

DENSE PHASE FLUID MEASUREMENT

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Many people in the industry, including probably most measurement specialists, have no experience with the measurement of dense phase fluids (if you don't count water). When the un-initiated are asked to develop or operate such a system, they tend to repeat the same mistakes others have made over and over due to trying to treat the streams like natural gas liquids or liquefied petroleum gases (NGL's or LPG's). Hopefully, this paper will assist the un-initiated reader avoid some of those mistakes.

Although definitions can be boring, I would like to cover a few that will help the student be sure they understand the fluid properties unique to dense phase fluids and eventually clarify the unique handling these fluids require.

Dense Phase Fluids, often referred to as supercritical fluids, by definition, are simply those fluids that are above their critical point at operating conditions.

The Critical Point is a combination of temperature and pressure for a particular composition fluid where the phase boundary simply ceases to exist. The fluid is neither a gas nor a liquid, but will have physical properties generally between the properties of the same stream at pressures and temperatures just below the critical point. Note that the density of the stream just above its critical point is extremely sensitive to even slight changes in temperature and pressure. As pressures rise above the critical point pressure, the rate of change of the density slows and eventually becomes tolerable from a measurement performance perspective.

Note that Dense Phase Fluids will effuse (spread out and flow) thru porous solids like a gas and they will dissolve materials like a liquid solvent.

In my experience, without a doubt, the largest single contributor to poor measurement performance when metering dense phase fluids is operating the system at a pressure too close to the critical point. In this region, density changes are so sudden and severe that our technologies can't accurately track them.

In our industry, ethylene and CO₂ are probably the two most common supercritical/dense phase fluid products. Purity ethane can be in this category too, since it's Critical Temperature is ABOUT 90 f and it's critical pressure is about 708 psia. They are often metered using orifice meters. When doing so, the orifice metering system must be operated at pressures low enough and temperatures high enough to avoid the supercritical region and successfully treat the stream like a gas – or the system must be operated at pressures high enough and temperatures low enough to be able to treat the stream like a liquid, even though it isn't.

Orifice metering systems should meet the usual industry standard recommendations when metering dense phase fluids, but you need to configure them for the typical operating condition. For example, when metering a dense phase fluid in the gas region of it's phase envelope (operating conditions), the transducer should be above the centerline of the meter run so that any liquid phase contaminants in the stream will be free to drain back into the pipeline and leave the metering system. If adequate slope is not available (one inch of drop for each foot of horizontal distance), any liquids will tend to be trapped in the meter manifold system and cause an error in the differential and static indications.

Similarly, when operating in the high pressure region well above the critical point, the orifice system needs to be operated as if it were handling a liquid stream, with the transducer below the centerline of the pipe so that any non-condensable vapors will be free to rise back to the meter taps and leave the metering system with the flowing stream.

Orifice meters operating in the gas phase region usually perform volumetric measurement. Systems operating well above the critical point may utilize volumetric or mass measurement. The general preference for this condition calls for mass measurement, where the meter indicates the volume at flowing conditions and a densitometer provides a measured density at flowing conditions (not corrected to standard conditions). In U.S. units, simply multiply the flowing volume in barrels times the density in pounds per barrel. This calculation leaves units of pounds, appropriate for mass measurement. In some cases, for a very pure stream, the assumption may be made that the full stream is one component on a \$/pound basis. For lower purity streams, the composition of the full stream will need to be determined so the % of impurities will be properly accounted for and not be treated the same as the more valuable product. Note that a calculated density at flowing conditions is an alternative source for the density input as well, but be careful using tables or calculated densities since they may be very composition dependent and/or increase overall uncertainty. API Chapter 11.3.2.1 is the source document for tables for estimating ethylene densities over a broad range of operating conditions.

Coriolis meters can provide mass measurement directly since the meter responds predictably to the mass flow rate. Proving Coriolis meters is often an issue however, since few provers are set up to prove on a mass basis (optimally, one incorporating an online density meter and compatible software). The user will need to decide whether to use a Master Coriolis meter, properly certified, to check the performance of the primary meter periodically, or to use a volumetric prover to prove the meter volumetrically at flowing conditions, then apply the Coriolis meters density output at flowing conditions(#/bbl) to multiply times the volume at flowing conditions(bbls) to get pounds of flow thru the system during a given time period.

Turbine meters are sometimes used in dense phase streams also, with the meter providing a volumetric output like the meters described above and either a calculated or measured density to develop the mass flow rate thru the system. Again, the user may chose to use volumetric measurement techniques to determine the volume at standard conditions, then use physical property tables or calculations to determine the mass of the stream determined volumetrically.

Note that it is important to have a rugged meter when working with supercritical fluids in the dense phase region. Unless start-ups and shut-downs are handled very carefully, flashing will occur and the meter will be destroyed or damaged. Turbine meters are particularly susceptible to this damage.

Ultrasonic meters can accurately produce the flowing volume at operating conditions by measuring the average flowing velocity thru the meter very precisely. Using measured density at flowing conditions times the volume metered at flowing conditions will yield the mass flow in pounds.

Another very critical concern specifically for ethylene is the potential for auto-ignition anytime pipelines or vessels are being filled and small amounts of air are still in the system. Auto-ignition can be devastating and dangerous. By definition, auto-ignition occurs when sudden pressure changes cause the ethylene to ignite even when no ignition source is available. Ethylene and oxygen alone can produce an explosion.

As with most fluids, supercritical fluids allowed to release to a low pressure from a very high pressure will produce a great deal of Joule-Thompson cooling, sufficient to cause severe injury or death, depending on the level of exposure.

Polymerization is another pitfall with ethylene streams. When the reaction occurs, the system will be loaded with a soft polymer, usually dark gray in color, which will foul instruments, meters and perhaps the entire pipeline. The system has to be mechanically cleaned when severe polymerization occurs.

In the case of ethylene, it's relatively easy to measure the stream accurately if it is a gas below it's critical pressure and temperature (pressure generally 600 psi or below and temperatures generally above 70 F.. As pressures exceed the critical point, typically about 750 psia at 50 F, you enter the "no man's land" where densities change extremely fast in response to temperature and pressure changes. If you increase operating pressure to approximately 1200 psi or so (over 1800 psia for CO2 streams), maintaining flowing temperatures as low as practical, density settles down and you can perform good mass measurement using online densitometers and a repeatable volumetric meter that has been properly proved, or an orifice

metering system that meets the recommendations of API Chapter 14.3. In a system utilizing a prover, all you need to calculate the mass flow rate is the flowing volume (indicated volume at flowing conditions – usually in barrels), flowing density (indicated density at flowing conditions – usually in pounds per barrel), a meter correction factor from proving the meter (VCF, which is dimensionless) and the density correction factor from the pycnometer proving system (DCF, which is dimensionless). The basic equation then looks like this:

$IV_f \times MCF \times D_f \times DCF = \text{Mass Flow in Pounds}$

Metering CO₂ is similar to metering ethylene, with the advantage that you don't have to worry about polymerization or auto-ignition. On the other hand, the operating pressures required to meter CO₂ above the critical point are higher than those required for ethylene. For example, I once reviewed a CO₂ system using orifice meters and calculated density which had a stock loss of almost 40%. The system was being operated at approximately 1000 to 1200 psig and 70 to 90F. The company was in the process of changing out the flow computers and upstream sections of the meter run (with improved flow conditioning), hoping to improve measurement performance, but had not implemented the changes when I arrived. By requesting system pressure be increased to 1800 psig, most of the imbalance disappeared. In fact, during the first 24 hour period after the adjustment, the system balanced within 2% of flow. By the time other issues were resolved (run lengths, beta ratios, some pulsation interfering with measurement, etc.), the system balance improved to better than +/- 1%. It was interesting to note that the system was designed for a maximum operating pressure of 2150 psig, so the engineering company must have had some experience with ethylene. Local personnel, as is so often true, had never received the information and support they should have for handling CO₂, so they immediately commented, "We always wondered why the MAOP was 2150 psig."