CONTROLLING VESSELS and TANKS

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INTRODUCTION. It would seem that controlling a vessel should be a very simple matter -- They don't *really* do anything! But then, if they didn't do anything why are there so many of them? And why do they have so many different names? Going through a typical set of Piping and Instrumentation Diagrams (P&IDs) I see the following vessels:

- Degassing Drum
- Feed Flash Drum
- Surge Drum
- Lube Oil Separator
- Gas Separator
- Reflux Accumulator
- Suction Scrubber
- Head Tank
- Storage Tank
- Day Tank
- Slug Catcher
- Deaerator

Although each of these is essentially a simple vessel or tank without any special internal structure, each serves a different purpose. Once it is clear what the purpose of a piece of equipment is, and how it functions, it will also be clear how to control and protect it. Different purposes require different controls.

SURGE TANKS. The most common function of a vessel or tank is to match two flows that are not identical in time but are expected to average out over the long run. Take a feed surge drum, for example. Flow into the unit is more or less steady but is subject to interruption. The flow to the processing unit should be as constant as possible, avoiding sudden change. Nevertheless, it, too, may be subject to interruption due to downstream conditions.

The purpose of the surge drum is to maintain sufficient inventory to feed the process and to maintain sufficient void capacity to continue receiving feed as it arrives. Clearly the tank must be large enough to accommodate any normal discrepancies between input and output over a reasonable period of time. Between the upper and lower bound, the exact value of the level does not matter.

Two separate control parameters are implied: Level and flow. Level control is no problem. Greg Shinskey[1] refers to "The easy element -- capacity". A high-gain level controller connected to a valve at either the inlet or the outlet will maintain the level very accurately at its setpoint. The only problem with this approach is that it absolutely defeats the purpose of the vessel. The same effect would be achieved by blocking in the vessel and bypassing the inlet directly to the outlet.

To control flow alone is also quite simple. A flow controller at the outlet, properly tuned, will maintain a steady flow to the process. Unfortunately, there is nothing to make this flow equal to inflow. It will not even equal the average inflow unless there is something to make it do so.

What we need is an instrument that measures the accumulated error between inflow and outflow. The tank itself is that instrument!

$$\text{Level} = \text{Starting Level} + \int (\text{Inflow} - \text{Outflow})\, dt \,/\, \text{Tank Area}$$

(To a process controls engineer, every piece of equipment is just a big, non-tuneable instrument!) The level transmitter only transmits the process value to the control system. If we now cascade the output of the level controller to the flow controller, we have a system that has one process variable: Accumulated flow imbalance. It has only one point of control: Outflow to the process.

To start this simple process:

- Fill the tank about half full.
- Give the level controller the current level as its set point. (PV tracking does this automatically.)
- Switch the flow controller to automatic with an estimated average flow as its setpoint.
- Switch the flow controller to cascade.
- Switch the level controller to automatic.

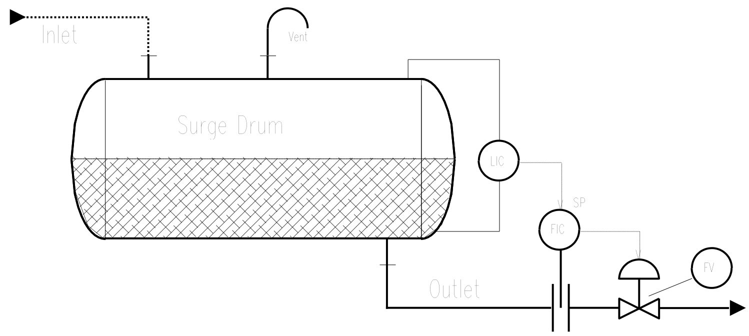


Fig. 6-1. Surge Drum Control

The control system will keep the flow "constant" but that constant varies in response to the imbalance between outflow and inflow. It is not important that the initial estimate of average flow be exact. A low guess will result in the tank level rising a little. A new, higher, estimate will result and the outflow will be adjusted accordingly. In the long term the average flow out is not an independent variable at all. It **will** be exactly equal to the average flow in. This can be accomplished at any arbitrary tank level. The level setpoint is based on the operator's estimate of the nature of the flow interruptions and whether the most probable upset will require additional flow or void capacity.

Should a pump be necessary to transfer the liquid from the vessel to its destination it should be placed between the vessel and the flow measurement. Further information on the control of pumps is found in *Controlling Centrifugal Pumps*[2]. This article also includes a section titled "On / off Control" for less critical level applications.

There is a long discussion on the special requirements for level control of steam heat exchangers and condensate receivers in *Controlling Steam Heaters*[3].

Surge drums are sometimes used for gas. The abrupt flow variations of a Pressure Swing Absorption (PSA) unit, for example, often need to be smoothed out before the tail gas can be introduced into a down-stream process. In these cases, pressure takes the role that level has in a liquid process. That is, a pressure / flow cascade is the appropriate solution.

TUNING SURGE TANK CONTROLLERS. Since the exact level of a surge drum is not important, the controller can be tuned very loosely allowing the level to rise and fall in response to any short term imbalances. This exactly serves the purpose of the surge tank; tight tuning defeats it. There is a non-linear control algorithm which specializes in the type of loose control required by surge tanks. One common name is the "gain on error squared" controller. Figure 6-2 shows its characteristic. The controller responds to small errors with a small gain; it responds to large errors with a large gain. This means that in the vicinity of the setpoint, the controller allows the level to drift freely and the flow to remain almost constant. With luck, the level will average out again before the deviation from setpoint is too great. If the level changes far from the setpoint so that the danger of running out

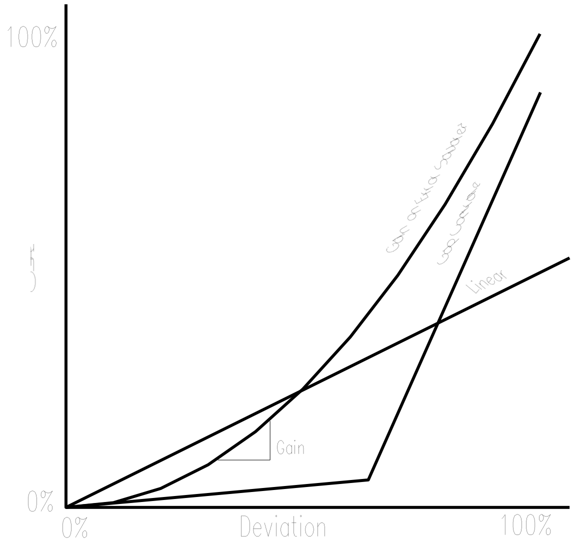


Fig. 6-2. Gain vs. Deviation for Three Types of Controllers

of capacity exists, the controller responds with a strong signal and rapidly brings the level back to near setpoint.

Another form of non-linear controller is also available: The notch or gap controller. This algorithm has the gain divided into three segments by two break points. The middle segment, on either side of the setpoint, has a low gain to avoid excessive action while the outer segments have a higher gain for a rapid return. It has the advantage of allowing the user to set the breakpoints and gains below the setpoint differently from those above. Its disadvantage is that it has four tuning constants instead of only the one found in the gain-on-error-squared controller. Some gap controllers have a zero gain in the centre segment. This is totally useless as the controller will never bring the level back to the setpoint. (No gain, no action.) Instead it will tend to use either the upper or lower breakpoint as its effective setpoint and return the level with a high gain. It should be noted that for a controller using a velocity algorithm an abrupt change in gain does **not** imply an abrupt change in valve position, only a change in the rate of movement. This function is more difficult to implement with controllers using the position algorithm as the controller has to be re-initialized with every gain change.

A simple proportional mode controller is sufficient for many surge drum applications. A slow integral may be used to bring the level back to the setpoint during a prolonged change in flow rate, but it should be turned off if cycling results. Do not use the derivative mode! Besides amplifying noise, derivative provides tight control by cancelling out the integrating capacity of the tank and thus defeating its purpose. A tuning rule I have heard of, but have not tested myself is

$$K = \Delta F/F * \Delta L/L$$

Where K = controller proportional gain

$\Delta F/F$ = the proportion of flow variations in the uncontrolled flow

$\Delta L/L$ = the proportion of level available for surge. This is the distance between the level setpoint and the nearest alarm.

This formula attempts to put the loosest level control consistent with keeping the level away from the alarms. There is a catch, however: It is necessary to predict the amount of flow variation to be expected in the future. Of course it is also necessary to do this to a certain extent when the vessel is sized.

SUCTION SCRUBBERS. A compressor suction scrubber is an example of a vessel whose purpose is to separate, collect, and dump relatively small quantities of liquid from a gas stream. The following conditions generally apply:

- Precise level control is of no value.
- The liquid flows to some form of drain.
- Smoothness of liquid flow is of no value.
- The average liquid flow is quite small.
- The pressure differential across the valve is high.
- Relatively large slugs of liquid occur occasionally.

The last three conditions would result in a valve that is usually operating near its seat with a high ΔP. It would experience severe erosion resulting in a short, unhappy life. The solution is to control the valve in on / off or "snap acting" mode. There are several ways to accomplish this. The simplest is to tune the controller to a very high gain. This would cause the valve to spend almost all its time in the full open or closed position. Unfortunately the high-gain controller would also try to maintain accurate level control by rapidly switching the valve between these extreme positions. Any saving in seat erosion would be cancelled by a high rate of stem and packing wear. The same response can be achieved by using a simple level switch connected to the control valve via a solenoid. (Pneumatic level switches tubed directly to the valve actuator diaphragm are also available.) A level switch can be viewed as a controller with an extremely narrow proportional band (0%!) and consequently an extremely high gain (100% / 0% = ∞ !).

Selecting a switch with a broad deadband results in a great improvement. The valve now remains fully open until a significant reduction in level is achieved. It then remains fully closed until the level substantially rises. With this arrangement it is possible for the valve to have both long life and peak capacity. Experience indicates that transmitters are more reliable instruments than switches and also demand less maintenance[4]. If a transmitter is used, the deadband function is accomplished through logic in the control system. This would have the added advantage of allowing the operator access to the high and low setpoints. In some ways the suction scrubber acts as the exact opposite of a surge drum -- it collects slow dribbles of flow and releases them as intermittent surges.

Sometimes there is a third option -- specialized liquid dump valves. These behave somewhat like steam traps in their ability to pop open in the presence of liquid and snap shut in the presence of vapour. Since they are not general purpose instruments, it is best to use them only when there is an opportunity to test their performance; the vendor should be consulted. These devices might be very cost effective in packaged equipment such as on the discharge receiver of an instrument air compressor.

STEAM DRUMS. The purpose of a boiler steam drum is to provide space in which the water and steam may disengage. Since the drum serves at high pressures and temperatures, perhaps up to 3600 psi and 1000ºF (25 MPa and 540ºC), it is expensive to manufacture and there is considerable economic incentive to keep it as small as possible. The techniques of boiler feed water (BFW) control can be applied whenever extremely tight level control is a requirement.

The level of the feedwater in the steam drum must be kept above the bottom of the drum or a catastrophic explosion may result. It must also be kept below the steam outlet or liquid water will be carried over. Water droplets will damage superheater tubes, turbine blades, and other equipment. The diameter of the steam drum, and hence its cost, is determined largely by the ability of the control system to keep the water level within bounds.

Thus level control of a steam drum has exactly the opposite purpose of that of a surge drum: The water level must be kept within an extremely narrow band and tight control is of essence. It is a simple matter to maintain tight level control... use both the proportional and integral modes and turn up the gain! Figure 6-3, Single-Element BFW Control, shows this very simple arrangement. As always, there are complications. Firstly, high gain means extremely rapid swings in flow rate. The BFW pumps suffer under that type of abuse. There is a second problem, peculiar to boilers, called "swell". Swell is the phenomenon in which a rise in steam demand causes a drop in pressure. This in turn results in a rapid boilup within the tubes which causes the water level to rise. Paradoxically, an increased steam removal rate causes a rise in level due to the swelling of the steam bubbles. The level controller responds by reducing BFW flow at the very moment it is needed most. The swelling water soon collapses as the steam rises to the surface. Now the controller reverses its response and adds a large amount of essentially cold BFW into the system. This causes the water temperature to fall. The cooler water shrinks, lowering the level further. The use of single-element control is not very highly recommended for boilers!

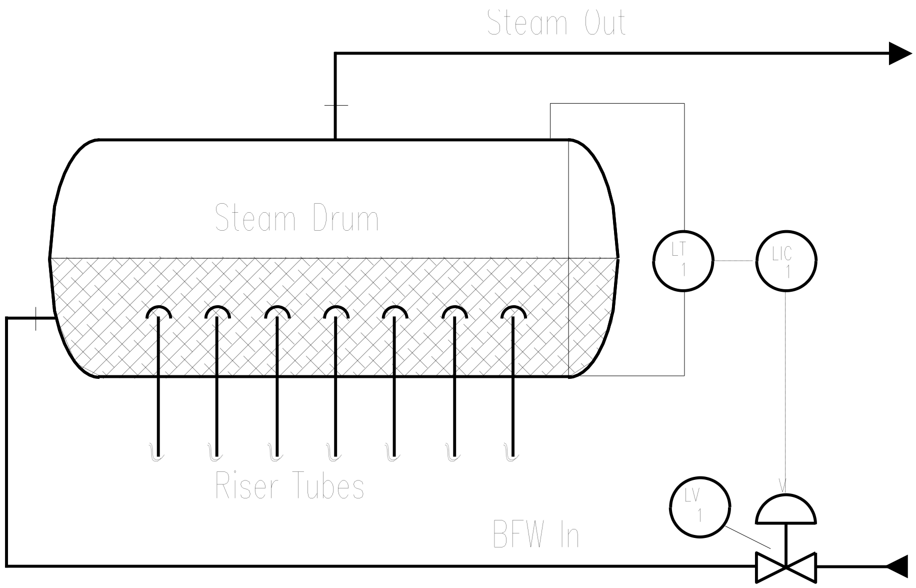


Fig. 6-3. Single-Element BFW Control

The disturbance to the level is caused by a change in steam withdrawal rate. Since this is a measurable quantity, feed forward can be applied to the level controller output. Figure 6-4 shows how this is accomplished. The compensated steam flow is added to the output of the level controller. Thus a rise in steam withdrawal and the swelling of the water is accompanied simultaneously with a surge of cold BFW. Ideally the two cancel out exactly and the controller sees no change in level at all. They will not cancel out exactly for two reasons: Firstly, there is no reason why they should. One effect or the other will predominate. They won't even be simultaneous. Secondly, the BFW

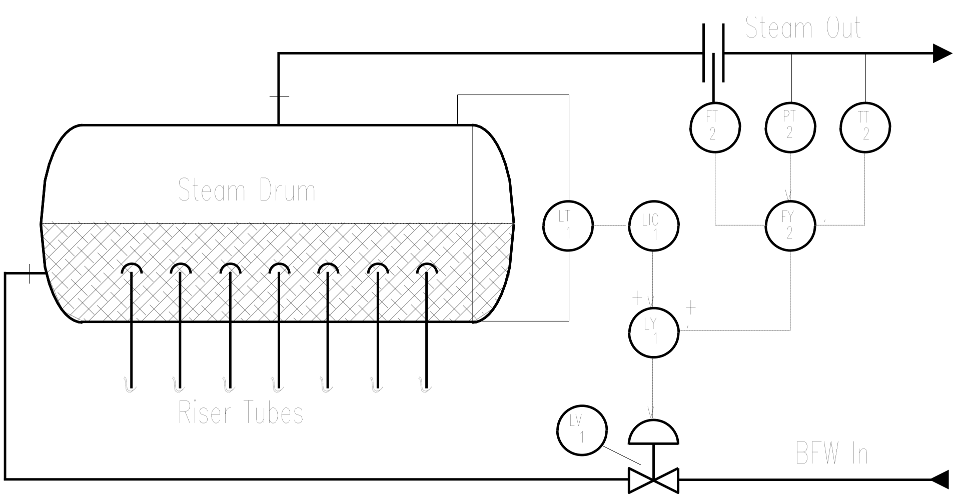


Fig. 6-4. Two-Element BFW Control

flow can only equal the steam withdrawal if the range of the valve is exactly equal to the range of the compensated steam flow. Since these two functions must be exactly equal over the entire operating range, it means that the valve must be perfectly linear and that its ΔP is absolutely constant. Not likely! So the level controller still has some work to do to keep the accumulated error at zero.

The rather farcical suggestion in the previous section, piping the inlet to outlet and bypassing the vessel, suggests a solution to the valve linearity problem: Use the measurement of the steam leaving the boiler as the setpoint to a BFW flow control loop. The level should remain constant once the shrinking and swelling have reached the new equilibrium. This simplistic solution overlooks a basic principle of process control: No **two measured quantities are ever identical**. In other words, the two flows will never be the same and the level will rise or fall at a rate proportional to the difference. Since level is a measure of the accumulated difference, a level controller is used to correct the BFW flow. What I have just described is the classic three-element boiler level control arrangement as shown in Figure 6-5.

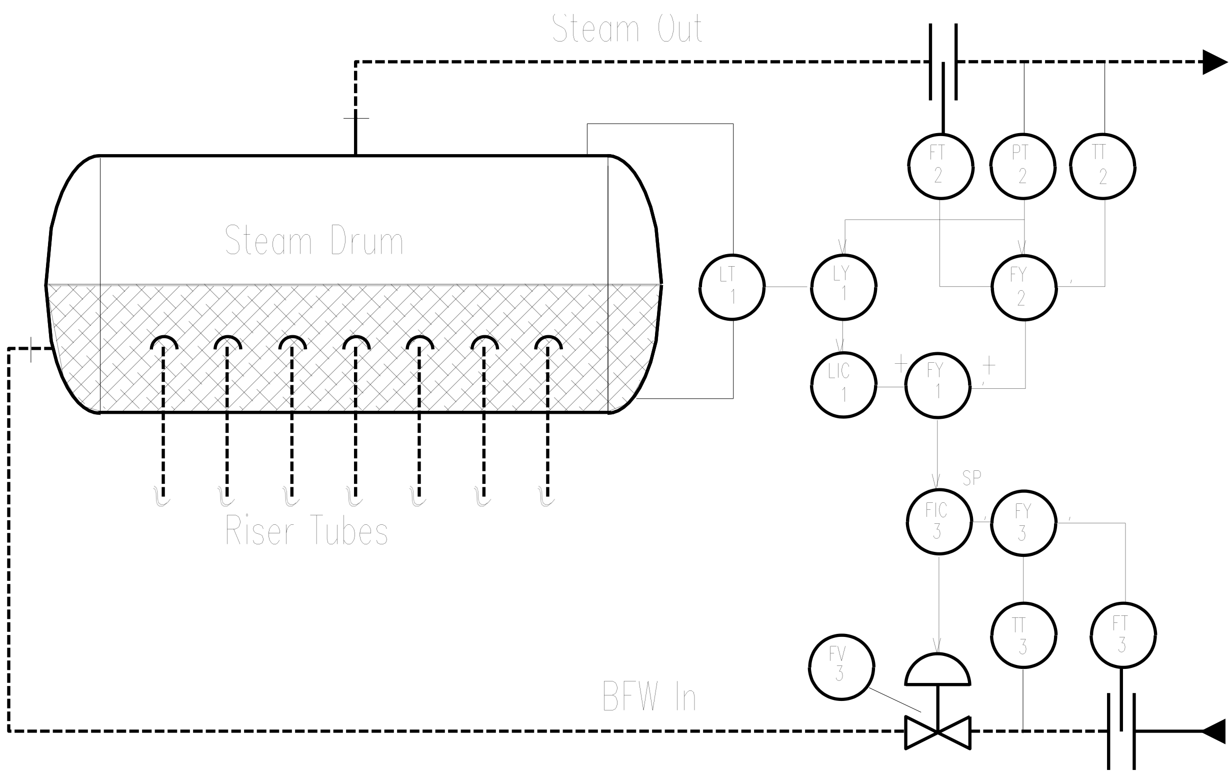


Fig. 6-5. Three-Element BFW Control

The diagram also illustrates a few other features. Compensation has been applied to account for the effect of pressure on the steam density and its effect on the level transmitter. BFW flow is sometimes temperature compensated since it is most probably preheated and its temperature may vary. For a temperature change from 0ºC to 300ºC (32ºF to 572ºF) the specific gravity changes from 1.000 to 0.712 and a measurement error of 15% will result.

This detailed exposition of boiler level control is presented only to provide an example of how extremely tight level control can be accomplished when necessary. Boiler control is a rather broad subject and many articles and textbooks have already been published concerning it.

CONTROLLING LIQUID INTERFACES. It is generally assumed that level control refers to the control of a gas / liquid or vapour / liquid interface. It ain't necessarily so. An interface can occur between any two immiscible fluids. Since all gases are miscible with each other in all proportions, interface level control is always taken to mean the interface between two liquids such as oil and water.

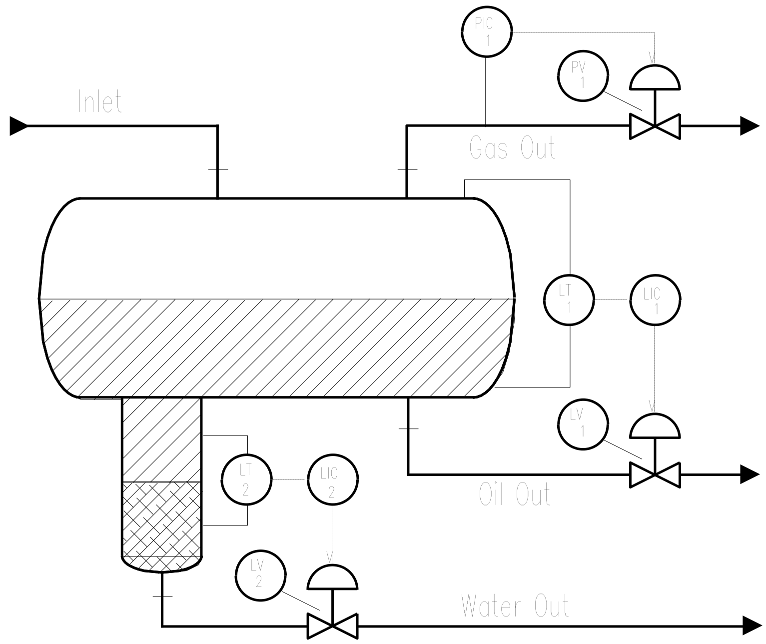


Fig. 6-6. A Crude Oil Separator

Figure 6-6 shows an example of a boot on a crude oil separator. This vessel serves three purposes: It is a gas / oil separator, a feed surge drum and a water separator. A real vessel in this service probably contains inlet baffles and demister pads. Each of the three phases must be individually controlled. But it is possible for the gas phase to discharge to an externally pressure controlled system or even to atmosphere. (Possible yes, acceptable no.) The key to understanding the function of any separator is to think in terms of a constant inventory of each component. To repeat: each component must be controlled individually. The amount of gas flowing in must be balanced by gas flowing out. Changing pressure is a measure of the gas imbalance, therefore pressure control is the appropriate way of controlling the gas outlet. Similarly, oil level is an indication of oil imbalance and water level indicates water imbalance. None of the three streams may be controlled on flow, although a level / flow cascade is often used to smooth out flow variations to the downstream equipment. Pressure / flow cascade is unlikely to be used unless the volume of the vessel is large enough to serve as a gas surge drum. Level / flow cascade on the water is unlikely since the water probably drains to a collection system that itself serves as a surge drum to a number of separators.

Sometimes the ratio of water to oil is too great for a boot separator. In such cases a weir may be used to divide the vessel as shown in Figure 6-7.

Certain precautions must be taken to make sure that the level transmitter actually gives a true indication of the interface. There are basically two varieties of level indicating devices: The first measures the distance of an actual interface from some fixed point. Ultrasonic and radar devices belong to this group. These would be ideal for the purpose except that they are often not suited for installation in pressurized vessels. Furthermore they have difficulty "seeing" anything other than the very top interface. Even surface foam and condensation on the instrument “window” can confuse them.

The second, and more traditional, variety integrates some particular property, such as density or dielectric constant, over a span. Displacers, differ-ential pressure transmit-ters, bubbler tubes, nuclear densitometers, capacit-ance probes, and even gauge glasses all belong to this variety. The key to successful measurement is that the level sensing device must sense only the two fluids bounding the interface. For a gauge glass this means that the lower tap must be in the lower of the two fluids and the upper tap must be in the fluid immediately above it. There may be **NO** intervening phases.

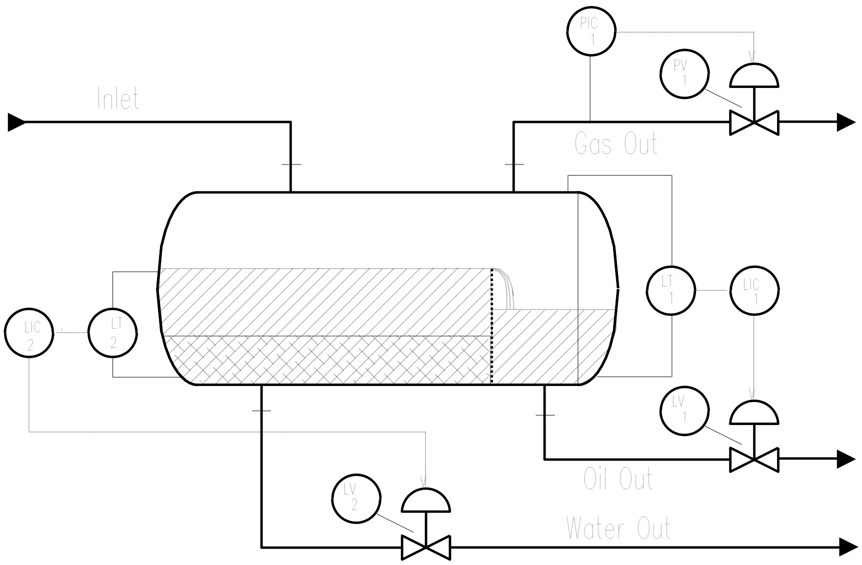


Fig. 6-7. A Three-Phase Separator

Figure 6-8 shows what happens when a gauge glass is connected to a vessel containing a vapour and two liquid phases. Assume that equal amounts of a liquid with Sg = 1.0, e.g. water, and a liquid with Sg = 0.5, perhaps propane, gradually flow into the vessel. Assume further that the span of the gauge glass is four feet, beginning one foot from the bottom of the vessel.

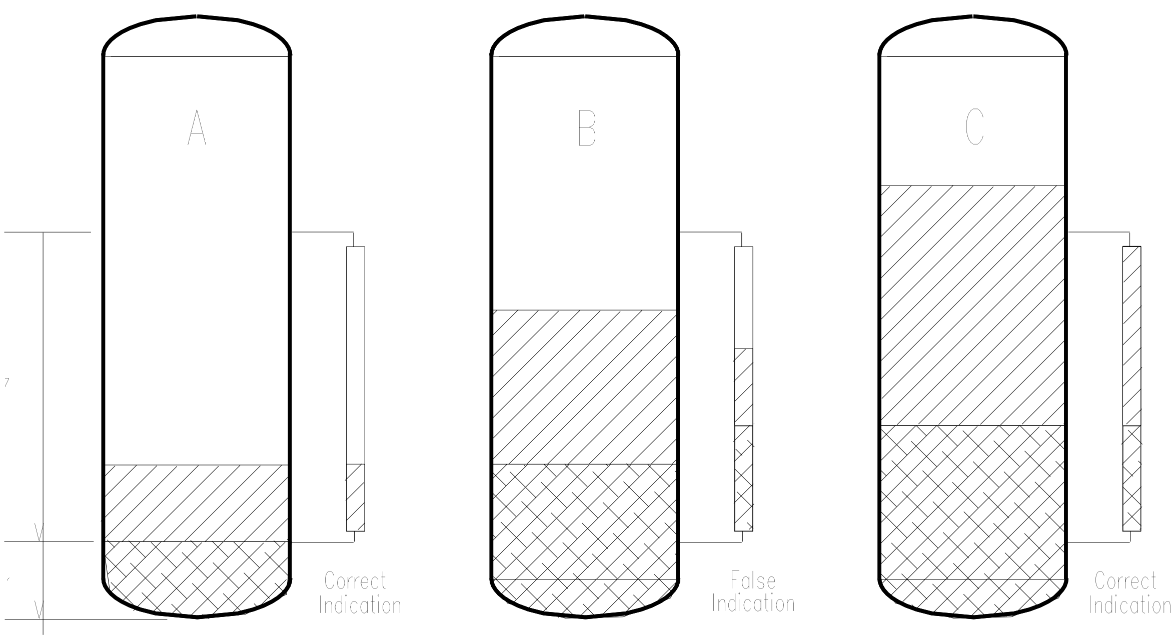


Fig. 6-8. Gauge Glasses and Interfaces

As the level of the propane rises, it flows into the glass. As both liquids rise further, water *begins* to enter the bottom of the glass. This is the state shown in vessel A. Up to this point, the glass shows a true indication of the level of propane in the vessel. Once water enters the glass, the propane is cut off. A constant plug, one foot thick, floats on top of the water. Its level no longer bears any obvious relationship to the actual level in the vessel. This is state shown in vessel B. The only relationship between the vessel and the glass is that the hydrostatic pressure is the same for both at the point where the glass taps into the vessel. A gauge glass is really nothing more than a manometer.

Once the level of the propane rises above the upper tap, it flows into the glass and the two interface levels adjust to the same elevation, as shown in vessel C. The gauge will continue to read correctly as long as its lower tap is in the water and the upper tap is in propane. If either fluid is withdrawn so that the upper tap is in the vapour space, the glass will once again read falsely.

This same analysis applies to any type of level indication based on density. Remember that a ΔP transmitter only gives a single reading, i.e. differential pressure. Therefore only a single quantity can be inferred. If the instrument is affected by only two fluids, it can yield the correct interface level between the two. If there are more than two distinct phases within the span of the two taps, it will give a reading based on the average densities of all the fluids within its span.

Capacitance or nuclear level transmitters will give similar results in multiphase situations, based on the average dielectric or nuclear absorption constants, respectively.

So... how can the process controls engineer be assured that the level readings are meaningful if even a gauge glass can't be trusted? Plan "A": Make the entire vessel out of glass. This isn't usually practical so we must fall back upon Plan "B": Every section of a gauge glass must have separate taps into the vessel so that each pair of taps has no "hidden" phase floating in between. Either that, or accept the fact that until the interface reaches its "normal" range, gauge glasses and transmitters will read falsely.

SLUG CATCHERS. Slug catchers are a special instance of three-phase separators frequently found in oilfield service. In addition to the usual separation functions, they are required to serve as surge tanks that can smooth out intermittent flow and also handle occasional very large surges in inlet flow. This is done by having two controllers connected to the oil-side transmitter. The oil overflow controller has its setpoint slightly below the top of the weir. In this manner, any surges can be accommodated by the large volume above the weir. This is in fact a non-linear, adaptive gain transmitter since transmitter gain = Δ output / Δ volume.

The inlet controller responds to the same level but has its setpoint just below the top of the vessel. It takes action only when the level rises to its set-point. This would happen if an unusually large slug of liquid arrived or if an upset in the downstream process caused the system to back up into the slug catcher. The facility would then be operating under "capacity control". Facilities lacking the capacity control feature are likely to experience a high level shutdown precisely when they are attempting to operate at maximum throughput. Not a very desirable occurrence.

It is common for level controllers to be tuned using both the proportional and integral modes. Since the inlet controller is normally functioning with the level well below its setpoint, reset windup will occur. This is a phenomena in which the controller attempts to raise the level to the setpoint by forcing an ever higher signal to the valve. This does not work, of course, since the valve is already wide open. If a sudden surge arrives that abruptly raises the level to the setpoint and beyond, the

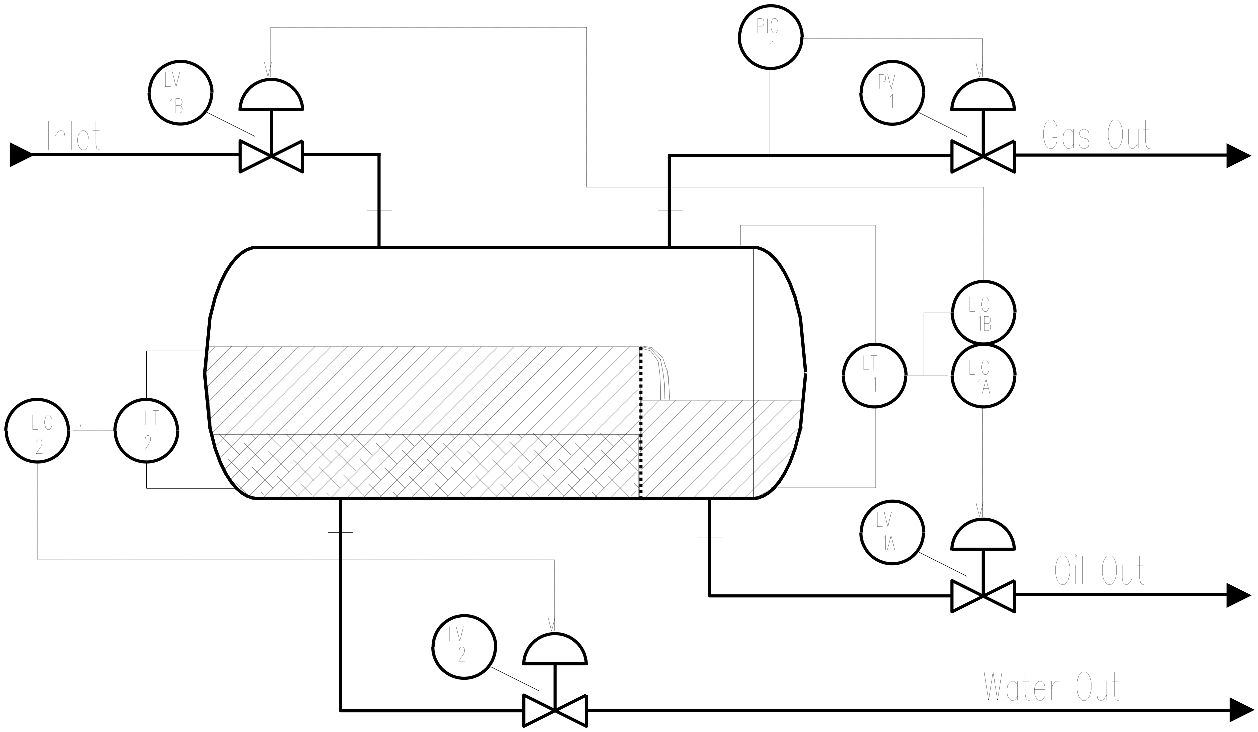


Fig. 6-9. A Three-Phase Separator / Slug Catcher

controller will be slow to close the valve since it has "wound up" in the opposite direction. Some form of anti-reset windup is required to prevent an unwarranted high level shutdown under these circumstances. It is probably a bad idea to use an equal percent valve in this application since it, also, is likely to respond slowly to a sudden demand.

It is possible to control the outlet and inlet valves with a single, split-range controller. This method accomplishes the required function of preventing high level shutdowns but has a serious disadvantage. If the setpoint of the combined inlet / outlet controller is set below the top of the weir, it will not take full advantage of the surge capacity of the vessel since the inlet will begin to close well before the top of the vessel is reached. If the setpoint is above the weir, it defeats its purpose by allowing mixed feed to flow directly to the oil outlet before it has time to separate. Thus a split-range controller will sacrifice either separation quality or surge capacity.

PRESSURIZATION SYSTEMS. A tank, vessel, or drum may require a pressurization system for any of a variety of reasons:

- The surface of the liquid in a reflux drum consists of a liquid at equilibrium with its vapour. There may not be sufficient gravity head to provide the net-positive-suction-head required to operate the reflux pump without cavitation. Raising the vessel high in the air above the pump is one way of providing this. Unfortunately the condenser providing the liquid, drains by gravity so it must be raised even higher. The entire arrangement can become extremely expensive. An-other method is to use a canned pump which is sunk deep into the ground. This can also get pricey. A blanket gas pad may be a relatively inexpensive way of providing the necessary pressure.

- The liquid in a storage tank is subject to oxidation, e.g. the surge tank of a glycol-based heat exchange system. A blanket of fuel gas will prevent the tank from breathing air as it cycles from empty to full and back again.

- The liquid in a storage tank forms an explosive mixture with air. (A rather extreme form of oxidation!) Continuous gas pressurization may be required to prevent this.

- The storage tank vents to a flare or other vapour collection system. A gas supply must be provided to make up any volume withdrawn when the withdrawal rate exceeds the fill rate. In other words, the pressurization system serves as a vacuum breaker.

A simple way of providing pressurization is to have a regulator connected to a source of pad gas and a second, back pressure regulator, connected to the vent. Care must be taken to set the back pressure regulator setpoint slightly higher than that of the inlet regulator. If there is no gap between the two settings, the pad gas will blow straight through to the vent. Remember that setting them to the "same" pressure is meaningless.

Often it is necessary to install a complete control loop including a pressure transmitter, a controller, and two valves. This has the advantage of allowing the panel operator to monitor and adjust a single setpoint. It also allows over- or under-pressure alarms to be easily provided. Figure 6-10 shows how the complete pressure control loop is arranged. For the most part it is pretty simple but there are two things to watch for: Firstly, there must be a gap between the action of the two valves. That is the reason for the split range values not meeting at 50%. Secondly, the failure mode of the valves must be taken into account. Since the two valves have the opposite effect, they must have opposite failure modes if they are to be operated by the same control signal. A DCS allows the output of the controller to drive two separate output modules, each characterized in its own way. This means that it is possible for the first 45% of the controller output to produce a 100 to 0% signal for the fill valve, and the last 45% of the output to produce a 0 to 100% signal for the vent valve. In this way both failure modes are accommodated and overlap of valve openings is impossible. The gap in the middle does not cause a problem for the controller. Integral windup will move the output quickly through the gap whenever there is a deviation from the setpoint. The reader should note that the split range control described in this paragraph is not at all the same as that described in the section on Slug Catchers.

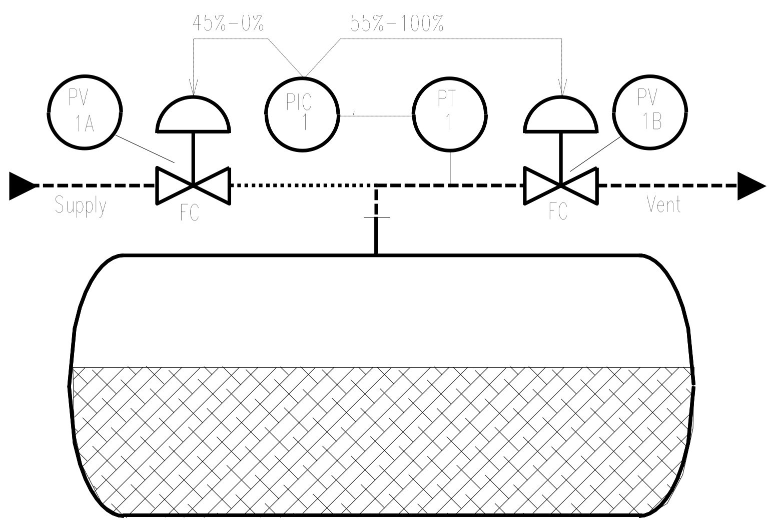


Fig. 6-10. A Gas Blanket Pressure Control Loop

It is possible to achieve the same effect by using a specialized, three-way valve that provides a gap in the middle. Most three-way valves are designed to have full overlap as they are intended for use in diverging / converging service. (If anyone knows of a centre-gap, non-overlapping valve, let me know.)

A number of vendors sell specialized gas blanketing systems capable of self-contained action. They consist of regulators with the very large diaphragms required to drive valves with pressures as low as 0.5" WC (125 Pa, 0.3 oz/in^2). Such systems are especially useful now that ever more stringent regulations concerning the emission of volatile organic compounds (VOCs) are being enforced. Figure 6-11 shows one typical arrangement.

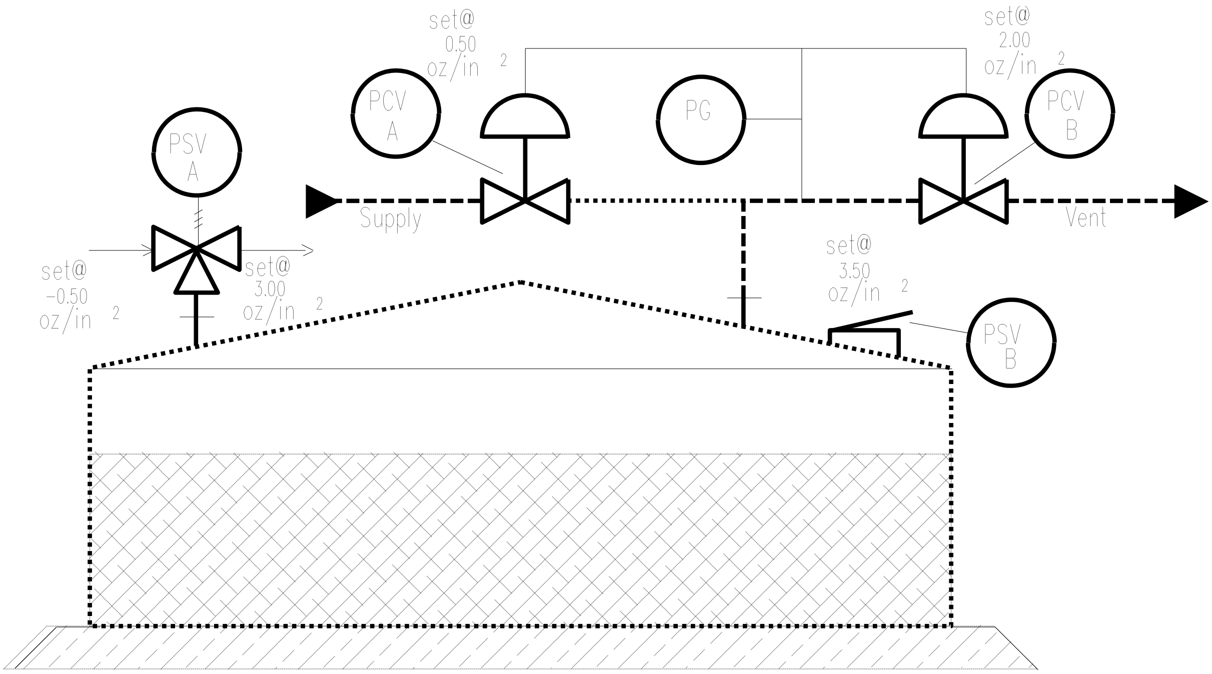


Fig. 6-11. A Typical Tank Blanketing System

Many factors enter into the correct specification of the setpoints and sizes for the various regulator and relief valves. These include:

- The maximum allowable pressure of the tank.
- The maximum allowable vacuum of the tank.
- The vapour pressure of the stored liquid.
- Inbreathing rate dependent on pump-out rate.
- Outbreathing rate dependent on pump-in rate.
- Vapour thermal expansion and contraction rate.
- Tank surface area and insulation.

Table 6-1 provides setpoints applied in a specific case. It must be remembered that actual values differ widely. *API 2000, Venting Atmospheric and Low-Pressure Storage Tanks*[5] and tank vendors provide much information, however it may be advisable to consult a specialist in the field.

	oz/in^2	"WC	kPa
Maximum Allowable Pressure	4.0	6.9	1.7
Manway Setting	3.5	6.1	1.5

Relief Valve Pressure Setting	3.0	5.2	1.3
Vent Regulator Setting	2.0	3.5	0.9
Fill Regulator Setting	0.5	0.9	0.2
Relief Valve Vacuum Setting	-.05	-.09	-.02
Maximum Allowable Vacuum	-1.0	-1.7	-0.4

Table 6-1. Typical Tank Blanket Pressure Settings

A brief sermon on tagging: According to *ISA 5.1, Instrumentation Symbols and Identification*[6], all forms of relief valves including pressure, vacuum, spring- or weight-loaded, with or without a pilot are tagged "PSV". Common abbreviations such as "PVSV", "PVRV" or "PRV" have absolutely no official status and therefore are not acceptable as tags on P&IDs.

LEVEL MEASUREMENT. Level measurement is deceptively easy, yet it seems that more time is spent specifying level instruments than any other. The reason is that the correct installation of level instrument is an interdisciplinary effort involving Process, who set the basic functional requirements; Mechanical, who have various constraints such distance of taps from seams; Piping, who have accessibility and orientation requirements; and Instrumentation, who must select from a finite catalogue of available instruments.

Actually this task has become considerably easier in recent years due to the increased use of ΔP transmitters and other instruments such as ultrasonic and radar which do not have a predetermined span. There is no longer any significant penalty in either cost or accuracy if the instrument is specified to cover a broad span. For horizontal vessels the top connection should be vertical at the top of the vessel. The bottom connection should be horizontal a few inches from the bottom. The gap is necessary to prevent the accumulation of sediment at the instrument. These connections no longer need to be in the same vertical plane nor do they require the same orientation.

Vertical vessels may still require a bit more attention. While a top-to-bottom span would be ideal, there may be trays, packing, or other internals that would cause a differential pressure in response to flow. It is also necessary for the level connections to remain clear of welding seams. This requirement may cause problems if alarms or other setpoints need to be near the bottom or top of the vessel.

The design process begins with the basic information on a P&ID in a form similar to that shown in Figure 6-12. A brief outline of the vessel including the bridle, if any, holding a gauge glass and a transmitter are shown. The desired values for the level alarms and the setpoint of the controller are also shown.

The Control Systems engineer must first decide if a ΔP transmitter is the appropriate choice. Let us assume it is. He/she must then try to find an appropriate span for the transmitter. A good rule of thumb is that alarms should not be set any lower than 10% or higher than 90% of the transmitter span. (Shutdown trip settings should not be closer than 5% from either end of span.) Since the two alarms are 42" - 6" = 36" apart, the span should be 36" x 1.25 = 45" thus allowing the alarms to be at 10% and 90% of span. This seems fine, but there is a problem. The first thing to determine is whether the vessel measurements are from the tangent line, from the seams or from some other reference point. In this particular case the vessel title block indicates that measurements are tan-to-tan. Since seams are generally 2" inside the tan lines, the lower tap of the transmitter is ½" above the seam. That is not acceptable. Mechanical considerations often dictate that nozzles may not be welded within 6" of a seam. This means that the lowest transmitter tap cannot be lower than 8" above the bottom tan line. The highest tap cannot be higher then 8" below the top tan line. This implies that the maximum transmitter span on a 48" T/T (tan-to-tan) vessel is 32". Alarms at 10% and 90% must be placed at 11.2" and 36.8". At this point, the Instrument Engineer becomes a broker between Process and Mechanical to help them find a compromise. Alarms at 11" and 37" are agreed upon. Don't forget to transfer this new information back to the P&ID!

It is a great convenience to the maintenance staff if the span of the transmitter is exactly equal to the span of the gauge glass. This is not always possible with displacers since both the gauge glasses and the transmitters come in fixed spans. However, it can easily be done for ∆P transmitters. The centre line of the sensing taps must be located at the tops and bottoms of the visible glass. Unfortunately, the associated valves bring the centre line of the gauge glass tap 4½" below the bottom of the glass. Fortunately, in extreme cases, it is possible to place the lower gauge glass tap *below* the lower bridle tap. There can be no meaningful readings below the lower bridle tap of course, since the bottom of the glass can never drain back into the vessel. As long as the glass itself does not go below the lower tap, it's OK. Since gauge glasses come in fixed lengths, it may not be possible for the upper tap of the transmitter to match the top of the visible glass. Remember that transmitter calibration will be more difficult if the upper tap does not fall within the range of the visible glass.

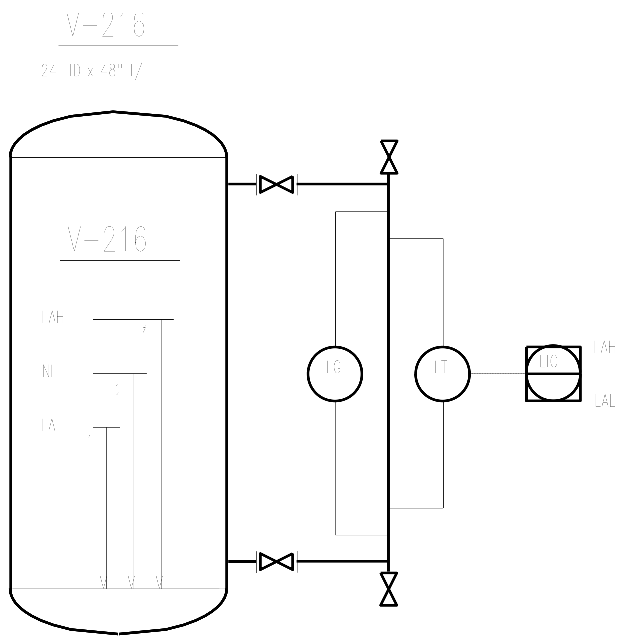


Fig. 6-12. Level Setpoints

Occasionally it is necessary to connect either the top or the bottom taps to interconnecting pipe instead of to the vessel itself. If the taps are attached to inlet or outlet lines, the level signal will be affected by flow rate. This effect can be seen in coffee percolators: The level in the gauge glass bobs up and down as coffee is drawn into a cup. It definitely could not be used for level control.

SEALS. Diaphragm seals have become a very popular accessory to ∆P-based level transmitters. A very thin metal diaphragm isolates the transmitter from the process. The space between the diaphragm and the sensor itself is filled with a fluid such as silicone. The pressure changes are communicated through the diaphragm to the transmitter via an armoured capillary tube. The volume change between minimum and maximum pressure is extremely small in a modern transmitter; the amount of flex in the diaphragm is correspondingly small. The net effect of this is that the introduction of a diaphragm into a measurement system introduces an error of only several centimetres or less. This is seldom significant in level applications. A second effect is that only an extremely small amount of liquid movement actually occurs in the capillary tubes. This, together with modern low temperature fill fluids, means that the instrument response does not slow down too much on cold days. Figure 6-13 is an example from one vendor's catalogue.

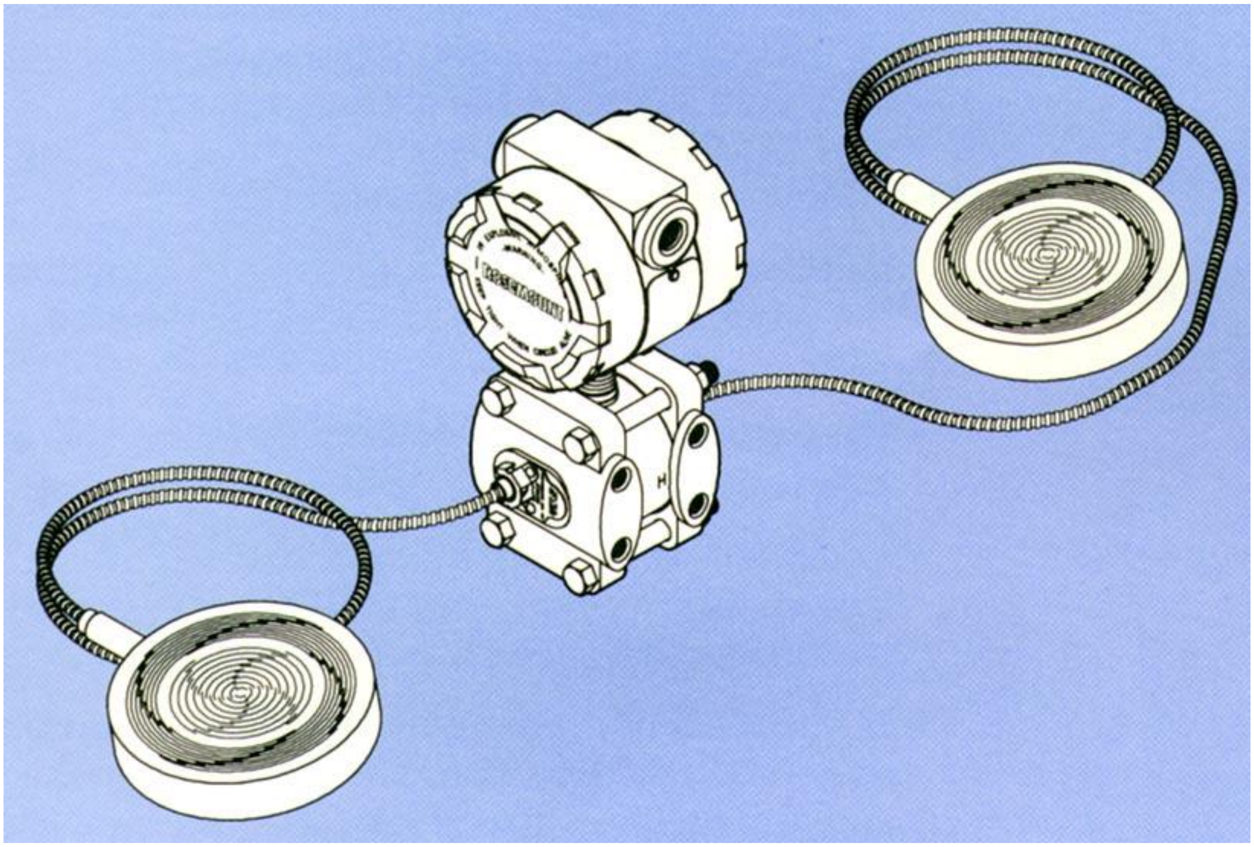


Fig. 6-13. A Typical Differential Pressure Transmitter with Diaphragm Seals

Seals should be considered whenever one or more of the following conditions apply:

- Dirty service - Whenever the process fluid is liable to plug the impulse lines, a diaphragm seal may be installed. It should be isolated from the vessel by a full-ported valve, NPS 2 or 3. Note that two NPS ½ taps are provided on the diaphragm housing for calibration and flushing connections. Seals are especially useful in sanitary service where all hardware in contact with process fluid must frequently be thoroughly washed. It is a good idea to use seals and capillaries of equal length on both the upper and the lower leg in order to maintain a balanced response to errors.

- Corrosive service -- Diaphragm seals made of corrosion resistant materials originated in corrosive service where they were referred to as 'chemical seals'. While the use of full-ported connections is not required in corrosive service, it is a good practice to maintain even if it might look strange to have 'such a big valve' for 'only' an instrument.

- Freeze protection -- Diaphragm seals may eliminate the problem of freezing impulse lines. However, in extremely cold weather it may still be necessary to heat trace the capillaries to prevent measurement response from being excessively slow. Self-limiting electrical heat trace is the only way to go! Any heat trace system involving a thermostat will introduce spikes into the measurement system as the heat is switched on and off.

- Uncertain phase -- This is the most frequent of all seal justifications. A warm vapour in equilibrium with its liquid will undergo condensation in the upper impulse line. Cold equilibrium liquid may experience boiling in the lower impulse line. Thus the measurement will slowly drift as the tube fills. Depending upon service and ambient temperatures, condensation and boiling may even alternate throughout the day. If this situation exists, the measurement becomes worthless. The traditional solution is to fill the upper line with a non-volatile, process compatible fluid. Depending on process and ambient conditions this might be water, glycol, oil or something else. The use of fill fluids introduces maintenance problems because any attempt to 'null' the transmitter by opening the equalization valve will drain the upper fluid into the process. It can only be replaced by climbing to the top of the vessel and filling the tube again. Bubbles are also a source of error. Seals provide a captive fill fluid that cannot be lost, does not form bubbles and cannot contaminate the process. (Did I say foolproof?)

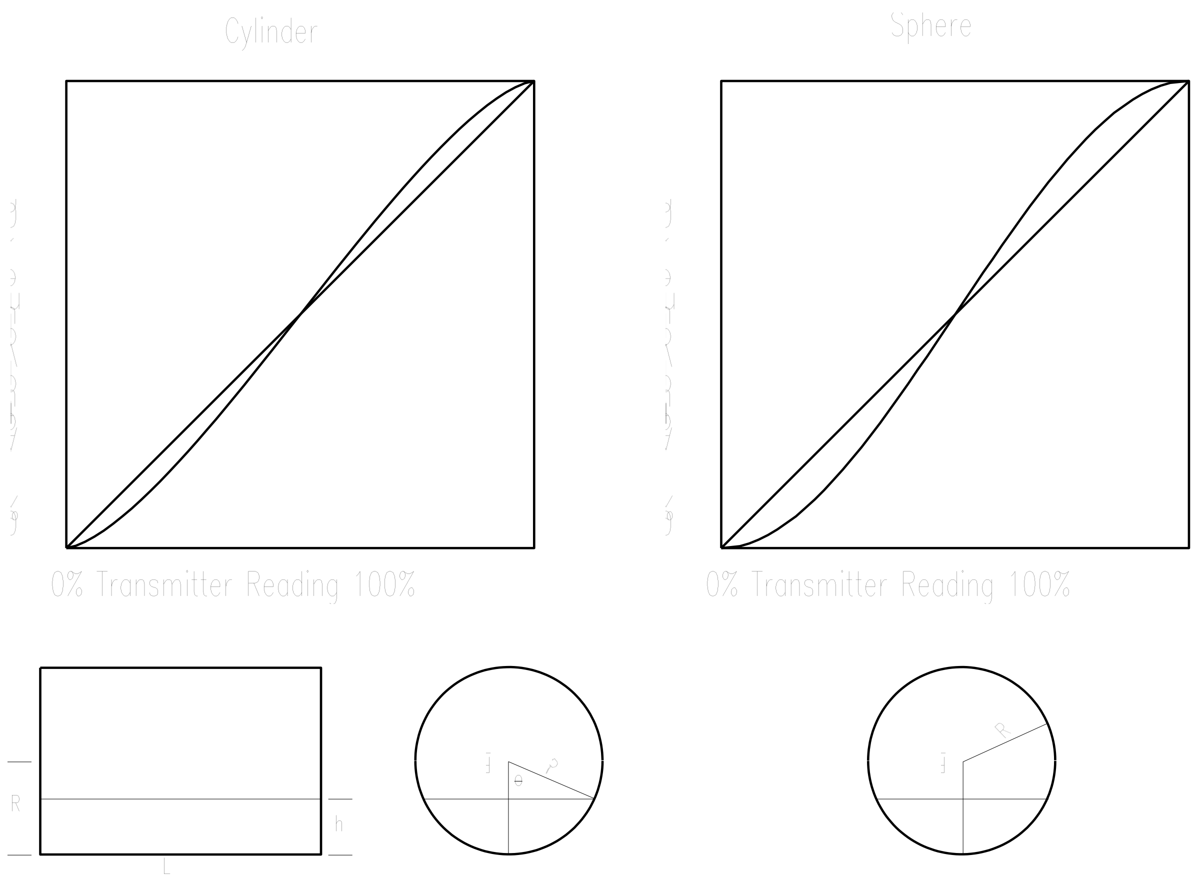
UNDERGROUND TANKS. A special requirement concerns underground (UG) tanks. Modern steel UG tanks have a double wall construction. Requirements are outlined in *CAN / ULC-S603, Standard for Steel Underground Tanks for Flammable and Combustible Liquids*[7]. The two walls of the tank are approximately an inch (2 cm) apart. A vacuum of 51 kPa (7.5 psi) is drawn on the interstitial space so that the two surfaces are, in many places, actually in contact with each other. A vacuum gauge is connected to the interstice. It must read at least 42 kPa (6.1 psi) of vacuum before the tank may installed. If the reading is ever less than 34 kPa (4.9 psi) the tank should be removed from service and steps taken to determine the cause of the leak. These values are summarized in Table 6-2, below. If a facility has many UG tanks, it may be desirable to connect the tanks to the central control system by means of vacuum transmitters. Low vacuum alarms can then be configured to alert the operators of any cases of leakage.

Interstitial Vacuum	psi	kPa
Required at manufacturing	7.5	51
Minimum acceptable for delivery	6.1	42
Minimum allowable in service	4.9	34

Table 6-2 Interstitial Vacuum Requirements for Underground Tanks

VOLUME MEASUREMENT. Most vessel and tank content measurements are made in the form of level. When true volume is required for such purposes as custody transfer, the tank volumes are calculated taking into account all details of their geometry as well as dimensional changes resulting from the pressure exerted by the density of the liquid. The results of these calculations and calibrations are tabulated by the manufacturer in a form known as “strapping tables”.

True volume measurement is seldom relevant for control purposes since setpoints for controllers and alarms are usually related to specific geometric features of the vessels. The level must often be kept below the vapour outlet or a weir. A frequent requirement is that a specific head be maintained to prevent pump cavitation. Sometimes the requirement is simply to maintain the level near midpoint in order to provide surge capacity. None of these applications benefit from true volume compensation. Figure 6-14, Volume vs. Height for Cylinders and Spheres, provides the correct mathematical relationship between level and volume for these two vessel styles if true volume measurement is actually required. It can be seen that between 10% and 90% little is gained by applying the rather complex calculations required for volume control.



Volume = $(R^2L/2)(2\theta - \sin 2\theta)$

Volume = $(\pi/3)h^2(3R - h)$

Where
- h = height of liquid in vessel
- R = radius of vessel
- L = length of cylinder
- θ is radians and $\cos\theta = (R-h)/R$

Note: The volume contained by elliptical vessel heads is ½ that of a sphere of equal radius.

Fig. 6-14. Volume vs. Height for Cylinders and Spheres

SAFETY. Vessels and tanks are probably the most hazardous pieces of equipment in any plant. Duguid's database[20] shows that 22% of all safety incidents are related to storage and blending. This may seem a little surprising until one considers that they store energy as well as material. For example:

- A vessel holding a compressed gas can cause a tremendous explosion if it ruptures. That is why "hydrotesting" with air or nitrogen is **far** more dangerous than with water.

- Storage tanks can hold a considerable amount of gravitational energy. The most notorious example of this energy being released is the infamous "Boston Molasses Disaster" which occurred January 15, 1919. A tank located at the top of a hill ruptured and released two million gallons of molasses down a narrow street. Twenty-one people were killed and 150 were injured.

- The contents of the tank or vessel can be flammable. While a line rupture external to the tank may be the cause of a fire, it is the reservoir of flammable fluid inside a tank that turns a minor fire into a major one. *API RP 750, Management of Process Hazards*, specifically addresses this point, however, it does not offer much in the way of solutions.

> **Torrent Kills Nine**
> The Associated Press
> MELILLA, Spain November 19, 1997
>
> A water tank burst and sent a river of water surging through a busy market area in a Spanish enclave on the northwestern African coast, a city official said Tuesday.
>
> Nine people, including two children, died and 30 were injured.
>
> The accident unleashed 24,600 litres of water which dragged cars and market stalls through Melilla's streets.

- The contents of a tank can be lethal. The February 1984 release of methyl isocyanate in Bhopal, India was the worst non-nuclear industrial accident in human history. Over 2000 people were killed by the toxic vapour.

The single, most comprehensive guide to the design of vessels is the *ASME (American Society of Mechanical Engineers) Boiler and Pressure Vessel Code*[8]. This rather large document deals with all aspects of vessel design, construction and operation. Section VIII, Parts UG-125 to 136, in particular, deal with the requirements for pressure relief.

The pressure relieving requirements for non-pressure vessels, i.e. tanks, are covered in detail in *API Standard 2000, Venting Atmospheric and Low-Pressure Tanks*[5].

Most safety related design practices applying to vessels and tanks are beyond the scope of the instrumentation and controls engineer; relief valves are an exception. Their correct sizing and selection is too broad a subject to be covered in this article especially since there is already a lot of material in print concerning them. Items 9 through 14 of the references below deal extensively with this subject. The earlier section, Pressurization Systems, gives a typical example of pressure protection for an atmospheric tank.

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