does not directly measure heat load on the reactor.	Fluctuations in feed flow rate, temperature, and enthalpy of heating medium
Jacketed CSTR w/ Exothermic RXN	Inlet temperature of coolant	Coolant flow rate	Coolant feed stream	There is a slow response to fouling on heat transfer surfaces, but there is a fast response to changes in inlet coolant temperature and coolant supply pressure.	Changes in feed flow rate, temperature; Fluctuation of coolant enthalpy

For most equipment, a degrees of freedom analysis is first performed and then control scheme is designed based on the degrees of freedom.

Heat Exchanger

To monitor the performance of a heat exchanger, the product stream is important. Usually, the product stream must be within some temperature range before it continues to downstream process units. The outlet temperature of this stream can be used to calculate the heat transfer. The steam is controlled in order to obtain the desired product stream temperature. One way to influence the product temperature is by controlling the flow of the heated steam. This flow-based control may take some time to implement and therefore cause fluctuations in the process. Depending on the process, these fluctuations may or may not be acceptable. Instead of monitoring the flow-rate, the steam pressure may be monitored, achieving tighter control of temperature. A change in pressure is much easier to monitor and correlates directly with a change in temperature of the steam. This offers an effective way to control process temperatures. Pressure control also enables the physical condition of the piping to be monitored, since pressure changes occur as fouling progresses. The pressure, combined with the flow-rate and temperature of the heated steam, can be used to calculate this fouling occurring inside a piece of equipment. It is important to remember that the steam should always be controlled on the inlet side of the heat exchanger for better pressure control and safety reasons, as shown in figure 1 below.

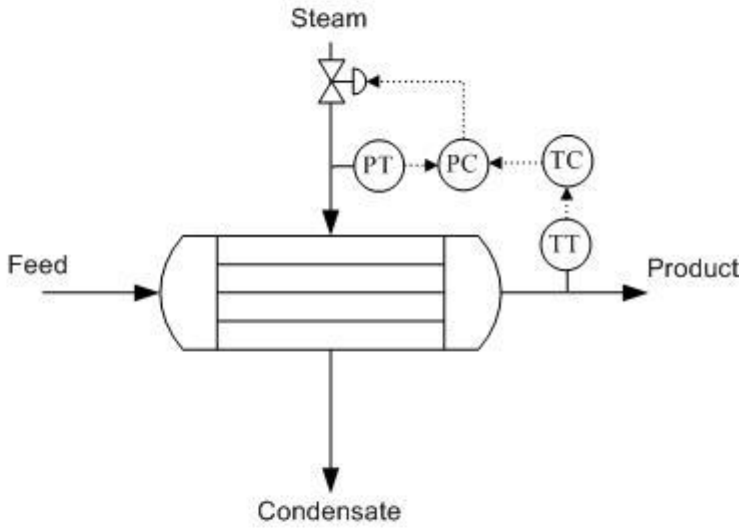


Figure 1. Heat Exchanger with pressure control on steam inlet and temperature control on the product stream.

Distillation Column

Because the economic viability of an overall process is based significantly on product purity, it is important that distillation columns maintain stable operation. Changes in composition and flow-rate of the feed stream are common disturbances in distillation column operation. Improper functioning of controllers can undermine the effectiveness of the product composition. A degrees of freedom analysis can help place sensors and actuators in appropriate places, while not including too many sensors and actuators, in order to obtain an efficient control system.

Degrees of Freedom Analysis

A simple degrees of freedom analysis can make design of a control scheme easier as well as improve the control scheme overall. A process control approach to degrees of freedom, adapted from ECOSSE Module 3.1, requires tabulation of streams and extra phases.

$DOF = Streams - ExtraPhases + 1$

A typical distillation column contains the following streams; feed (1), bottoms (2), distillate (3), reflux (4), product (5), vapor-liquid mix (6), cooling water (7), and steam (8). There are also three locations where there are two phases, vapor and liquid, present in equilibrium. These are denoted as one "extra phase", since temperature and pressure are not independent in a two-phase system.

$DOF = 8 - 3 + 1 = 6$

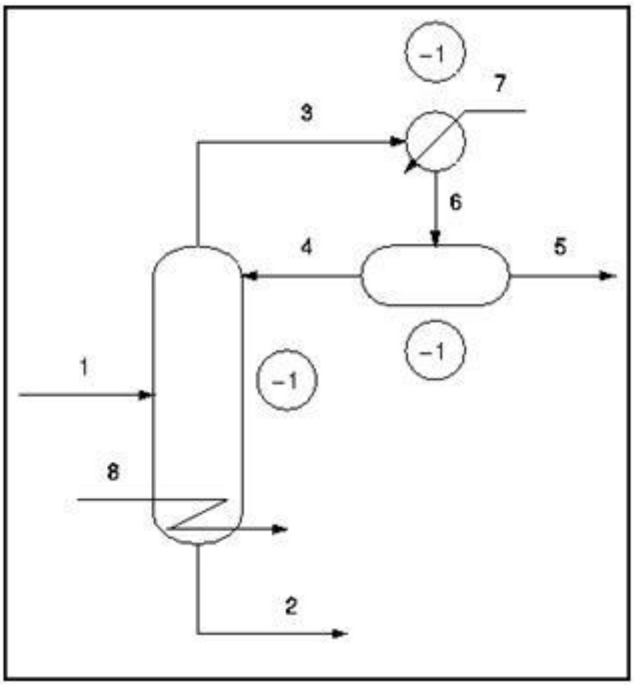


Figure 2. Degrees of freedom analysis on a typical distillation column (adapted from ECOSSE).

In typical practice, the condenser and reboiler pressure are specified as atmospheric pressure. Once the pressure is specified in the condenser and reboiler (both two-phase systems), the temperature is specified and therefore the degrees of freedom are reduced by two. The total degrees of freedom, in practice, is six.

$DOF = 8 - 5 + 1 = 4$

A simple degrees of freedom analysis, in any system, can help define the variables and reveal where the critical sensors for process control should be located. This type of analysis also minimizes the design of too many sensors, actuators, and valves, which may in theory control and regulate the same variable.

Sensor and Actuator Options

Product streams, reboiler steam, and the reflux stream should have flow sensor/transmitters so that each flow can be adjusted to meet the column's control objectives. A flow sensor should also be placed on the feed stream because disturbance to the feed flow rate are common. Each flow sensor should be connected to a flow valve to control the corresponding stream.

Differential pressure level sensor/transmitters should also be used for the accumulator (the vessel that collects condensed distillate) and bottom of the column because maintaining these levels is essential for reliable operation of the column. If flooding is an issue, a pressure differential across the column should indicate the onset of flooding. A pressure indicator should be installed at the top of the accumulator to monitor column pressure. Temperature in a distillation column is typically controlled by manipulating steam flow to the reboiler.

Control Schemes

There are a number of common control schemes for distillation columns. Optimally, a distillation column should be run with *dual composition* control because it saves energy. In dual compostion control, the temperature of both chemicals in a binary distillation is controlled. The system is more complex to setup and measurements required for control may be difficult. As a result, many distillation columns use single composition control instead. Common control configurations for distillation columns include reflux-boilup and distillate-boilup.

In the reflux-boilup configuration, the distillate composition and bottoms composition are the control variables. The reflux flow and the heat input control (vapor boilup) are the manipulated variables, which allow control of the liquid and vapor flow-rates in the column. With this control system, a quick response to changes in the feed composition is possible. In the distillate-boilup control configuration, the distillate flow and the vapor boilup are used to control composition. This configuration is a better choice for columns where the reflux ratio is high.

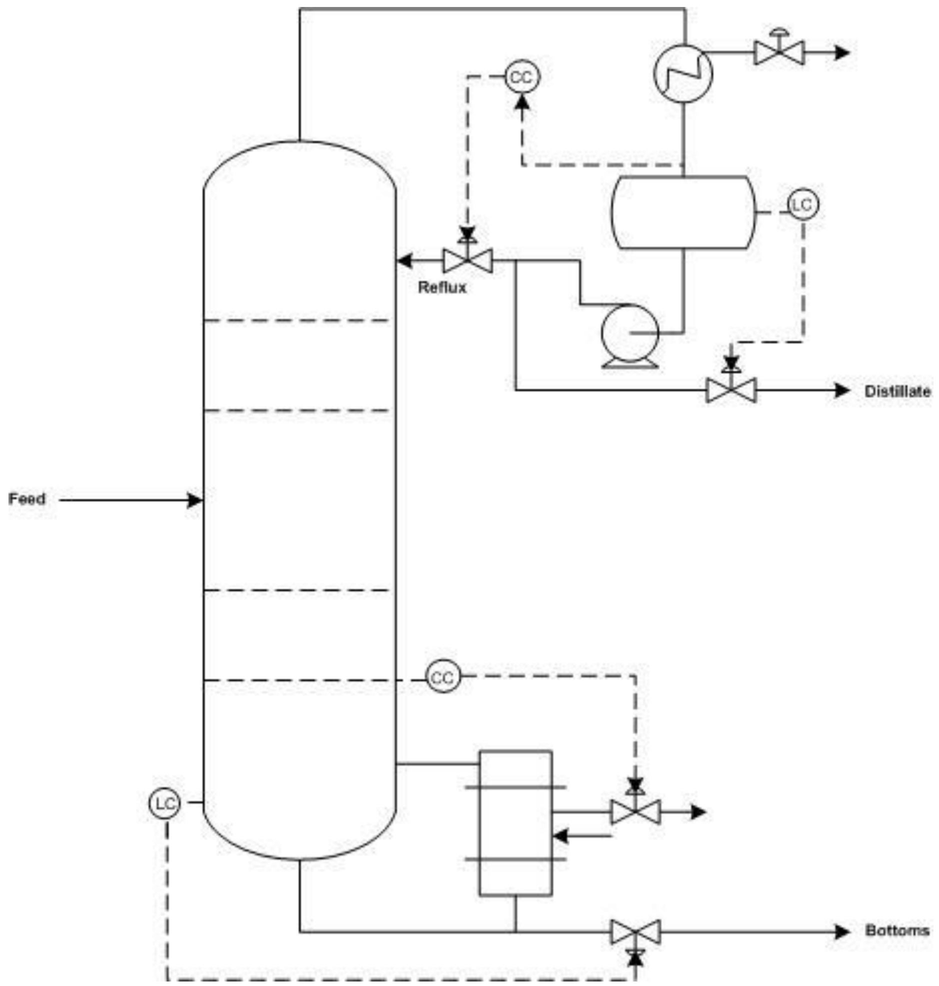


Figure 3. Distillation column with reflux-boilup control scheme (adapted from Luyben).

Reactor (Exothermic Reaction in CSTR)

In implementing controls in reactive systems, temperature is a good indicator for unit performance. Temperature is often related to reaction rate and varies with time in most reactors. However, a reactor exit stream with a constant temperature is often desired downstream in the process.

In an exothermic reaction in a CSTR, lower initial temperatures result in lower reaction rates and low heat generation. As the reaction progresses, heat generation increases rapidly due to higher reaction rates and high concentrations of reactants. As reactant concentration decreases, the heat generation once again becomes low. Exothermic reactor temperatures must be controlled to assure stable reactor operation.

To remove heat from an exothermic reaction, basic heat transfer principles are employed. A coolant is pumped through a shell outside of the reactor. Since the heat removal is linear, the temperature of the coolant should be controlled. In doing this, it is possible to increase the driving force for heat transfer to slow a reaction, or conversely, to allow the reaction to further progress by decreasing the driving force for heat transfer.

Degrees of Freedom Analysis

In order to determine where to place controls, sensors, and valves in a exothermic reaction operation, a degrees of freedom analysis, similar to that carried out in the distillation section 3.2, can be helpful.

A typical CSTR contains the following streams; reactant A (1), reactant B (2), product (3), and coolant to the jacket (4). There are no locations where there are two phases, vapor and liquid, present in equilibrium - assuming the reaction is liquid phase with no simultaneous evaporation or sublimation.

$DOF = Streams - ExtraPhases + 1$

$DOF = 4 - 0 + 1 = 5$

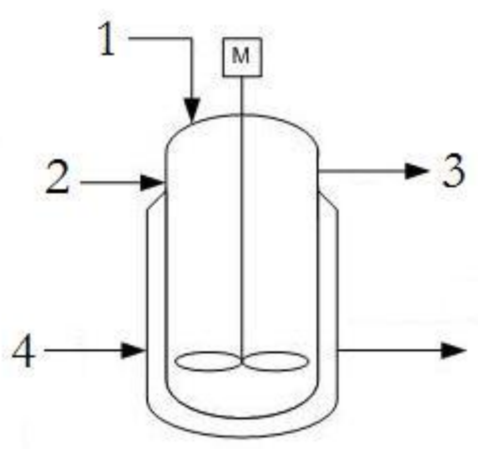


Figure 4. Degrees of freedom analysis on a typical exothermic reaction in a CSTR.

Since in most practical applications the reactants (1 and 2) as well as the product (3) flow rates are defined by demand for the product, there are only 2 degrees of freedom which allow for placement of a 2 valves or controllers on the coolant stream (4) as well as one of the reactant streams (1 or 2). Only one of the reactant streams need to be specified as the other can be determined by a ratio controller using stoichiometric coefficients.

Control Schemes

As discussed previously, a valve can be placed on the coolant inlet stream to ensure proper temperature control of the reactor. By controlling the coolant stream based on inlet conditions of the reactant streams, the control can respond quickly. But, if the coolant is controlled based on outlet conditions of the product stream, there is a lag in response, but it is easy to monitor heat transfer performance.

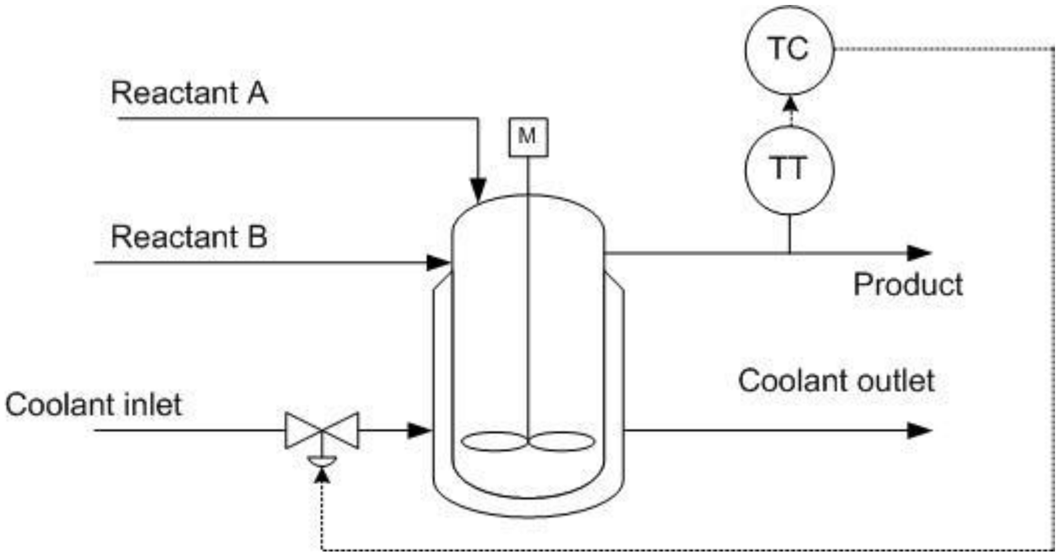


Figure 5. Jacketed CSTR with coolant control based on the outlet temperature conditions.

To translate this example to an endothermic reaction, the coolant flow would simply be translated to steam flow in order to provide heat to the reaction through the CSTR jacket.

Other Common Process Equipment

Beyond the heat exchangers, reactors, and distillation columns, many other pieces of process equipment, including furnaces, compressors, decanters, refrigerators, liquid-liquid extractors (LLEs), and evaporators, are subject to disturbances and require careful control.

Furnaces

For example, furnaces may be subject to frequent load changes as a process or customer requires more energy. To cope with these demands, the temperature of the outlet stream must be monitored and manipulated. Information from a temperature controller at the outlet stream can be used to effect changes in valves that control the flow-rate of air and fuel to the furnace. At this point, the best setup of the control system must be considered in light of factors such as safety, lag time for changes to occur, and equipment wear. For a furnace, the controls should generally be set up so that an excess of air is more likely than an excess of fuel. This reduces the risk of explosion due to the furnace filling with uncombusted fuel.

Liquid-Liquid Extractors

In liquid-liquid extractors, the interface level and pressure are the controlled variables. Disturbances in flow rate of the entering stream can affect interface level and prevent complete separation of the heavy and light components. From this, it is obvious that there should be valve controls on both exit streams. The best control scheme depends on the operation of the process. When the heavy phase is continuous (light phase flows upward through heavy phase), changes in interface level should be controlled by adjusting the flow-rate of the light product, while the pressure is controlled by adjusting the flow-rate of the heavy product out of the column. Figure A is a representation of what occurs in a single stage extractor. Generally, single stage extractors are using in chemical labs, where as multistage extractors are used in industry. A multistage extractor uses the immiscible liquid stream from the previous stage as the feed in the following stage. Figure 6 depicts the control scheme previously described.

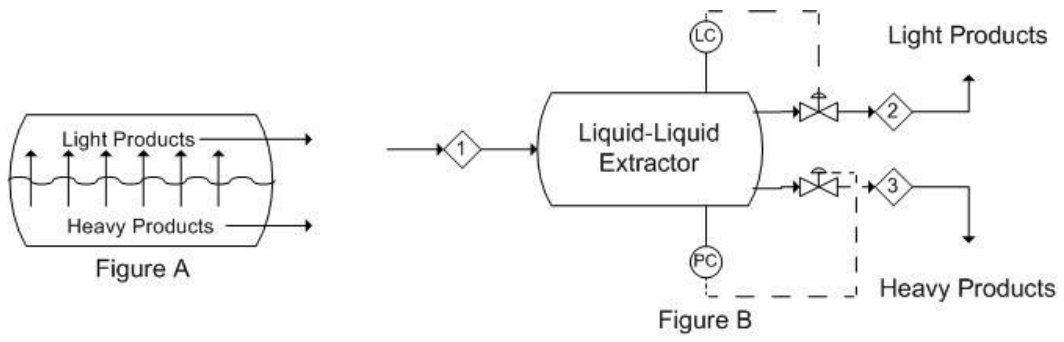


Figure 6. Liquid-liquid extractor control scheme with interface level and pressure control.

When the light phase is continuous, the control system must be set up in the opposite manner. Figure C is again a depiction of what occurs in a single stage extractor. Figure D is a representation of a control scheme that could be implemented. This is the reverse of the control scheme in Figure 6.

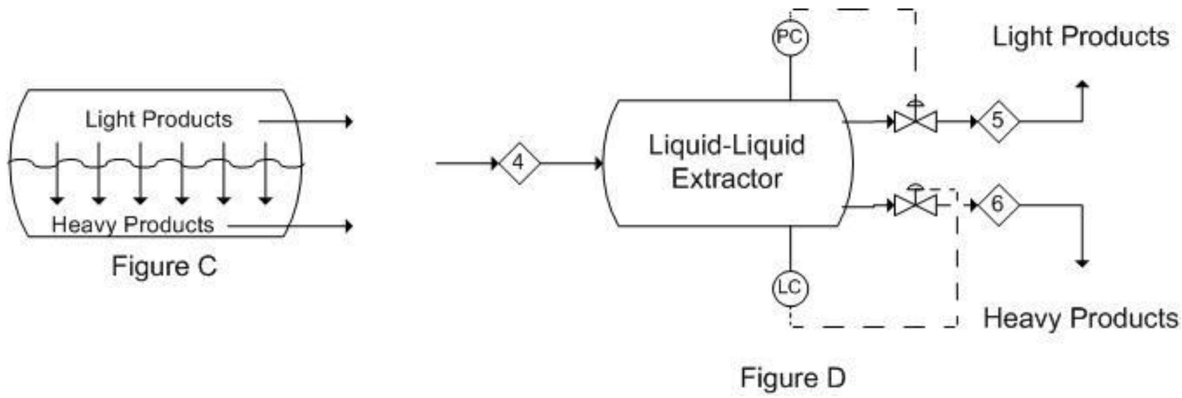


Figure 7. Liquid-liquid extractor control scheme with light-phase control.

These representations are only two possibilities for basic control schemes that can be implemented in a process as there are several controllers and aspects of the specific processes that may need to be controlled or monitored.

Compressors

Compressors are another valuable component in process design. They allow for the reduction of the volume of an incoming stream by increasing the pressure at which the stream is maintained. They can also be used to pump liquids through the process, as liquids are highly incompressible compressors cannot be used for volume reduction. For this, there must be a specific control system as to prevent adverse effects due to extremely high or low pressures. There are several types of compressors, including: dynamic, axial, and rotary [1] to name a few. Because the increase in pressure is governed by the ideal gas law, there is most often and increase in temperature as well. This can be left as is, or sent to a heat exchanger for temperature reduction. Heat exchangers were discussed above.

One such example is the use of a centrifugal compressor to reduce the volume of a fuel stream for storage. Using a compressor will allow for volume reduction as gasses are easily compressed, this is also economically friendly as it reduces the size of tank necessary to store the fuel stream. The tank should also be equipped with a pressure reducing valve, to bring the stream back to a desired pressure, depending on the process. A diagram of this scheme is as follows:

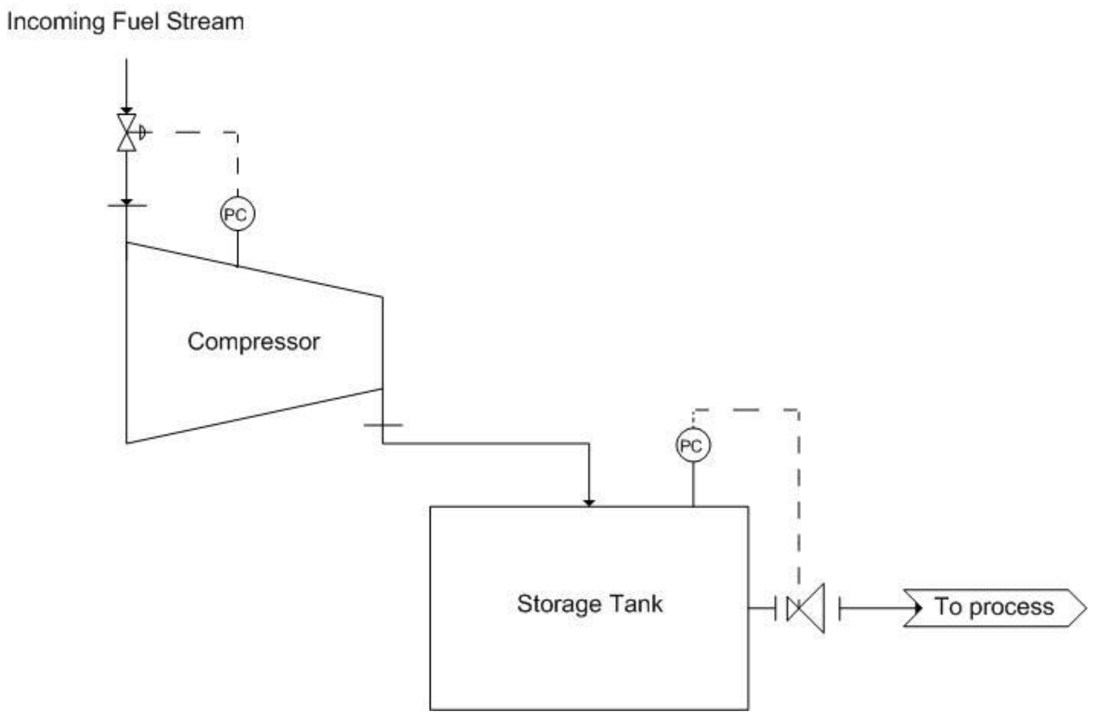


Figure 8. Compressor control scheme with pressure control.

The pressure controller on the compressor controls the valve on the incoming fuel stream. This ensures that if there is a build up in pressure, the flow into the system will be stopped in time. Also, a pressure controller should be placed on the storage tank. This is controlled by the pressure reducing valve mentioned earlier.

Decanters

Decanters, much like Liquid-liquid extractors, use solubility as their principle of separation. Unlike Liquid-liquid extraction, these require some time for the separation to occur. Generally, the separation is a liquid-liquid or liquid-solid separation. Decanters are widely used in the wine industry to separate out sediment from the wine. Utilizing the wine separation example, a possible control scheme is as follows. Here, there is only a level sensor on the decanter as it is a liquid-solid separation. Note that there is a release stream, used to remove collected sediment. An analytical or pH sensor could also be used here to maintain the correct chemistry or pH, as wine is the final product. It is also important to note that the exact placement of the level sensor would vary depending on the desired level in the tank. Here, the level sensor is not shown in its exact placement.

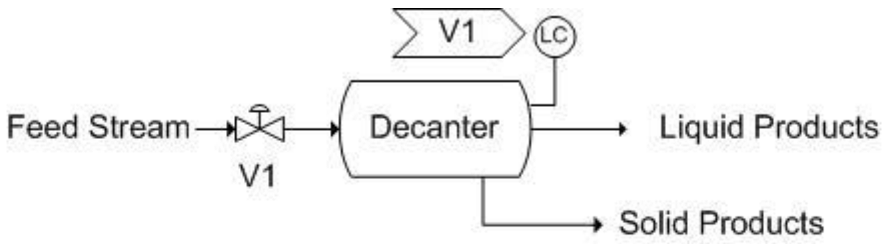


Figure 9. Decanter control scheme with level control and pH control.

Another vessel that is very similar to decanters are knockout drums. These vessels are generally located after heat exchangers or other pieces of equipment that result in a multiphase system. These vessels are used to separate the two phases, generally gas-and liquid separation. A possible control scheme is depicted below. The incoming stream is a liquid-gas mixture coming from a heat exchanger. Thus, there is a pressure sensor on the knockout drum. A level controller could be used, but this is effectively measuring the same thing, so it has been omitted. Also, because it is coming from a heat exchanger, a temperature controller has been included to control the amount of cooling taking place in the heat exchanger.

The pressure controller (PC) controls V2, the stream into the tank. The temperature controller (TC) controls V1, the valve on the coolant stream to the heat exchanger.

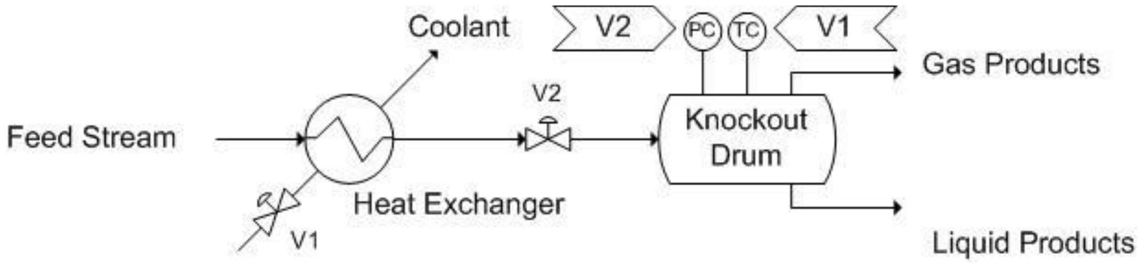


Figure 10. Knock-out drum control scheme with pressure and temperature control on different streams.

These examples illustrate the typical method of locating control systems for process equipment. After choosing the location for valves based on process constraints, there still remain a number of possibilities for the actual manner in which to control the valves. The lag-time for changes may be longer for certain configurations, while the danger of damaging equipment is smaller. The controls configuration will depend strongly on the safety concerns associated with a specific process.

Selecting Controls and their Locations for a Multi-Unit Process

The following steps should be followed when setting up controls for multi-unit processes. For more in-depth descriptions, refer to the Step-by-Step Method For Describing Controls and Their Purpose.

1. Determine process objectives, taking into consideration product specifications, economic constraints, environmental and safety regulations, etc.

2. Identify boundaries for normal operation. These can be based on equipment limitations, safety concerns, environmental regulations, and the economic objectives of the processes.

3. Identify units and streams in the process that are susceptible to significant disturbances. These disturbances commonly occur in feed streams, product streams, and reactor vessels, but can be present anywhere that temperature or pressure or other variables are changing.

4. Select the types and locations of sensors in order to properly measure and monitor critical process variables.

5. Determine the appropriate types and locations for control valves in order to appropriately adjust process variables so that they remain within the normal operating boundaries. Controls should be set up to minimize response time between sensing a change and taking corrective actions. The ideal location for any given control depends on the process unit or units that it affects.

6. Perform a degree of freedom analysis.

7. Energy Considerations. An energy balance should be performed for the process. This step involves transporting energy to and from process units. This may include removing heat generated by a reactor and using it elsewhere in the process. Control valves will help regulate the flow of such streams.

8. Control Process Production Rate and Other Operating Parameters. Adjusting process inputs, such as reactant feed rates, can alter other variables in the process. Process controls must be able to respond to these adjustments to keep the system within operating boundaries.

9. Set up control system to handle disturbances and minimize their effects. (See Chemical Process Controls > PID Control)

10. Monitor Component Balances. Accumulation of materials within a system is not desirable and can lead to inefficiency in the process or catastrophic failure.

11. Control individual unit operations. Each unit of a multi-unit process needs to be individually controlled in order for control of the entire system to be possible.

12. Optimize the process. If the system has degrees of freedom, process variables can be manipulated in order to more efficiently or economically create product.

Example 1: Completing a Control System

Consider the following reaction in a semi-batch reactor: $(CH_3CO)_2O + H_2O \rightarrow 2CH_3COOH$

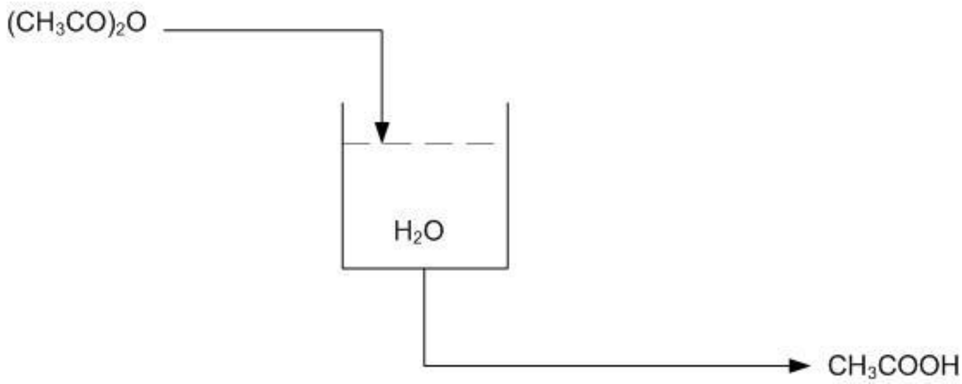


Figure E-1. Basic Flow Diagram of reaction of Acetic Anhydride with water to produce Acetic Acid.

The objective is to produce CH3COOH at a given concentration, i.e. 15-20%. The tank should also be drained once the given concentration is achieved and the reaction has gone to completion. The tank can only hold a specified volume and 80% is filled with water. The reaction should be run in excess water. Once a given amount of acetic anhydride has entered the reactor, it should be left in the tank until the reaction has gone to completion. The completion of the reaction can be determined by temperature, as the reaction is exothermic.

Using the multi-unit controller design from above, determine the location of needed sensors and valves on the flow diagram.

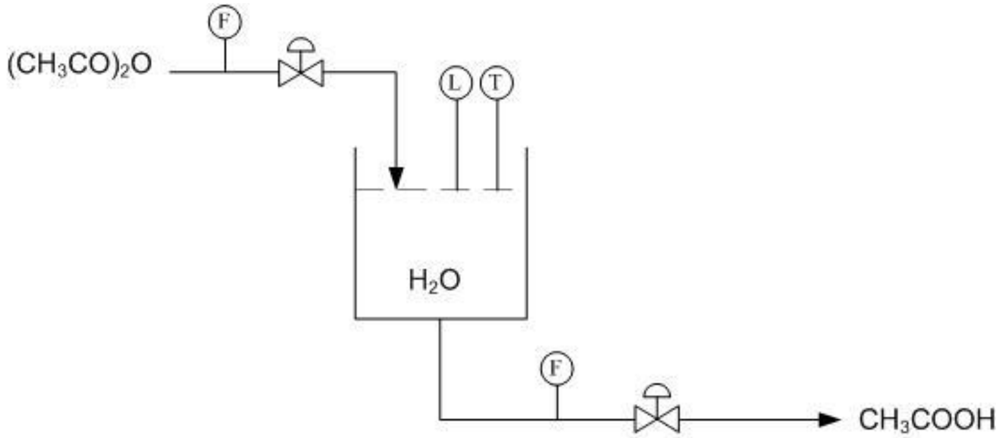


Figure E-2. Location of sensors and valves.

Using Figure E-2 determine how to use the level and temperature sensors to limit the amount of acetic anhydride entering the reactor and also how to drain the system once the reaction has gone to completion.

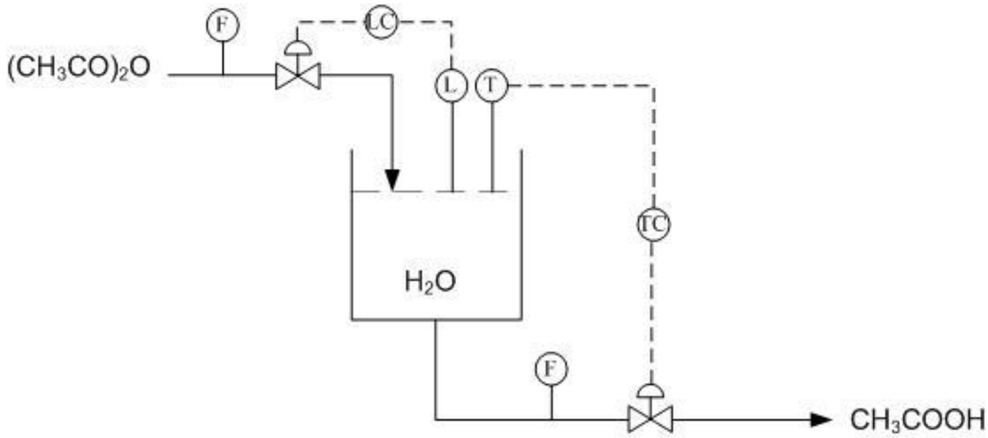


Figure E-3. Control system for production of acetic acid.

Using the above system, when the right amount of acetic anhydride has entered the reactor the level sensor will stop the flow into the reactor. Once the reaction has gone to completion and the temperature sensor has been triggered, due to a peak in temperatures from the exothermic reaction, the tank will drain.

Example 2: Control of a Heat Exchanger

Examine the figure below, which shows two different control schemes for a heat exchanger. The outlet stream temperature must remain constant. How is this goal accomplished using the control systems shown below? Which stream is the manipulated stream? What are the advantages and disadvantages of each setup?

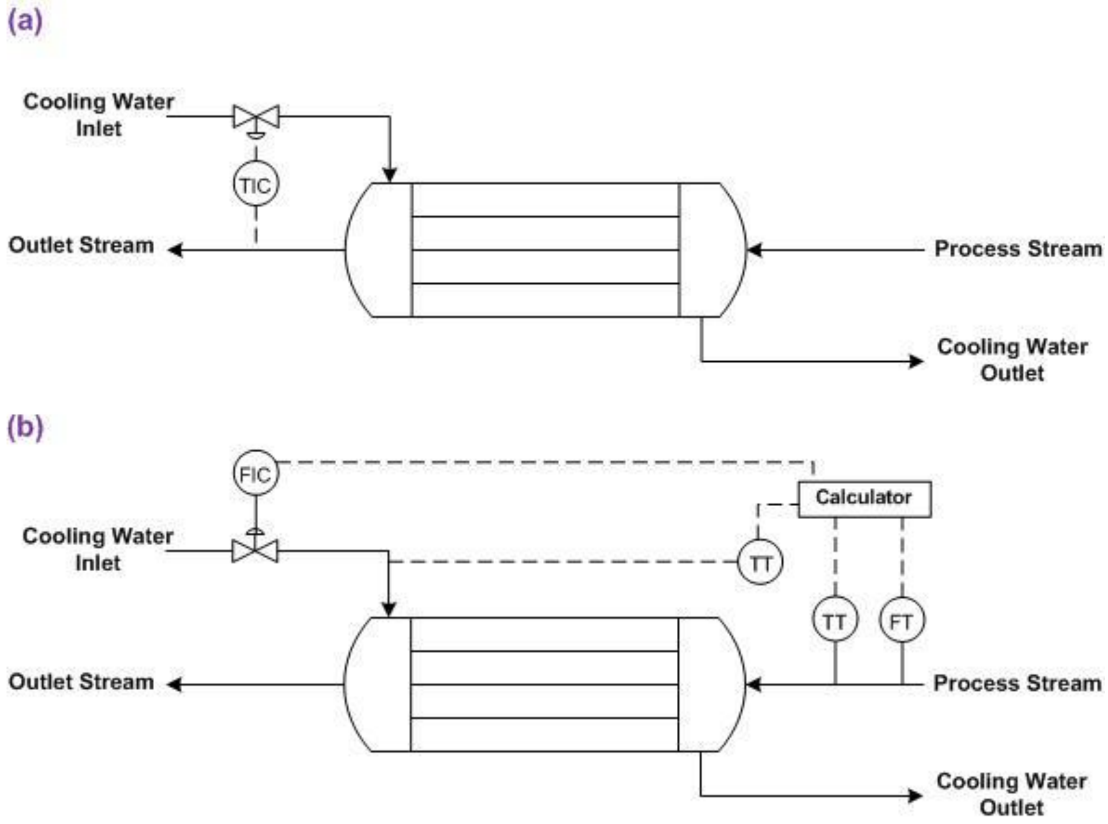


Figure E-4. Two different control strategies for a heat exchanger (adapted from Turton)

In both setups, the cooling water flow is manipulated to control the temperature of the outlet stream. The control strategy for each heat exchanger is different.

Figure E-4 (a) shows a heat exchanger where the flow of cooling water is adjusted based on the temperature of the outlet stream. The advantage to this setup is that it is quite straightforward and the cause of the change in the temperature of the outlet stream does not have to be identified. The disadvantage is that a temperature change has already occurred before action is taken, and if the lag-time for the additional cooling water to cool the process stream is long, the outlet stream temperature may be inconsistent.

Figure E-4 (b) shows a heat exchanger were the flow of cooling water is adjusted based on the flow-rate and temperature of the inlet stream and the temperature of the coolant stream. Based on a set of calculations using the heat transfer coefficient of the heat exchanger, the outlet stream temperature can be predicted. The advantage of this system is that action is taken before changes occur in the outlet stream temperature. The disadvantage is that if the actual heat exchange capacity differs from the expected value due to fouling or mechanical problems, the outlet temperature will not be correct.

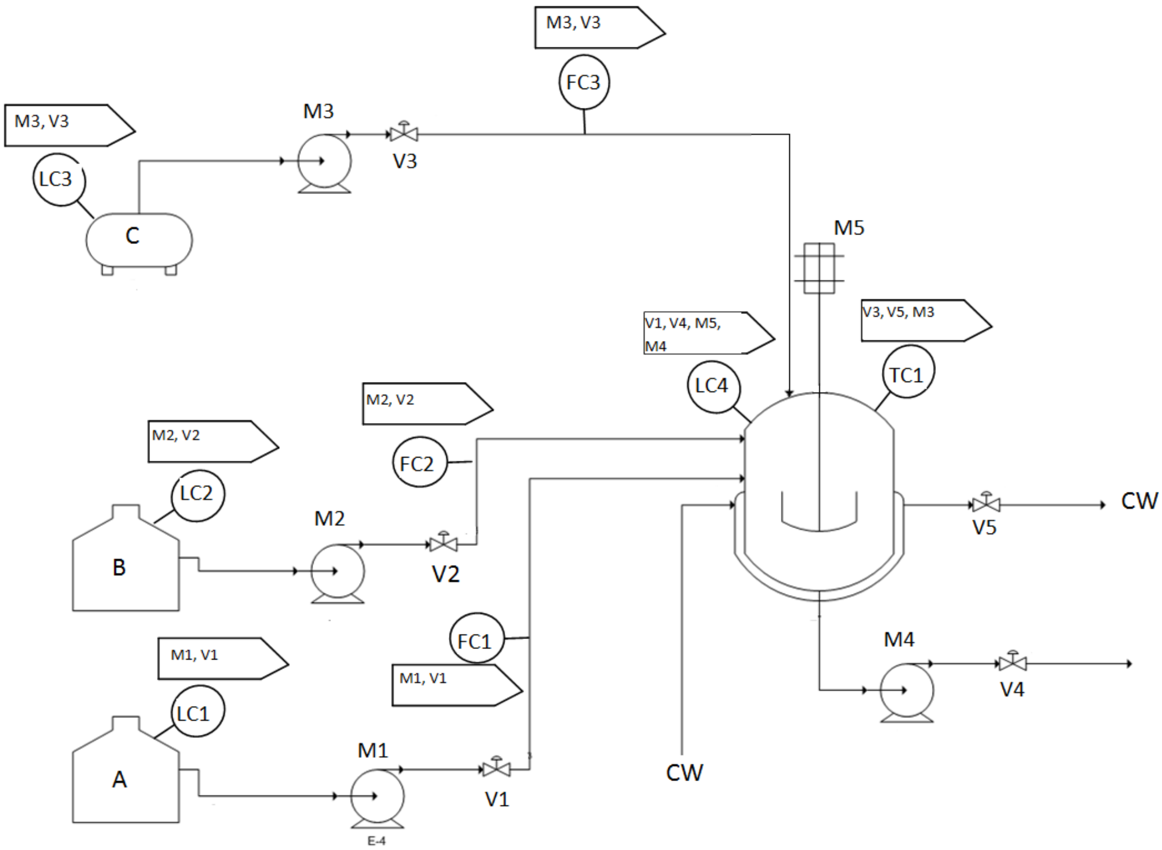
Example 3: An alternate notation, with explanation of controller location and function

This presents an alternate notation of controls. The arrows above the controller indicate what parts of the process that the controller controls.

The Problem

A + B − > AB

1. B is added slowly into a full charge of A
2. The temperature is maintained at TR (RXN temperature)
3. Upon completion of reaction, AB is cooled to TP
4. C is added to AB to make final product, which leaves the tank



The Solution

- **LC1**: This controls M1 and V1.

  This controller is used to keep the level of A in the A tank above a preset minimum.

- **LC2**: This controls M2 and V2.

  This controller is used to keep the level of B in the B tank above a preset minimum.

- **LC3**: This controls M3 and V3.

  This controller is used to keep the level of C in the C tank above a preset minimum.

- **FC1**: This also controls M1 and V1.

  This controller totalizes the amount of A that has flowed out of the A tank, and turns off flow when the flow amount has reached a preset value.

- **FC2**: This also controls M2 and V2.

  This controller totalizes the amount of B that has flowed out of the B tank, and turns off flow when the flow amount has reached a preset value.

- **FC3**: This also controls M3 and V3.

  This controller totalizes the amount of C that has flowed out of the C tank, and turns off flow when the flow amount has reached a preset value.

- **LC4**: This controls V1, V4, M4, and M5.

  V1 - If level of liquid in the reaction tank goes over a preset maximum, V1 is closed.
  V4 - If level of liquid in the reaction tank goes under a preset minimum, V4 is closed.

  M4 - If level of liquid in the reaction tank goes under a preset minimum, M4 is turned off.

  M5 - If level level of liquid in the reaction tank goes under the level of the agitator, M5 is turned off.

- **TC1**: This controls V3, V5, and M3.

  M3, V3 - If the temperature in the reaction tank is at TP, M3 is turned on and V3 is opened.
  V5 - To maintain the temperature in the reaction tank at TP, V5 is opened and closed accordingly.

PIDStandardPitfalls

Title: P&ID Standard Pitfalls

Note: Video lecture available for this section!

Authors: Angela Yeung, Curt Longcore, Ben Van Kuiken, Jeffrey Carey

Date Presented: 5 October 2006 /Date Revised: 19 October 2006

Stewards: Eric Black, Megan Boekeloo, Daniel Carter, Stacy Young

Date Updated: 16 September 2007

- First round reviews for this page
- Rebuttal for this page

Contents

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Introduction

Piping and Instrumentation Diagrams (P&ID) are standardized in many ways, and there are some fundamental safety features that are absolute requirements for all P&IDs.

Unfortunately, many people forget these features in their designs unintentionally. Lacking these safety features could lead to serious engineering problems. It is important to eliminate these pitfalls when designing a P&ID.

In the following sections, different pitfalls of P&IDs will be discussed. The equipment design section will concentrate on how equipment might be accidentally neglected and misplaced while designing a safe and functional process. The design process section will describe how a lack of consideration of process conditions would lead to serious design pitfalls. Overspecification and underspecification of equipment and design will be discussed. There are also some miscellanous pitfalls associated with interpretation and inherent problems of the P&ID itself. Examples are introduced at the end to illustrate common errors and pitfalls of P&IDs.

P&ID Production Pitfalls

Most of the common pitfalls of P&ID production result from the engineer forgetting that the design on paper represents a real physical process and that there are practical and physical limitations that need to be considered in the process design. The following are some errors to be particularly careful of when drawing up a P&ID:

Equipment Design

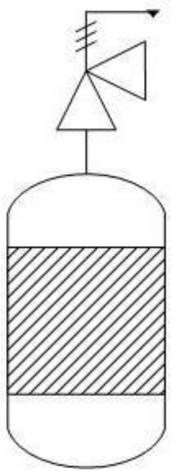
Safety valves

Safety valves are part of the essential valves system for P&IDs. Together with isolation valves, they are an absolute requirement for instrument design. Safety valves are required to install for all gas, steam, air and liquid tanks regardless of the tank's function for pressure relief purposes. Engineers should be aware of their system's set pressure, relief pressure, percent overpressure, maximum allowable pressures etc. when selecting a safety valve. The US law requires all tanks of pressure greater than 3 psig to have safety valves installed.

Different pressure tanks require different safety valves to best fit their safety design. Therefore, engineers must be very careful in selecting the right safety valves for their systems. However, when constructing a P&ID, engineers sometimes forget adding safety valves to their design, and this could cause serious problems. For example, if one forgets to add a pressure relief valve or safety valve on a reaction tank of gas and liquid, the extra pressure accumulating would exceed the preset pressure limits for safety design. This could lead to a serious explosion!

A pressure relief valve is symbolized by two triangles orientated at 90 degrees to each other, as shown below.

Pressure Relief Valve

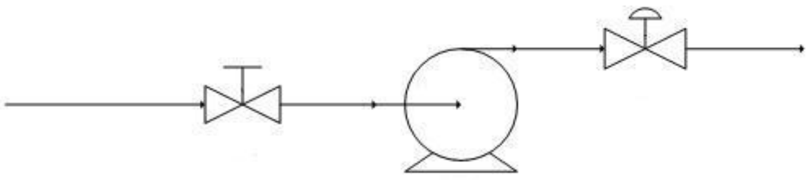


This eliminated excess pressure that might build up inside a reactor.

Isolation valves

The isolation valve is used to isolate a portion from the system when inspection, repair or maintenance is required. Isolation valves are placed around the junctions in the distribution system. They are also part of the absolute requirement for P&ID construction. Engineers should be aware of their system's pressure, voltage, process medium, pipe sizing and flow rates when selecting the right isolation valves that work under the conditions for their system.

If engineers forget to add isolation valves on their P&ID design, serious problems would occur when the system needs to be partially shut down for maintance or other reasons. There would be no way to control the unit's operation other than existing flow valves. Some upstream production problems could affect downstream production since appropriate isolation valves are not installed for safety and production purposes.



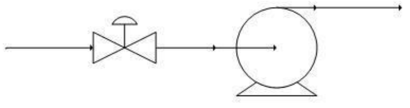
Isolation Valve Placement

Notice that the automatic valve after the pump is a flow valve, while the manual valve before the pump is an isolation valve. One example of this use is for cleaning. If some parts of the pump cannot be cleaned in place, the pump will need to be taken apart. The isolation valve can cut off the flow to the pump, allowing it to be safely taken apart.

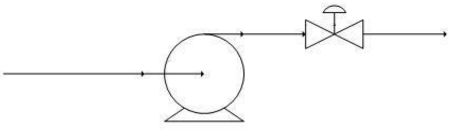
Valves and pumps

Some of the most commonly used pieces of process equipment that show up on P&IDs are valves. Valves control the flow of fluid through pipes by opening to allow flow or closing to stop flow. One of the problems associated with valves on P&IDs is the sheer number of them and deciding where to place them in the process.

A common mistake with valve placement has to do with the interaction between valves and pumps. It is important that valves be placed **after** pumps in a given pipe. When a valve is placed before a pump, should the valve close and the pump has not been shut off, there will not be a constant supply of fluid to pump. This is known as *starving* the pump. Starving the pump can create a disturbance known as cavitation, and it is very hard on the equipment and can cause the pump to break. Placing valves after the pump ensure that even if the pump does not shut off, it is still filled with the proper fluid.



Improper Valve Placement

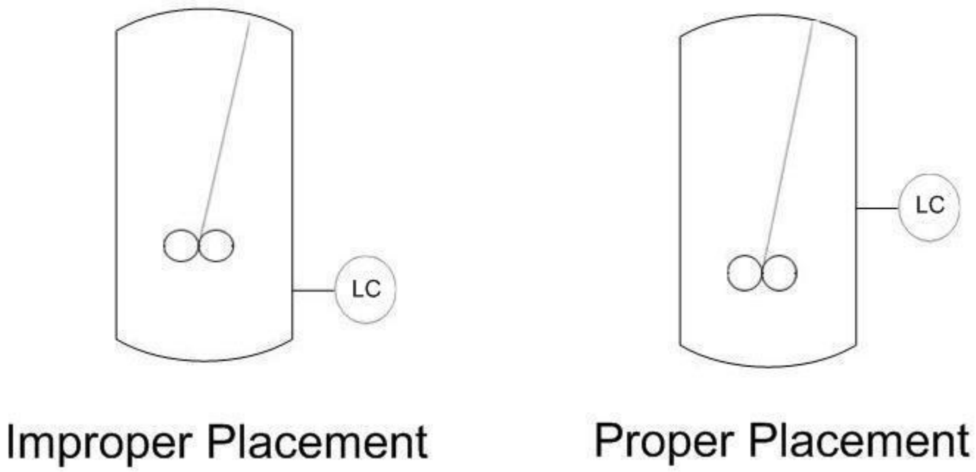


Proper Valve Placement

These same principles apply to valve placement with respect to compressors. Placing a control valve upstream of a compressor eliminates your ability to control pressure in the pipeline downstream of the compressor and introduces a risk of starving the compressor.

Agitators

A point that is very easy to miss, and very expensive if it is missed, is that if a vessel is equipped with an agitator, the vessel **must** be filled enough to submerge the agitator before the motor is turned on. Agitators are designed to operate against the resistance of fluid. Without this resistance, the agitator will rotate much faster than the equipment is designed for. This rotation may result in hitting the harmonic frequency of the structure, compromising its integrity and causing the agitator to rip from its foundation. This creates not only a fiscal predicament (not only ruining the agitator but also the vessel), but a safety nightmare. When designing a process, one must make sure he or she knows and accounts for how much fluid must be in an agitated vessel to avoid this situation. This can easily be solved by adding a level sensor to the tank that the agitator is being used in.



When placing the level sensor on the tank, make sure to place the sensor above the level of the agitator. This will ensure that the agitator is submerged in the fluid. It would be incorrect to place the level sensor below the agitator.

Instrument selection and placement

Instruments are designed to operate properly under specific conditions. Every instrument has a range over which it functions properly, and instruments must be selected that are appropriate for their applications. For example, a pressure gauge might have a working range of 5 psig - 50 psig. You would not want to use this gauge for sensitive measurements in the range 3 - 6 psig. Instrument material must also be considered during the selection process. If the substance being monitored is corrosive, for example, the instrument must be made of a corrosion-resistant material.

Once an appropriate instrument has been selected, it must be appropriately placed. For example, a level control is not useful in a pipe because there is no need to measure any water level inside of a pipe, much like a flow controller is not useful in a storage tank because there is no flow. Similarly, a flow controller should not be placed on a valve, but instead downsteam from the valve. However, level controls **are** useful in storage tanks and reactors, while flow controllers are useful in pipelines. Instruments must be selected and placed to reliably provide useful information and to accurately control the process.

Equipment Selection

When creating a P&ID, the equipment that is selected to be used is very important, not only to maintain a smooth process but also for safety purposes. Each and every piece of equipment from 100,000 liter storage tanks to temperature sensors has *Operational Limitations.* These are the conditions under which a given piece of equipment operates as expected, with safe, consistent, and reproducible results. For example, when storing a highly pressurized gas at 2,000 psig, one wouldn't want to use a storage tank that has been pressure tested up to 3 psig. The process conditions are way outside the operational limitations and would pose a serious safety hazard. The same goes for sensors and gauges of all types. On a pressurized vessel at 2,000 psig, it would be no good to use a pressure control system that has a sensor that is meant to measure up to 100 psig, the results would not be accurate.

Operational limitations can usually be found in the equipment manual sent by the manufacturer or possibly on the manufacturer's website. If it is not found in either place, an engineer is obligated to contact the manufacturer and find the operational limitations before using a piece of equipment in a process. More can be found on equipment selection for some (but by no means all) equipment below:

Temperature Sensors

Pressure Sensors

Level Sensors

Flow Sensors

Composition Sensors

pH and Viscosity Sensors

Biosensors

Valves

Process Design

Unit operation input/output

Providing clearly specified inputs and outputs to and from process units is vital. For the safety of the system, proper control and the prevention of disaster, it is important to show where each substance came from and where it is going. The P&ID must show all material streams to and from separation units, heat exchangers, and reactors.

For example, if the P&ID is not clearly denoted in a reaction to create construction explosive materials, then a large and perhaps fatal calamity can occur at the chemical plant. The exiting streams of a plant may pour into a river which may violate environmental regulations. Therefore, specifying inputs and outputs is imperative in a P&ID.

Pressure and flow

The movement of fluid is essential in many production lines. Transporting material from a tank to a reactor, a reactor to a distillation column or from a column into a tanker truck all involve the movement of fluids. When designing P&IDs, a process engineer must decide how they are going to attack this problem, namely whether or not a pump is needed to move the fluid.

The main issue here is pressure. Fluid, both gaseous and liquid, moves down the pressure gradient from high pressure to low pressure. The rule of thumb is that if the source of the fluid has a much higher pressure than the destination of the fluid, a pump is not needed. Even if the source has a pressure only a fraction higher than that of the destination, a pump may yet be needed for the fluid to flow through the pipes, from the source to the destination. However, the flow and pressure from the source must also be compared to the needs of the process. For example, if a inlet flow has highly fluctuating pressure, a pump or valve should be used to regulate the pressure for the process. Also, if the diameter of the pipe does not remain constant and is fluctuating as well, perhaps due to outside pressure or force, a pump should be used to control the pressure inside the system.

A common mistake young process engineers make is when charging a pressurized vessel, they do not use a pump powerful enough to overcome that pressure. This causes *backflow* and can ruin process equipment. This problem is especially prevalent in recycle streams. In some situations it may be appropriate to use a check valve (a valve allowing fluid to flow through into it from only one direction), to add an additional barrier against backflow. An example of such a situation would be when a fuel is to be mixed with air for combustion: if any air were to backflow to the fuel source, a dangerous, explosive situation would arise. In this case the use of a check valve would be appropriate.

An example of when a pump is never needed is when liquid from one vessel at ambient pressure is being transported to another vessel at ambient pressure that is at a lower elevation. This is known as a *gravity feed* and utilizing gravity feeds where possible can significantly decrease the cost of a process. As long as the pressure at the exit of the pipe is lower than the pressure created by gravity, a gravity feed can be used. An example of such a gravity feed is the flushing of a toilet, by which the water from the cistern at a higher elevation falls to the water closet at the lower elevation.

Underspecification

For safety and control purposes, **redundancy** is desirable in control systems. A process lacking in suitable redundancy would be **underspecified**. For example, a reactor containing an exothermic reaction would be underspecified if it only contained a temperature control. The potential danger posed by an exothermic reaction warrents a high degree of safety: if one controller were to malfunction, a backup must be in place. A pressure control should also be included, and perhaps a level control as well depending on the specific process.

Overspecification

On the flipside of underspecification is **overspecification**. Adding too many controllers and valves on vessels and lines is unnecessary and costly. Since process control is highly dependent upon the situation, it is difficult to set specific limits on the necessary amount of specification. However, a controller should not be used unless there is a specific need for it. Misused redundancy is a common example of overspecification. It is unlikely that a water storage tank needs level, temperature, and pressure controllers. A degree of freedom analysis, as shown in a distillation column example, can be helpful in determining the exact number of controllers, sensors, actuators, and valves that will be necessary to adequately control a process.

Problems When Using a P&ID

Other than equipment and production pitfalls, there are also some general P&ID interpration and inherent problems.

Interpretation of P&IDs

Although it is essential for P&IDs to represent the right instruments with the right references, many P&IDs do not support scaling or do not require scaling as part of the system. P&IDs are drawn in a way that equipment and piping are displayed for ease of interpretation. They do not show the physical placement and location of different systems or the actual sizes and length of equipment and pipelines.

Experience tells us that many engineers have overlooked certain pieces of equipment or over-estimated the size of equipment and piping, leading to calculation and construction errors. For example, an engineer could be designing new equipment on the existing P&ID and not realize that existing equipment is so densely populated that extra equipment will not fit. It is important to check the physical space of an area before adding equipment to a process.

Inherent Problems of P&IDs

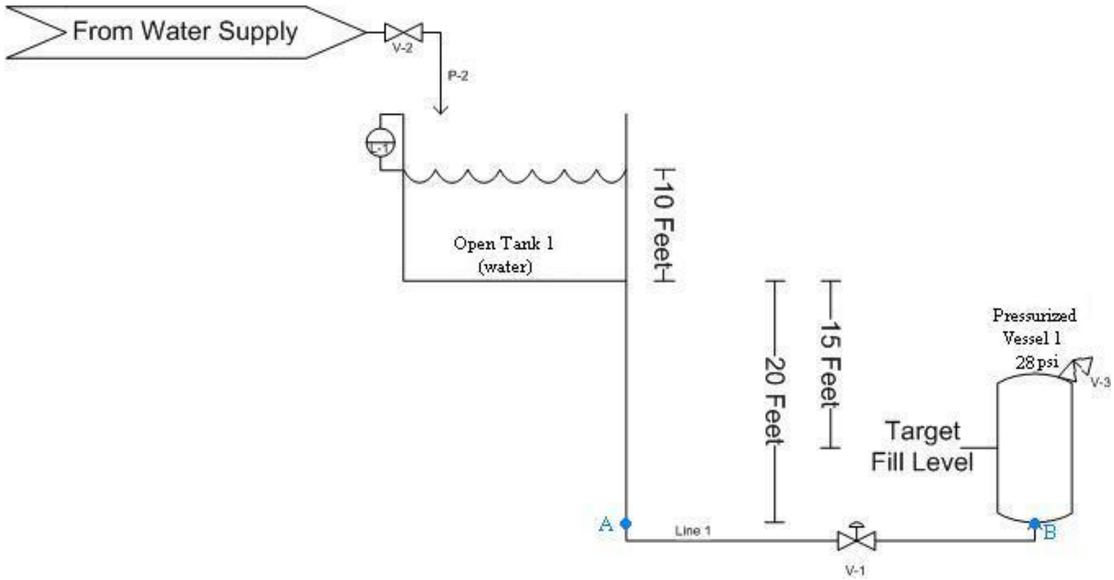
There are a few problems with P&IDs that are only solved by being conscious of them. One problem is that P&IDs are constantly being updated, revised, changed, and added to. When reading a P&ID it is always important to check the date it was last revised and if there is a later revision available. Making adjustments to equipment that isn't there anymore not only causes confusion and frustration, but is a waste of everyone's time. Moreover, making sure you are using the most recent edition to the P&ID will ensure that you do not purchase equipment to install that has already been installed.

Usually, companies will have a computer database with the most current P&IDs. Before modifying or working with an old paper copy of a P&ID, check to make sure it is up to date with the most current revisions.

Another problem with P&IDs is that even if they are well made and technically perfect, complex processes often appear cluttered. There is an inordinate amount of information contained in a P&ID, and as such P&IDs are next to impossible to take in at a glance. It is important to carefully study each document to fully understand the process. The mistake of assuming you gained all the information off of a P&ID from simply scanning it can leave you with mental holes that you don't even realize you have. This can be avoided by reading the P&ID carefully on the first pass and constantly referring back to it when questions arise.

Worked out Example 1

A pipe connects a water tank (open to the atmosphere) and a vessel pressurized to 28psi. The open tank has 10 feet of water in it. A level control system ensures that the 10 foot level is maintained in the open tank. The bottom of the pressurized vessel is 20 feet below the bottom of the open tank and starts with no water in it. The goal is to fill the pressurized vessel up to 5 feet. Due to a poor design by the project engineer, the water is fed into the bottom of the pressurized vessel. Given that the density of water is 62.4 lbm/ft^3 and the gravitational constant is 32.2 ft/s^2, is a pump needed? If so, where should it be placed? Assume that there is no pressure drop due to friction against the pipe and that the air pressure of the pressurized tank remains at a constant 15psi. The figure below may be helpful in visualizing the process. (Hint: Remember in English units, a “fudge factor” is needed when using P = ρ*g*h)



Answer: First, we must determine the pressure at the stream exit (point B). When the vessel is full, we know that the pressure created by the water is:

$$P = \frac{\rho g h}{32.2}$$

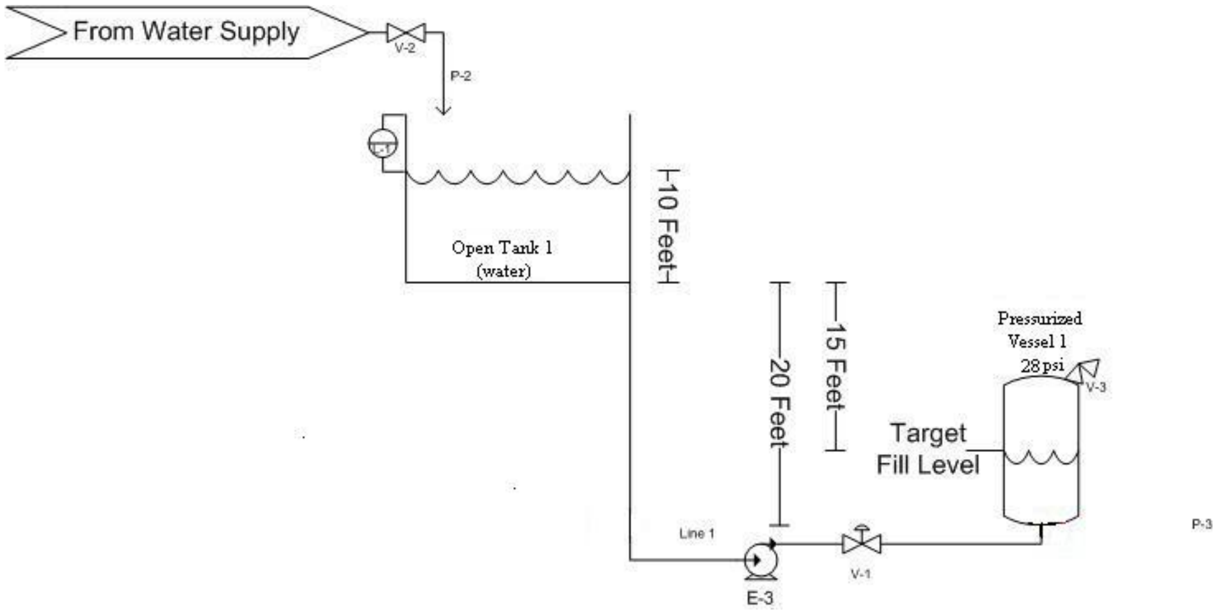
Plugging in 62.4 for the density, 32.2 for g and 5 for h, we find that the pressure caused by the water is 312 lbf/ft^2. Dividing by 144 to convert to psi, we see this is equivalent to 2.2 psi. When we add the 28psi from the air pressure in the vessel, we get a grand total of 30.2 psi at the pipe exit.

The pressure in the feed pipe at the height of the exit to the pressurized vessel (point A) is again given by:

$$P = \frac{\rho g h}{32.2}$$

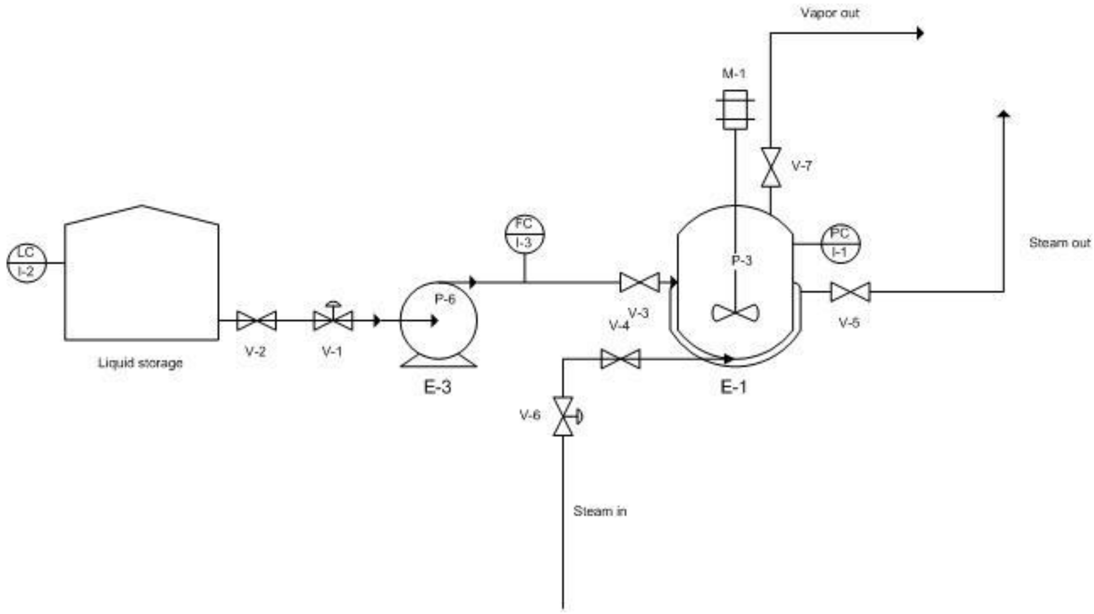
Only this time, the top of the water is 30 feet above the pipe exit instead of 5 feet. Plugging in this value, we find that the pressure is 1872 lbf/ft2. After converting, we find this is equivalent to 13psi. Since the water tank is open, we must also add 14.7psi for atmospheric pressure. This gives us a total pressure of 27.7psi.

The pressure of the feed at point A in the diagram is LESS than the pressure at point B, so a pump is needed. The pump should be placed BEFORE the valve labelled V-1 on the pipe as shown below. This is to prevent starving the pump when the valve closes.



Worked out Example 2

There are several errors in the following P&ID, which shows a vessel where a liquid is vaporized using heat transfered from steam passing through a surrounding jacket. Identify the errors and explain why they are mistakes.



Answer:

- 1. Control valve V-1 should be after the pump, not before. Placing a pump after a control valve takes away your control of the flow rate and can also damage the pump if the valve is shut off while the pump is running.
- 2. Vessel E-1 is only being monitored with a pressure control. This is not adequate redundancy for process control. There should also be a temperature control (to make sure the vessel is at an appropriate temperature for vaporization) and a level control (to make sure the agitator is always submerged in liquid).
- 3. Although the vaporization takes place at ambient pressure, there should be a safety valve on the vessel. What if there is a malfunction and too much vapor is produced?
- 4. The input and output streams for vessel E-1 are not sufficiently labeled. The source of input streams and destination of output streams should be specified to ensure adequate control of a process.
- 5. There should be a flow meter and control valve on the vapor stream exiting the vaporization vessel so that you can monitor and control the amount of vapor exiting.

PIDSafetyFeatures

Title:Saftey Features in Chemical Processes

Note: Video lecture available for this section!

Authors:Bradley Anderson, Aaron Bennick, Michael Salciccioli Stewards: Jocelyn Anleitner, Stephanie Combs, Diane Feldkamp, Heeral Sheth

Date Presented:8/3/06 Date Revised:

- First round reviews for this page
- Rebuttal for this page

Contents

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Introduction

Safety has become integral to the manufacturing world. The implementation of proper safety techniques and accident prevention can not only save time and money, but prevent personal injury as well. P&IDs, when properly utilized, are a powerful resource to identify safety hazards within the plant operations. The following sections provide an overview for the safety hazards that exist within a process, and illustrate the importance of P&IDs in a chemical plant.

Safety in Design

During the early stages of plant design it is critical to determine important safety features that remove potential hazards from effecting the facility environment. Regulations require that plant designers play a major role in minimizing the risks associated with these hazards. However, in order to do so, designers need to be aware of the hazards that exist during plant activity. The facility design team must develop a detailed drawing (P&IDs) including specifications of the plant process and environment to ensure that every aspect with regards to safety is covered.

Hazard vs Risk

When discussing safety, the terms hazard and risk are often used interchangeably. However, the difference in definition between the two terms is critical in utilizing the information they provide in increasing the safety within a plant. Hazard is defined as a potential source of danger and risk is defined as the level of threat associated with the hazard. A risk of a hazard occurring can be represented mathematically by the following equation:

$$Risk = Frequency \times Consequences \quad \textbf{(1)}$$

Frequency represents the probability that a hazard will occur, and consequence represents the impact of that hazard. Values for each parameter of the equation above are assigned by using either experimental information or educated judgment based on engineering models and experience. Each plant design process will have specific safety hazards and risks associated with it. Therefore, there is no predetermined value that can be assigned to each variable. For example, in one situation, a water tower may be well supported inside the plant facility, and in another situation a water tower of similar structure may be hoisted against a rusted frame work outside of the plant facility. Both of these situations have different levels of risk associated with them. Needless to say, the risks associated with a plant setup can be reduced substantially by minimizing the probability that a hazard will occur.

Hazard Locations and Risk Hotspots

Common hazard locations exist in any place containing large amounts of energy. The degree of danger is proportional to the amount of energy stored at that location. Risk can be directly linked to kinetic energy, potential energy, work, heat, enthalpy and internal energy sources. Kinetic energy, otherwise known as energy in motion, is present in any moving component. The component may be vibrating, rotating, or translating, and this motion causes the kinetic energy of that part to greatly increase. Within industry, personal injuries and fatalities are common hazards associated with moving parts. Reactors and cooling towers placed in high locations contain potential energy. If a structural failure were to occur within a plant, these structural units and their contents could fall from a large elevation onto another processing system or a human being, releasing all the chemical contents stored inside. Similarly, the stored work in springs and other devices can cause fatigue and wear on the mechanical system over time, and result in eventual machine failure. Heat released from a reaction within a chemical reactor can be rapid and fatal if not accounted for. The buildup of heat can cause serious consequences with runaway reactions and boiler explosions. The enthalpy and internal energy of a reaction typically are the cause of most runaway reactions and destructive fires in a plant.

Risk Hotspots

Uncontrolled chemical hotspots created within a chemical plant are a common source of hazard, besides the hazard locations pertaining to energy sources such as kinetic and potential energy. Risk hotspots mostly occurs in the piping system and associated valves of the system, joints, traps, and various other piping elements. Possible malfunction of the system due to structural corrosion can be triggered by the failure to maintain the piping systems efficiently and periodically. Even if one of the valves in the system has corroded and is unable to function properly, the fluid flowing through the pipe might get trapped, and the resultant buildup of pressure in the pipe may cause major safety hazards including fatal ones such as an explosion. If the system is not shut off before the pressure gets out of control, the pressure buildup in the pipe will cause it to burst, releasing all of its internal contents to the surrounding environment.

Storage vessels are other pieces of equipment that must be cared for properly. Since the plant operators and engineers do not usually interact with the storage vessels, as compared to other parts of the plant process such as the piping system, they are considered to be of secondary importance and commonly overlooked. Storage vessels have much more content inside them than pipes, so a leak or a burst would take longer to get back under control than pipes, which can by plugged more easily. Problems which may arise from storage vessels are not associated with their design, but in fact from not thoroughly and periodically maintaining them. Possible complications that can arise from neglecting the storage tanks are over-pressurization, overfilling, heating element malfunction, or simply equipment malfunction.

Chemical reactors are another common location where risk hotspots occur. The nature and design of commercial chemical reactors is to handle a controlled explosion. However, if the control element is removed, disaster is bound to occur, just as in any other part of the chemical plant. The most common type of hazard is a runaway reaction inside a batch reactor. When the plant facility looses electricity or cooling water, a runaway reaction will strike inside the reactor. Once a runaway reaction has spun up inside the reactor, many other hazards may follow, such as flow reversal in the pipes, incorrect reagent charging, heat exchanger failures, and external fires. Other hazards may perhaps be even more serious, such as engineering errors that could potentially cause a runaway reaction to occur, including inappropriate material selection, inadequate equipment inspection or failure to fully understand the chemistry or exothermic nature of a reaction. A runaway reaction originating inside the chemical reactor can easily cause a chain reaction across the rest of the equipment at the facility, and can result in the entire system malfunctioning.

Other process equipment that may be hazardous and where risk hotspots commonly arise, are vacuum operators, furnaces, pumps, gas movers, compressors, and heat exchangers. The location and type of specific piping and unit operations are available on the process P&ID. A responsible process engineer should use the P&ID to identify all risk hotspots, and act accordingly to monitor and maintain a safe working environment. In addition, a standardized plan should be constructed so that in the event of a malfunction, the correct steps can be taken to bring the faulty part back under control. Supplementary precautions should be taken to prevent a comparatively minor malfunction in the system from becoming a disaster which may violate environmental regulations and even endanger human lives.

Safe Design Principles

The ISD or Inherently Safer Design movement was a doctrine striving for safer chemical processing procedures. This movement was pioneered by Trevor Kletz in 1976, and promotes the design of processes so safe, that no catastrophic failure can occur within the plant. The following principles apply to initial process design:

1. Use the fewest number of hazardous substances in the smallest quantities and still maintaining plant productivity

2. When possible, substitute hazardous chemicals with chemicals that are less dangerous

3. Practice moderate operating conditions in the plant

4. Use the simplest plant design possible

5. Design equipment in the plant to minimize the effects of a hazardous incident

The specifications determined by the process designers are communicated through the P&ID.

Hazards in Construction

In order to eliminate hazards, a operations personnel must be able to identify that a hazard exists. Hazards that may be encountered on plant sites may be categorized into three main types:

A) Hazards harmful to health

B) Hazards likely to cause personal injury

C) Hazards likely to lead to catastrophic events

Hazards harmful to health:

When workers are exposed to or come in contact with asbestos, corrosives, irritants, toxins, or noxious gases try to avoid by specifying the processes, which lead to this exposure

Hazards likely to cause personal injury:

Hazard awareness is increased when people have to work in situations likely to expose them to the risk of personal injury, including moving plant machinery or working in areas where objects are likely to fall. Situations where there are live electrical circuits overhead, buried power lines, and confined working conditions are likely to cause personal injury.

Hazards likely to lead to a catastrophic event:

These hazards have consequences beyond the site boundary. They include fire outbreaks, explosions, flooding, or premature collapse of structures, cranes, tunnels and excavations.

Fail-Safe Design

The fail-safe design of a unit operation (such as a reaction vessel) requires a complete understanding of the operation at hand, and the knowledge of all the worst-case conditions. A fail-safe system is a unit operation such that, if any or all of the worst-case conditions were to occur, the operation would shut itself down automatically and in a safe fashion. In the case of a run-away reaction, if reagent feed limits, interlocking controls, and integrated heat balances are all properly maintained, the reaction cannot "run away." Other precautions such as purges, vents, dump tanks and quenches are available for reaction vessels, and should be visible on the P&ID.

Inherent Safer Predesign

The table below provides a guideline for identifying and minimizing hazards partly based upon Kletz's rules for ISD in the Safe Design Principles section above. The step/rule column describes the action taken. The tools column describes the mechanism by which the actions are taken. The experimental and analytical resources column describes the knowledge by which the mechanisms are created. The Literature References column describes where the knowledge can be found.

Step/Rule	Tools	Experimental and Analytical Resources	Literature References
1. Identify	Fire triangle, Flammability-limit chart, Chemical	Thermodynamic calculations, Reactor	Perry's Handbook (10) Table 26-10

	reactivity chart, Safety compatibility chart, Safety stream chart, Toxicity ratings	design equations, calorimetry, flammability charts	and associated pages
2. Eradicate (ISD Rule 2)	Inert-gas blanketing, Failsafe design	see above	see above
3. Minimize, Simplify, Moderate, Attenuate (ISD Rules 1, 3, 5)	Moderate ignition sources, Keep reactive volumes small	Logical chemical engineering analytical and computational skills	Merk Index, MSDSs, Chemical Engineering textbooks and monographs
4. Isolate (ISD Rule 4)	Separate hazardous operations - Surround in an impenetrable structure	Logical chemical engineering judgment	

Relief Systems

Organizations such as the American Society of Mechanical Engineers, American Petroleum Institute and National Fire Protection Association layout recommendations and design standards so that most engineers with proper training can setup proper emergency relief systems for single phase flow. Unfortunately, this is often not good enough for reactive systems. These systems are much more complex and include multiple phase flow, runaway reaction potential and self reactant material. When designing an emergency relief system (ERS) it is necessary to understand all aspects of the chemicals and processes that will be in play. This includes but is not limited to: kinetics of the possible reactions, contamination, interactions with air, rust, piping, or water, phase changes and runaway reactions. The follwing topics deal with the hazards that require designing relief systems and the prevention of runaway reactions.

Planning and Design

When designing an ERS, it is important to consider the worst-case scenario. This is based upon a thorough knowledge of the reactions, materials, and environment of the process. Some of the most critical scenarios are over-pressurization in a reactor.

All disaster scenarios can be analyzed using a hazard and operability (HAZOP) study. This HAZOP study will analyze a process based upon human, equipment, and environmental factors.

The HAZOP team must take the following steps to ensure that all potential scenarios are taken into account:

• Review the potential hazards of all chemicals. This includes non-operational conditions and interactions with contaminants

• Study the chemical process including all possible reactions, rearrangements, decompositions, etc.

• Review the P&ID's for the process

• Study the specific reactor and storage vessels for material composition compatibility, size, surface area, instrument ranges, and set points

With this information, the HAZOP team can determine most of the potential disaster scenarios. Most likely the worst case scenario will involve fire induced runaway reactions.

Design Strategies

There are various techniques often used to prevent failure from over-pressurization, fire, runaway reaction or other disaster scenarios. The following are some design suggestions that will decrease this risk:

-The use of insulation in case of fire. Unfortunately, insulation will also minimize heat loss from the reactor during a runaway reaction. It is important to consider the ability of a vessel to drain when using insulation. If the contents can be drained and cooled before the reaction starts to runaway, catastrophe can be prevented.

- Design to avoid fire damage on sensitive equipment. The possibility of fire damage on electrical equipment or sensors will cause control difficulties when trying to slow down a runaway reaction, or monitor another emergency situation.

- Consider the structural integrity of the entire system due to fire damage.

- Use multiple purge streams and valves to separate materials. Separating reaction components into a storage vessel with a large surface area will allow for faster cooling.

- Install two separate relief devices in case one has been compromised due to fouling or solid particle blockage.

Overall safety relief plans must be made for all possible scenarios.

Reaction Kinetics

Reaction kinetics usually determine the potential for safety disaster. For the average exothermic reaction, the reaction rate doubles with every 10°C increase. This will lead to an exponential increase in energy which will force both the temperature and pressure in a system to uncontrollable levels. Since reaction rates are sensitive not only to temperature, but also pressure, contaminants, concentrations and phases, all possibilities must be adequately analyzed for plant operation to be deemed safe.

Relief devices should be designed to handle chemicals and/or mixtures in any phase. For example, if a reactor temperature increases suddenly and a safety relief valve is activated the material must be able to flow through the valve quickly and safely. If the chemical has changed phases, or has a higher pressure that the valve cannot accommodate, the chemical will not be able to escape and build up pressure in a reactor. This would greatly increase the risk of a disaster. Because of the extent of possibilities when considering reaction rates and kinetics, it is usually necessary to consult some sort of computer simulation or dynamic simulation tool to plan for every possible scenario.

Runaway Reactions

Runaway reactions are caused when exothermic reactions are fed more energy due to malfunctioning cooling systems. This causes an exponential increase in temperature, which in turn causes an increase in pressure, and finally damage to the reactor and/or plant. The possibility of malfunctioning cooling systems must always be considered for flammable materials. If materials are being stored at temperatures above their flash point, fire is always a possibility. Fire will cause a reaction to reach runaway conditions with very little reactant. Relief systems for fire induced runaway reactions must be larger than conventional runaway reactions.

While relief systems are often designed for over-pressurization, this might not be enough when considering flammable materials. Constant or prolonged exposure to flame will cause most normal reactant or storage vessels to fail causing chemical leaks or plant fires. Fire proof insulation must be used together with normal pressure relief systems to prevent system failure. An important design variable when considering fire induced runaway reactions is fire flux. This variable considers heat impact on a reactor due to fire. Formulas for calculating fire flux can be obtained from the National Fire Protection Association. The elevation of flames must also be considered when designing reactor vessels and safety relief systems. Pool fires can produce flames that are hundreds of feet high; using the P&ID will be important in ensuring that process components sensitive to fire are sufficiently protected from something like pool fires.

Two Phase Flow

For systems in which multiple phase flow is possible, all phases must be planned for. For vapor hybrid systems, all-vapor flow should be considered. Likewise, when foam flow is possible, all-foam flow should be planned for. This is all part of the mindset of planning for the worst case scenario. The most conservative design should always be used when faced with uncertainties of phase flow.

Often times, a runaway reaction will cause high-viscosity two phase flow. Relief valves and safety features must be ready for viscous flow. Many resources suggest averaging the viscosities of the two phases to plan for two phase flow. It is important to keep in mind however, that two phase flow discharge will separate in the discharge line. This will lead to higher pressure drop. Piping sizes are often underestimated due to this unplanned scenario. Undersized piping will lead to valve rupture and back pressure and could cause venting disturbances during the emergencies where venting is critical to the safe shutdown of a process.

Regulations

There are often conflicts and contradictions in federal regulations and recommended practices for safety design guidelines. When faced with such confusion the designer should at least design to the level of the Occupational Safety and Health Administration's (OSHA) requirements, as well as meeting regulations set by the Environmental Protection Agency (EPA), the Process Safety Management(PSM), and the Risk Management Program (RMP). While these guidelines and regulations produce a minimum standard to follow, when dealing with reactive systems, this does not always meet process needs. For more details on the various regulatory agencies that govern chemical processes, see Regulatory Agencies and Compliance.

Additional Safety Hazards in Chemical Plants

This section outlines potential safety hazards commonly found in places where chemicals are stored or chemical processes are taking place. While these hazards are rather easily prevented through attention to detail and general awareness, neglecting them can have catastrophic consequences.

Time Sensitive Chemicals

Some chemicals have a "shelf life," or an expiration date provided by the manufacturer. The chemical must be used by this date or properly discarded. These are typically reactive chemicals, which can become unstable after a certain period of time, possibly rupturing the vessel in which they are stored. An example of this is a monomer that begins to polymerize unless an inhibitor is present. This inhibitor is completely consumed after a certain period of time, allowing polymerization to occur, and therefore must be used or discarded by that time. Another example deals with the formation of peroxide, which can be a severe fire and explosion hazard. It may also be a health concern, causing severe mucous membrane, respiratory tract, skin, and/or eye burns. Peroxide-forming materials should be stored carefully, labeled with the date received and the date first opened on the container. Chemicals should be disposed of or checked for peroxide formation after six months; do not open any container with solid formation around the lid. There are several ways to prevent this from happening. Material Safety and Data Sheets (MSDS) should be available for any chemical the plant uses. These can inform you if a particular chemical becomes unstable after a certain period of time. It is also important to make sure there are procedures in place for handling time-sensitive materials. Lastly, investigating near-miss occurrences can help to ensure that future incidents do not occur.

Pressure Relief Systems

Any open pipe in a chemical plant is a potential discharge site. While operator convenience and maintenance remains a concern, safety takes precedence when dealing with relief systems. It is essential to operator safety that relief valve discharge sites are located in areas that pose a low risk of exposing personnel to chemical hazards and are directed away from all access platforms. It is common in industry to tie multiple pressure relief devices to an emergency vent header that releases on the roof of the plant. Long stretches of unsupported pipes also pose a potential threat. Force generated by material flow could bend or break the pipe impeding on plant operations, and more importantly, injuring personnel.

Dust Explosion Hazards

Most flammable solids can form an explosive dust cloud if the particles are small enough. Materials such as wood, grain, sugar, plastics, and many metals can all form these explosive dust clouds. Dust explosions occur when a combustible material accumulates in a confined area and is exposed to an ignition source. High risks areas are usually those that are neglected such as tops of vessels and tanks, on pipes, storage bins, bucket elevators, and dust collectors.

To prevent dust explosions, it is essential to implement good housekeeping practices. Care must be taken so that the cleaning process, such as sweeping, doesn't turn a dust layer into a more dangerous cloud. Operations and process engineers should be aware of all the mechanical and safety control equipment associated with preventing fires. Equipment like electric vacuums must be appropriate for use in an area where an explosive dust cloud could form. This entails no holes or cracks in the cord, sufficient grounding, and receiving site approval for use of the equipment.

Overfilling Tanks

The overfilling of vessels has long been a leading cause of serious incidents in chemical and petroleum industries. When a level sensor or high-level alarm fails, reactive material can spill over the tank and accumulate. If this material is exposed to an ignition source, there is the potential for an explosion, leading to serious property damage, environmental issues, and injury to operations personnel. When filling or draining a tank with material, operators should be aware of all relevant level, pressure and temperature controls in place, and watch for abnormal trends. Also, all safety critical alarms surrounding a vessel should be tested regularly at frequencies recommended in plant process-safety-management procedures. Conducting regular process maintenance on safety critical alarms have numerous benefits, which include reducing the risk of operating a plant with faulty equipment, and increasing operator awareness with the location and function of critical safety devices.

Containing Storage Tank Spills

Engineering controls are implemented into plant designs to account for potential disasters, such as a spill, leak, or complete emptying of a storage tank. Industrial-sized chemical plants store large amounts of raw materials, products, and byproducts on-site. The amount of each varies, but is typically between three to thirty (or more) days of the required supply or amount produced. The stored chemicals can be hazardous, flammable, explosive, and/or reactive with each other. In the event of a tank spilling, dikes are built around tanks to contain the spill and protect the surrounding community from the spill. The regulations of the dikes include the following: dike volume must be 1.5X the largest storage tank contained by the dike, reactive materials cannot be stored in the same dike, and scuba gear must be present on-site if any dike is deeper than four feet. Since the dikes cannot contain any reactive materials, the implementation of dikes affects the overall plant layout. Typically, dikes are designed to be like speed bumps and have a height less than one foot so fork-lifts and tankers can easily maneuver through the plant. Dikes are an effective engineering control that greatly improve the safety of a chemical plant.

Temperature and Pressure Ratings

Before a vessel is put into plant operation, it is rated and stamped by the manufacturer with temperature and pressure limits. Problems arise when personnel overlook the inverse relationship between temperature and pressure for gases. Be aware that equipment rated for a specific temperature and pressure, cannot be operated at the same pressure if the temperature is increased. While it may seem extremely intuitive (think ideal gas law), this relationship is too often overlooked, usually with serious consequences. When operating any process, pay attention to the temperature and pressure ratings. If they aren't readily known, review equipment files or contact the manufacturer before making any changes to the process. Also, operation and maintenance should always be performed according to strict standards laid out by a plant standard operating procedure (SOP). Any changes must be reviewed and approved by a cross functional team.

The above hazards and preventive measures have been outlined to illustrate how important safety is in a chemical plant, and the importance of being aware of your surroundings and all possible safety hazards. Many of the aforementioned topics may seem like common sense, but it is very easy to overlook small details in the scope of a large-scale chemical process. Keeping safety in mind at all times as the paramount of any process can ensure that people leave their shift the same way they arrived.

Alarms in Processes

Alarm configuration and specification is an important part in the design and operation of any chemical process. Alarms are implemented in a process design to aid in the control of the process. Federal and industrial documents only specifically reference alarms in the context of processes exceeding regulatory compliance limits. In this sense, alarms are used to control safety and environmental hazards. Other important uses of alarms are to control product yield, product quality, and operational limits of process equipment. This section will discuss the steps taken to implement alarms in processes, common different levels of alarms, and common instances in which alarms are useful to comply with regulations.

Alarm Lifecycle

When it has been determined that an alarm is needed to aid in the control of a process, the alarm must be specified. The following are basic steps in implementing an alarm.

1. First, the process designer needs to know what category the alarm fits into. This is important because responses from different categories of alarms are usually managed differently. They may be prioritized in case multiple alarms occur at once, so that the proper follow-up reports and procedures may be taken accordingly. Common categories include product quality parameters, safety, environmental considerations, and equipment protection.

2. After determining the specific use of the alarm, the limits must be set. If the process variable exceeds the set limits, the alarm will be triggered. (More information on limits will be discussed in Alarm Levels.)

3. Next, the computer system for the alarm must be configured. The computer system may contain logic loops that automatically change control parameters to offset the problem and merely inform the operator that it has done so. Alternatively, an alarm may trigger horns, flashing lights, or send a page to an operator, alerting them that there is a problem that needs attention.

4. The correct user response and interaction must then be defined for the alarm. This includes providing proper training for the particular process, procedures and operator manuals describing how the event should be investigated, guidelines on when action needs to be taken, and guidelines on when to escalate the situation to a more serious event.

Proper communication from the automated alarm system is critical. Care needs to be taken that alarms only signify abnormal conditions that require a response. For example, the successful completion of a batch operation is an important piece of information and should generate a computer message so that the operator knows. However, this does not represent an abnormal situation and should therefore not show up on the computer as an alarm. While this seems quite obvious, most industrial plants struggle with maintaining alarm systems free of "nuisance alarms." It is also important that alarms have proper descriptions so that the reason an alarm appears is clear. For example, an alarm could appear in a large plant that says, “LI-501 exceeds limits” by default. Although this might be useful to the engineer designing the alarm system as an indication that tank 501 has a high level of material, an operator that sees the alarm or other engineers working on the system may not know what the alarm is communicating. A more universally meaningful alarm indicator might say, “Material level in tank 501 is high.” With this simple change in the computer system, the alarm would be more effective for personnel to locate the problem quickly and act accordingly.

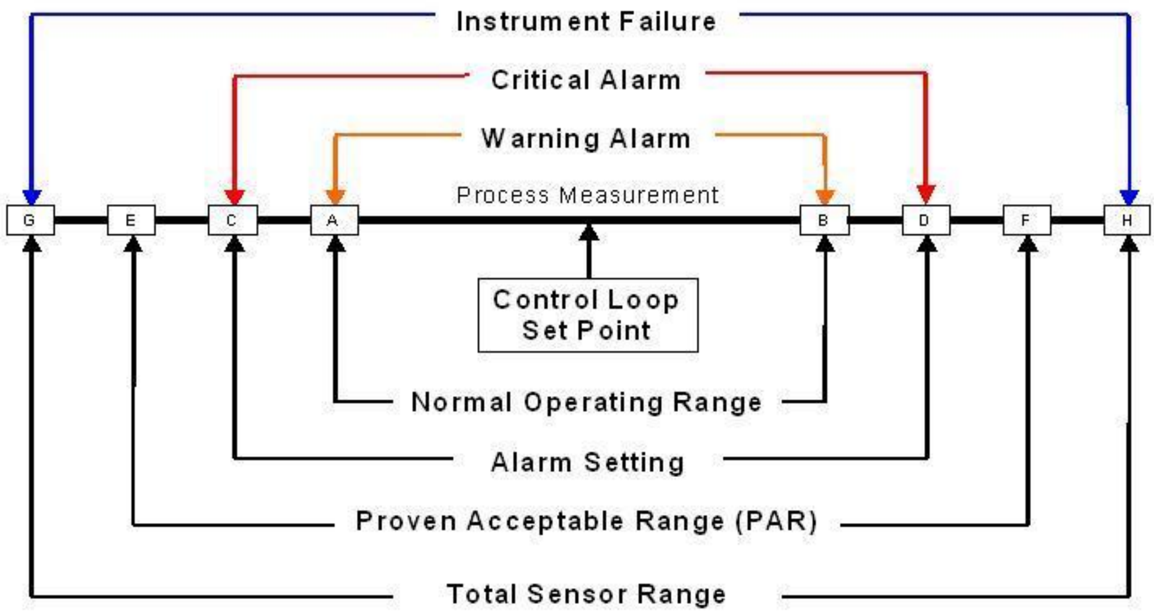
Alarm Levels

Alarms are available in a wide variety of types, with multiple levels of alarm. In all processes, disturbances occur that can shift a plant's operation away from normal. When this happens, measures are usually taken by computers, such as with the use of P&ID control loops, to keep the process under control. With these control systems, processes are designed to fall within a range of acceptable normal operating limits. When a process deviates beyond these normal limits, an alarm should be triggered.

For most processes, the minimum for safe operation is two levels of alarms: warning and critical. The warning alarm tells plant operators that the process has deviated beyond the acceptable limits and provides them with the time and ability to take corrective action so that the product quality is not affected and environmental and safety regulations are not exceeded. If the right actions are not taken or are not taken quickly enough to correct the problem, a critical alarm may then be triggered. The critical alarm tells the plant operators that conditions are dangerously close to breaching what is allowed. In many cases, the critical alarm will call for a systematic shut-down of the operation until the problems can be addressed.

The conditions at which warning and critical alarms are triggered are those conditions that exceed the limits determined for the process. Measurement uncertainty must always be considered because all devices in the control system will be subject to some possible error, even if it is small as +/- 1%. Setting an alarm at exactly the proven acceptable range for the process could allow a measured value to fall within this range, even though the actual value lies outside. This is called a "false acceptance." By performing error analysis and statistical distribution theory, the alarm limits can be adjusted as needed. This is a process called "guard banding," and it prevents real disturbances in the process from being ignored by the alarm system. Information about alarm limits should be well documented so if changes to the system are proposed, designers know how the limits were originally determined.

The figure below provides a visual representation of alarm ranges.



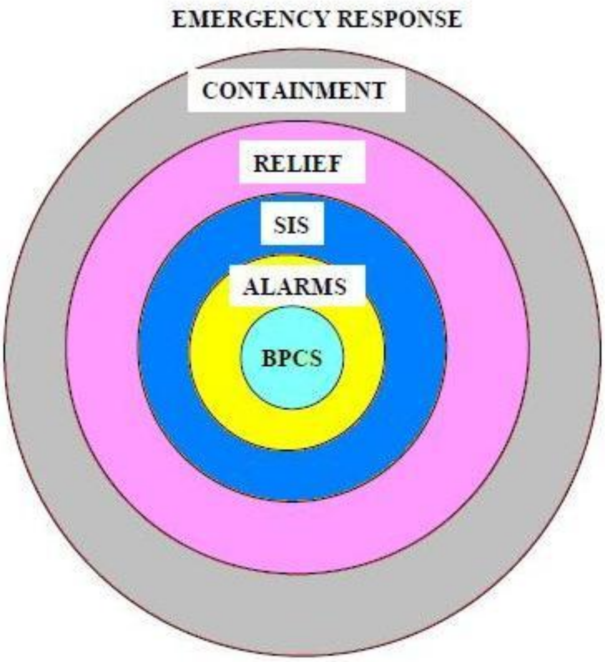
As seen in the center of the above figure, the control loop set point is the optimum point of control for the process (e.g. the optimum temperature and concentrations of reactants for a reaction). It is impossible to maintain the process at exactly this point, so there is a range of "normal" operation, inside which the process is still considered to be running in an acceptable way. The warning alarm would be triggered when the process goes outside of the limits of this range (lower than A or higher than B), allowing time for the process to be brought back under control. The critical alarm would then go off if the process goes beyond the alarm setting (lower than C or higher than D). This setting is determined by guard banding the process acceptable range (PAR) for uncertainty, seen in the figure by the fact that the alarm setting lies well inside the PAR. The spaces between E and C and between D and F are determined by the uncertainty. Lastly, this PAR must be inside the total range of the sensor that determines the instrument failure.

Alarms must be analyzed based on their priority:

- HIGH	= Hazard to people or equip., action required
- MEDIUM	= Loss of $$, close monitoring required
- LOW	= investigate when time available

SAFETY STEPS TO HIGH RELIABILITY

A safe system involves many layers of reponses when an incident occurs.



The center of the ring is the basic process control system.

The first layer of response is the alarm system which draws attention.

The second layer is the Safety Interlock System which can stop/start the equipment.

The third layer is the Relief system which leases pressure build-up in the sytem.

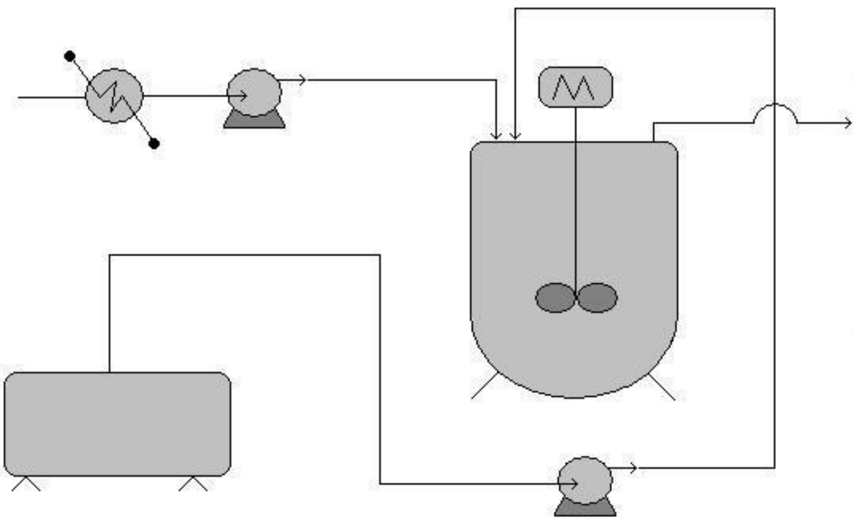
The fourth layer is containment which prevents material from reaching workers, community, or the environment.

The last layer to the ring is the emergency response system which involves evacuation, fire fighting, etc.

Worked out Example 1

A reagent recovery unit for a chemical process plant is being designed. The goal is to recover tin from a tin-plating waste stream through binary extraction with carbon tetrachloride. The equipment used in the process and P&ID are as follows:

- Carbon tetrachloride storage tank
- Mixing Vessel
- Pumps
- Heat Exchanger
- Associated piping



With knowledge of hazardous locations and risk hotspots, and safe design principles, analyze the flow diagram. Identify areas of risk and specific improvements that should be made to the process design before implementation begins. (note: no piping, valves, or controllers exist on the P&ID for the sake of clarity. Ignore these in your analysis) .

Areas of risk

Any piece of equipment could potentially become dangerous if the right situation were to arise. Each piece of equipment used in this process is no exception.

- Storage tank - These process units that are of secondary importance to the process goal, don't draw as much attention from plant operators and engineers, and tend to receive less maintenance.
- Mixing Vessel - The potential for heat buildup due to heat of mixing and the kinetic energy of the spinning motor makes this vessel a risk hotspot. Improper control of temperature within the vessel or motor speed (especially during periods when the tank may be empty) can lead to equipment malfunction or even explosions
- Pumps and Heat exchangers - The energy stored through pressurization by the pumps and potentially high temperature solutions handled by the heat exchanger present the possibility for danger. Wherever energy is stored, danger is associated with its potential release.
- Associated piping - Corrosion or failure to maintain pipes and associated elements are the main cause of the malfunction leading to danger. Leaks or total failures can release reactive materials on other equipment causing corrosion or malfunction

Specific Improvements

This process could be improved and made safer in the following ways.

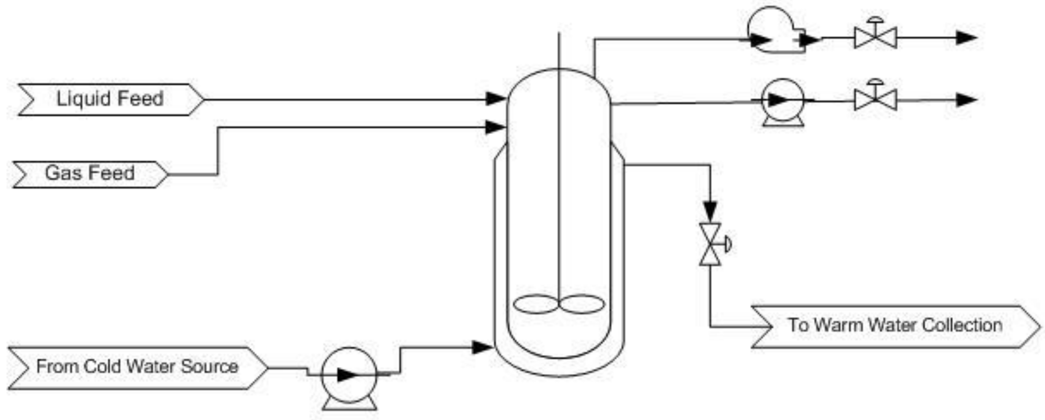
- Simplification of the piping used to connect the storage tank to the mixing vessel will reduce the risk of leaks and malfunctions due to piping complications.
- The addition of a emergency relief valve and temperature controlling/insulating element to the mixing vessel will greatly reduce the risk of equipment failure due to

overheating or overpressureization due to heats of mixing or outside heating influences.

- Use a reactor with a rounded top, as this type can withstand much higher pressures.
- The substitution of carbon tetrachloride with a less hazardous solvent, if the specific solubility required to extract the tin still exists, like cyclohexane, isopropyl alcohol, or 1,1,1-trichloroethane reduces the risk of health complications due to exposure, and possible explosions.

Example 2: Alarms in P&ID

A P&ID appears below for the production of a solution containing a pharmaceutically active compound. The reaction taking place in the CSTR is highly exothermic. After examining the P&ID for this part of the process, describe a possible alarm system.



Answer: The CSTR for the exothermic reaction is jacketed with a cooling water stream. An alarm should be in place to monitor the reactor temperature. A warning alarm can notify the operator that the temperature is too high and corrective action needs to be taken. A critical alarm should be in place to warn that the reactor is nearing runaway conditions and an immediate response is needed. If the necessary action is not taken, systematic shutdown of the reactor could occur. This would involve closing the valves, flooding the jacket with cooling water, and having the impeller on. Another possibility for an alarm, although we do not know how the products are being used specifically, is in a composition measurement of the product containing the pharmaceutically active compound. Depending on where this stream is going and how it is being used, too high a concentration could be dangerous if no other concentration-altering steps occur before the finished product goes out to consumers.

Actual Case

The bulletin below is taken for the US Chemical Safety and Hazard Invesitgation Board

Washington, DC, July 15, 2004 - *The U.S. Chemical Safety and Hazard Investigation Board (CSB) today released a Safety Bulletin calling on chemical plant and refinery* *operators to exercise appropriate caution when performing work on piping and equipment that could contain hazardous materials.*

The bulletin, set to be considered by the Board at a public meeting today, results from the CSB's investigation into a January 13, 2004, explosion and fire at the Huntsman Petrochemical facility in Port Neches, Texas, where two employees were seriously burned and significant damage occurred to nearby equipment. The explosion and fire occurred as workers attempted to purge a thousand-foot-long chemical process pipe in preparation for a cutting and welding operation.

Huntsman managers and workers were aware of the importance of completely removing hazardous material before cutting into the piping, and relevant warnings were contained in the company's written operating procedures. In this incident, workers first purged the piping with nitrogen to force out residual chemicals, including a hazardous mixture of peroxide and alcohol that reacts violently when heated. But unknown to the workers, the piping included a 300-foot-long section that was three feet lower than the rest of the piping, and despite the nitrogen purge, a significant amount of the hazardous mixture remained trapped. The next step in the operation was to use high-temperature steam to purge the piping of what workers believed would be a small amount of residual flammable hydrocarbon vapor. But the steam heated the peroxide that was trapped in the low section of piping. The peroxide then began to decompose, releasing heat and creating intense pressure. The pressure blew out a valve gasket and violently ruptured the pipe. Flammable vapors shot out of the openings and ignited into a large fireball, injuring plant workers.

After the accident, Huntsman found two drains in the low section of the pipe, which could have been used to remove the trapped liquid. Had Huntsman's procedures called for reviewing plant pipe drawings and physically walking the entire line within the work boundaries, the accident would likely have been avoided, CSB said.

Lessons Learned for this incident:

- *Physically examine all piping and components between isolation devices such as valves, and be sure piping drawings are current*
- *Use the drawings to identify key components, such as low-point drains that can be used to remove dangerous chemicals*
- *Prepare a specific written procedure for removing hazardous material and consider the consequences of working on piping that is not completely purged."*