Optimal system design requires the right vapor pressure. Here's how to calculate it

The pressure of the liquid at the inlet of a centrifugal pump must be high enough to prevent vaporization within the pump, because this vaporization (called cavitation) hinders the pumping and can damage the impellers. This pressure requirement must be taken into account when deciding how high to place the pump feed vessel relative to the height of the pump itself.

Basically, the pump suction pressure must be greater than the fluid's vapor pressure at the pumping temperature. The difference between pump suction pressure and vapor pressure is the net positive suction head (NPSH). For cases in which the liquid contains no dissolved gases, the vapor-pressure determination is straightforward. With dissolved gases, the situation is more complicated, because vapor-pressure data for such systems are usually not at hand.

Adding to the complication is the fact that centrifugal pumps generally can, as it happens, tolerate a small amount of vapor (about 2 to 3% by volume) at the impeller eye. If the solubility of the dissolved gas is low and the temperature is far below the boiling point of liquid, the amount of vapor released in a pump is likely not to exceed the tolerable value unless the pressure reduction is substantial.

Accordingly, if the engineer plays it safe and assumes (for lack of data) that the process pressure of the liquid-gas solution in the feed drum is its vapor pressure, and if he or she consequently demands enough feed-drum elevation to fully compensate for friction losses be-

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tween the drum and the pump, this elevation may well be greater than necessary. Conversely, if the engineer ignores the presence of the gas and bases the elevation calculation on the vapor pressure of the pure liquid, the drum may not be high enough.

For an economical pumping-system design, an effective vapor pressure that lies between the process pressure and the liquid vapor pressure should be used

in calculating the available NPSH, or NPSHA. Presented here is a relatively simple method to determine that effective vapor pressure.*

Setting the stage

The method builds upon two earlier approaches to the dissolved-gas problem presented in CE. First, Penney [1] developed

*For information on dealing with liquids that carry entrained undissolved gases, see CE, December 1991, pp. 74, 75.

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basic equations to express the volume fraction of flashed gas as a function of solubility, vapor and liquid densities, liquid vapor pressure and total pressure. He used these equations to analyze some pump-performance problems created by dissolved gas. However, he did not provide an explicit method to calculate the effective NPSHA of a pump.

Four years later, Tsai [2] introduced an "artificial," or effective, vapor pressure of a gas-saturated liquid for calculating the effective NPSHA. He defined this as the pump-eye pressure at which the volume of flashed gas is 2.5% of the total volume. Then he determined this effective vapor pressure by applying trial and error to Penney's equations.

The present article offers an analytic way to calculate Tsai's effective vapor pressure without trial and error. Furthermore, it presents some curves generated from the analytic equations, to further simplify the determination when the allowable gas volume at the pump eye is in fact 2 to 3%.

Deriving the equations

When the pressure of a saturated solution is reduced from the feed drum pressure P_o to the pump eye pressure P, the volume fraction of flashed gas (including liquid vapor) can be written as:

$$f = \frac{1}{\frac{(P/P_o - P_v/P_o)^2\,(1 - P_v/P_o)}{S\,(P/P_o)\,(1 - P/P_o)} + 1} \quad (1)$$

where $S = W_o \rho_L / \rho_{Go}$ (2)

Equation (1), the basis for the analytical equations derived in the present article, was developed by Penney. He assumed that the gas obeys the ideal gas law, Dalton's law and Henry's law. W_o is the weight fraction of dissolved gas in the liquid in the pump feed drum. If solubility data or a Henry's law constant is applied, partial pressure (not total pressure) of the gas is employed to calculate W_o. The ρ_L is the liquid density, assumed to be constant (because only a small amount of gas is flashing). The ρ_{Go} is the gas density at the temperature and pressure (total pressure, not partial pressure) in the pump feed drum.

For conservative calculation, the liquid is normally assumed to be saturated with the gas, as noted earlier. However, in real-life process operation the gas-liquid contact time is frequently not long enough to reach equilibrium. If the percentage approach to equilibrium can be estimated by test data or operational experience, the amount of dissolved gas can be reduced by a saturation factor a (0 to 1) and expressed as:

$$W_a = a W_o \quad (3)$$

When the saturation factor a is included, Equation (1) becomes

$$f = \frac{1}{\frac{(P/P_o - P_v/P_o)^2\,(1 - P_v/P_o)}{S\,(P/P_o)\,(b - P/P_o)} + 1} \quad (4)$$

where $b = a + (1 - a)\,P_v/P_o$ (5)

In a saturated solution, both a and b have a value of 1.

Define a new parameter N as follows:

$$N = [f/(1 - f)]/S \quad (6)$$

Then Equation (4) can be rearranged to obtain the following equation:

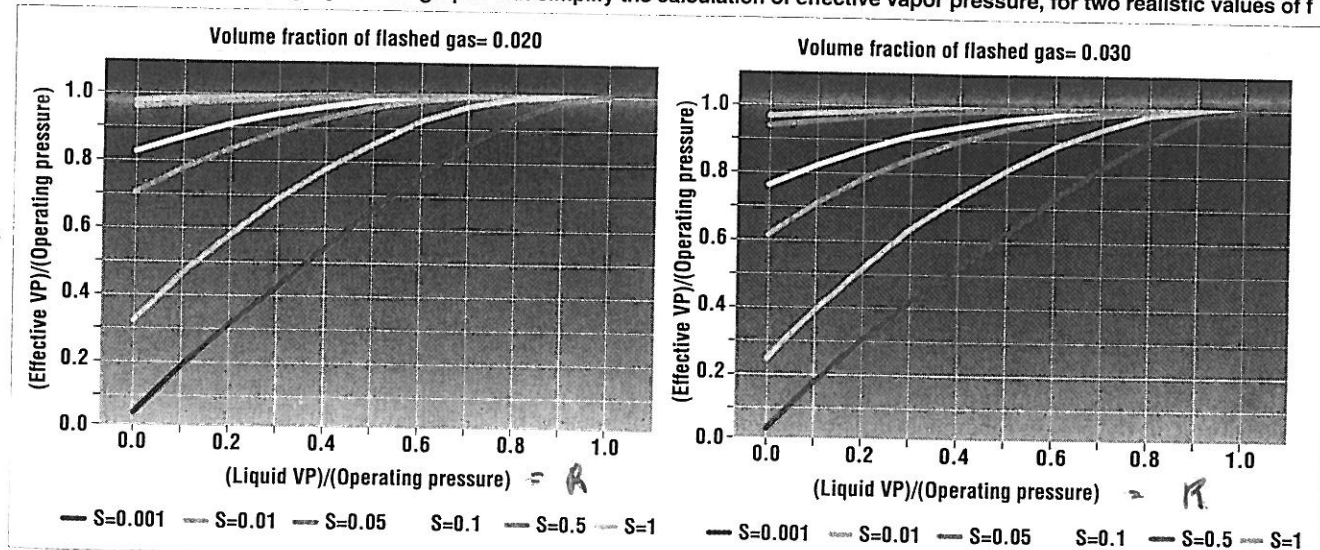$$N = \frac{(P/P_o)\,(b - P/P_o)}{(P/P_o - P_v/P_o)^2\,(1 - P_v/P_o)} \quad (7)$$

Now, let $P = P_e$, $y = P_e/P_o$, and $R = P_v/P_o$. Then Equation (7) can be written as:

$$N = \frac{y\,(b - y)}{(y - R)^2\,(1 - R)} \quad (8$$

When Equation (8) is algebraically expanded, it becomes a quadratic equation in y

$$A y^2 - B y + C = 0 \quad (9)$$

Figures 1 (left) and 2 (right). These graphs can simplify the calculation of effective vapor pressure, for two realistic values of f

Volume fraction of flashed gas= 0.020

(Effective VP)/(Operating pressure) vs (Liquid VP)/(Operating pressure) = R

— S=0.001   — S=0.01   — S=0.05   S=0.1   — S=0.5   S=1

Volume fraction of flashed gas= 0.030

(Effective VP)/(Operating pressure) vs (Liquid VP)/(Operating pressure) = R

— S=0.001   — S=0.01   — S=0.05   S=0.1   — S=0.5   — S=1

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PUMP CALCULATIONS

Example	W_o	L (lb/ft³)	G_o (lb/ft³)	f	S	N	P_v (psia)	P_o (psia)	R	A	B	C	Y	P_e (psia) Calc.	P_e (psia) Simu.
TABLE 1. SUMMARY OF CALCULATION RESULTS FOR THE EXAMPLES															
Blanket Gas (Cl) System															
Sat Low Press	4.68×10^{-4}	41.78	0.0777	0.025	0.2517	0.1019	18.65	44.7	0.4172	1.0594	1.0495	0.0103	0.9808	43.84	43.21
Sat Int Press	1.7188×10^{-3}	41.78	0.1994	0.025	0.3602	0.0712	18.65	114.7	0.1626	1.0596	1.0194	0.0016	0.9605	110.17	109.30
Unsaturated (50%)	2.34×10^{-4}	41.78	0.0777	0.025	0.1258	0.2038	18.65	44.7	0.4172	1.1188	0.8077	0.0207	0.6954	31.08	31.08
Steam Stripping System															
Low Press Column	6.5696×10^{-4}	41.58	0.0523	0.025	0.5226	0.0491	9.76	30.92	0.3157	1.0336	1.0212	0.0033	0.9847	30.45	30.25
Int Pres Column	3.9841×10^{-3}	40.38	0.2091	0.025	0.7694	0.0333	27.98	113.23	0.2471	1.0251	1.0124	0.0015	0.9861	111.66	111.20
Ammonia Wash System															
Based on NH_3	1.3723×10^{-3}	62.00	1.0047	0.025	0.0847	0.3028	1.471	364.7	0.0040	1.3016	1.0024	0.0000	0.7702	280.88	35.20
Based on H_2	4.6809×10^{-6}	62.00	0.1182	0.025	0.0025	10.4431	1.471	364.7	0.0040	11.4009	1.0839	0.0002	0.0949	34.62	
Synthesis Gas System															
Based on Individual Component:															
CO_2	2.1204×10^{-2}	62.00	6.3422	0.025	0.2073	0.1237	1.070	855	0.0013	1.1235	1.0003	0.0000	0.8903	761.22	380.00
CO	3.1000×10^{-5}	62.00	4.0342	0.025	0.0005	53.8191	1.070	855	0.0013	54.7518	1.1345	0.0001	0.0206	17.65	
H_2	$1.2000 10^{-5}$	62.00	0.2903	0.025	0.0026	10.0064	1.070	855	0.0013	10.9939	1.0250	0.0000	0.0932	79.70	
H_2S	1.1690×10^{-3}	62.00	4.9077	0.025	0.0148	1.7362	1.070	855	0.0013	2.7340	1.0043	0.0000	0.3673	314.08	
Based on Average MW and Z:															
Use Separator Gas	2.2416×10^{-2}	62.00	2.9054	0.025	0.4784	0.0536	1.070	855	0.0013	1.0535	1.0001	0.0000	0.9493	811.66	380.00
Use Dissolved Gas	2.2416×10^{-2}	62.00	8.6110	0.025	0.1614	0.1589	1.070	855	0.0013	1.1587	1.0004	0.0000	0.8634	738.21	
Use Released Gas	2.2416×10^{-2}	62.00	6.2046	0.025	0.2240	0.1145	1.070	855	0.0013	1.1143	1.0003	0.0000	0.8977	767.50	
Based on Molar-Average P_e:															
Use Separator Gas				0.025										339.97	380.00
Use Dissolved Gas				0.025										722.62	
Use Released Gas				0.025										619.52	
Use CO_2, CO and H_2 in Separator Gas				0.025										340.28	
Use CO_2 and H_2 in Separator Gas				0.025										362.80 (Best Approx.)	

where $A = N(1-R) + 1$ **(10)**

$B = 2NR(1-R) + b$ **(11)**

$C = NR^2(1-R)$ **(12)**

The meaningful solution of Equation (9) is:

$$y = \frac{B + \sqrt{B^2 - 4AC}}{2A} \qquad (13)$$

The solution with a negative sign in front of the square root cannot produce a physically meaningful result and hence is eliminated.

Once y is solved, the effective vapor is calculated as follows:

$$P_e = yP_o \qquad (14)$$

In summary, these equations can be applied to a given process situation as follows:

1. Determine the liquid density ρ_L and gas density ρ_{G_o} at the process conditions in the pump suction drum. As noted earlier, be sure to use total pressure instead of partial pressure for obtaining ρ_{G_o}

2. Determine W_o, the weight fraction of dissolved gas at the pump-feed-drum conditions, from simulation results, solubility data or Henry's law

3. Use the results of Steps 1 and 2 to calculate the solubility factor, S, by Equation (2)

4. Select a design value for the tolerable volume fraction f of vapor at the pump eye: 0.025, 0.03, or some other value suggested by the pump vendor

5. Determine the vapor pressure of the pure liquid, P_v, at the process temperature and calculate R from the relationship $R = P_v/P_o$

6. Determine the saturation factor a (between 0 and 1; use 1 if the factor is not known) and calculate the saturation coefficient b by Equation (5)

7. Use the values of S and f obtained above to calculate the parameter N by Equation (6)

8. Use the results of Steps 5, 6, and 7 to calculate coefficients A, B, and C by Equations (10), (11), and (12)

9. Calculate y by Equation (13), and then the effective vapor presssure P_e by Equation (14)

Some calculated results appear in Figures 1 and 2 for $f = 0.020$ and 0.030, respectively. One can see from the curves that the difference between the process pressure and the effective vapor pressure becomes significant only when the gas solubility is low.

The presence of dissolved gases crops up very widely throughout the process industries. As the following examples bring out, the engineer should neither overlook the gases nor, on the other hand, overreact to them.

The water-air system

When water is saturated with air at atmospheric pressure, the solubility of air and the physical properties of the system are functions of temperature only. The effective vapor pressures corresponding to f values (flashed air volume fractions) of 2, 2.5, and 3% are plotted against temperature in Figure 3 along with the vapor pressure curve for pure water.

As can be seen, the effective vapor pressure of water saturated with air at ambient temperature is significantly higher than the water vapor pressure. In fact, the effective vapor pressure is roughly equal to the average of atmospheric pressure (14.7 psia) and water-vapor pressure in the temperature range between 40 and 170°F. To see the impact of this difference, consider this example:

A centrifugal pump is used to move water from an atmospheric storage tank at 80°F. The allowable volume percent of flashed gas is 2.5%. The water density is 62.223 lb/ft³. As Figure 3 shows, the effective vapor pressure is 6.73 psia while the water vapor pressure is only 0.5068 psia. Assume that the pump suction line

has a friction loss of 0.5 psi. Calculate NPSHA without taking the dissolved air into account. Then recalculate it correctly.

Solution: Let h stand for the suction-side static head, in feet. The formula for NPSHA is:

$$NPSHA = (P_o - P_v - \text{friction loss, in feet of liquid}) + h \qquad (15)$$

a. Ignoring the dissolved air:

NPSHA $= (14.7 - 0.5068 - 0.5) \times 144/62.223 + h = (31.7 + h)$ ft

b. Taking dissolved air into account (i.e., using the effective vapor pressure):

$(14.7 - 6.73 - 0.5) \times 144/62.223 + h = (17.3 + h)$ ft

The difference between Case a and Case b is about 14 ft.

Pumping water from a tank that is higher than the pump usually offers no NPSH problem, and the error of using the vapor pressure of water instead of the effective vapor pressure may not make a significant difference. If water is instead to be pumped from a level below the pump, ignoring the dissolved gas may prove to be serious — in the present example, the lifting capability of the pump would be overstated by about 14 ft.

Penney [1] reports an incident involving a cooling tower pump. Although the calculated NPSHA (based on pure-water vapor pressure, ignoring the dissolved air) was 9.4 ft greater than the NPSH required according to the manufacturer's performance curves, the pump produced the gravelly noise that is typical of cavitation. Mechanical damage occurred to the impeller, which failed in about two years.

One-component systems

When a blanket of inert gas is employed over a liquid to maintain a desired process pressure above the liquid vapor pressure, some of the gas will dissolve in the liquid. As the liquid enters the pump, part of the dissolved gas will flash if the liquid static head is insufficient to overcome the pressure drop between the tank and the pump. This may cause severe pump cavitation if the amount of flashed gas is substantial. Consider these three blanket-gas cases:

1 - Saturation at low pressure: A petroleum distillate stream having a density of 41.78 lb/ft^3 and vapor pressure of
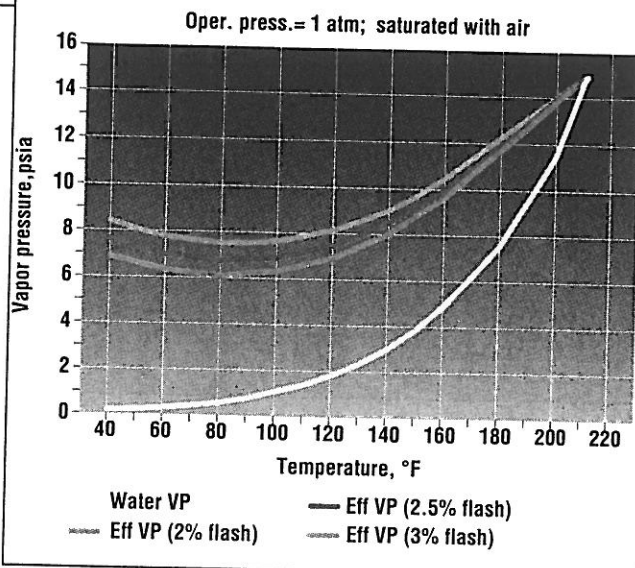

Oper. press.= 1 atm; saturated with air

Figure 3. The effect of dissolved air in water can be dramatic

Water VP   Eff VP (2.5% flash)
Eff VP (2% flash)   Eff VP (3% flash)

18.65 psia is pumped from a surge drum to a hydrotreater unit. The surge drum is at 400°F and 30 psig. This pressure is maintained by a blanket of natural gas (assumed to be pure methane). Assuming that the liquid is saturated with the blanket gas, the effective vapor pressure corresponding to 2.5% flashed gas volume, calculated by the sequence outlined above, is 43.84 psia. This value is very close to the result of 43.21 psia obtained from direct simulation, and is considered very accurate. The results of analytic calculation and process simulation for this and the following examples are summarized in Table 1, to show the accuracy of this method.

As in the first example, use 0.5 psi for pump-suction-line friction loss and h ft for suction static head for this and the following examples. With the use of Equation (15), the values of NPSHA calculated based on different vapor pressures are:

NPSHA based on $P_v = (44.7 - 18.65 - 0.5) \times 144/41.78 + h$
   $= 88.06 + h$ ft
NPSHA based on $P_e = (44.7 - 43.84 - 0.5) \times 144/41.78 + h$
   $= 1.24 + h$ ft
NPSHA based on $P_o = (44.7 - 44.7 - 0.5) \times 144/41.78 + h$
   $= -1.72 + h$ ft

If the required NPSH (NPSHR) is 10 ft, and a 2-ft design margin is stipulated, the required static heads above pump centerline are as follows:

Based on P_v:
   No positive static head is required

Based on P_e:
   A static head of 10.8 ft is required
Based on P_o:
   A static head of 13.7 ft is required

The result obtained by using liquid vapor pressure in NPSHA calculation, namely that no positive static head is required, is too optimistic. The result based on using the process pressure is conservative, though not too far off — the vessel support height thus calculated is unnecessarily high by about 3 ft.

2 - Saturation at intermediate to high pressure: If the pressure of the gas blanket system is increased to 100 psig, the difference between operating pressure and effective vapor pressure will be intensified. Although the calculated effective vapor pressure (110.17 psia) is as high as 96% of the operating pressure, the 4% difference represents 4.5 psi, which converts to a 16-ft difference in vessel-support-height requirement. Specifically, the calculated NPSHA's based on different vapor pressures are:

   $329.33 + h$ ft based on P_v; no positive static head is required
   $13.89 + h$ ft based on P_e; no positive static head is required
   $-1.72 + h$ ft based on P_o; a positive static head of 13.7 ft is required

These results indicate that for the intermediate-pressure blanket-gas system, the vessel-support-height requirement based on P_o is too conservative and can be drastically reduced (by about 16 ft) to save the cost. This difference between P_o and P_e increases as P_o increases. Thus, in order to minimize the cost, the allowance of a small amount of flashed gas in the

Table 2: MULTICOMPONENT SYSTEMS REQUIRE SPECIAL TREATMENT			
		In Liquid	
Component	Vapor Mole %	Mole %	lb/100 lb H_2O
CO_2	38.40	92.10	2.1204
CO	6.46	0.21	0.0031
H_2	54.04	1.13	0.0012
H_2S	1.10	6.56	0.1169

pump should particularly be considered in pumping system design when the pressure is high.

3 - Unsaturation: Realistically, the liquid leaving a surge drum may not have reached equilibrium with the blanket gas. However, it is a prudent common practice to assume the liquid is saturated, unless test data are available to determine the degree of saturation. This example is given merely to examine the difference under saturated and unsaturated conditions.

Assume that the degree of saturation as determined from test data is 50% (that is, $a = 0.5$) for the low-pressure blanket gas system ($P_o = 30$ psig) of Case 1. Then the calculated effective vapor pressure will be only 31.08, psia which is 13.6 psi lower than the operating pressure. The NPSHA calculated from the effective vapor pressure becomes

$$(44.7 - 31.08 - 0.5) \times 144/41.78 + h = 45.22 + h \text{ ft}$$

This result indicates that for the low-pressure blanket system, the NPSHA with $a = 0.5$ is about 44 ft greater than that with $a = 1$. This difference increases as the pressure increases.

Coping with highly soluble gases

A circulating water system, propelled by a centrifugal pump, is used to remove ammonia from a hydrogen-rich gas stream at 350 psig and 115°F. This system is somewhat different from those discussed before — the gas contains over 85 mole % of hydrogen, which has low solubility in water, and 0.04 mole % of ammonia, which has high solubility.

Although the contacted gas contains much more hydrogen than ammonia, the portion that dissolves in water is mainly ammonia. In other words, the key component of the contacted gas is hydrogen but that of the dissolved gas is ammonia. The question arises, which one should be used in a single-component approximation? It is

answered via the following investigation:

The calculated effective vapor pressure corresponding to $f = 0.025$ is 34.6 psia when only hydrogen is considered, and 280.9 psia when only ammonia is considered. A direct flash simulation reveals that the real effective vapor pressure is 35.2 psia.

This means that hydrogen, the main component of the feed gas, should be used as the single component in calculating the effective vapor pressure by the analytic method. Ammonia cannot be used, because the absorbed ammonia will stay in water when the pressure is reduced. Only the dissolved hydrogen will escape from liquid.

Note that the effective vapor pressure of the ammonia wash system is substantially lower than the operating pressure. Neither the NPSH nor the height of the wash-tower skirt should be a concern for pumping ammonia wash water around when the tower pressure is sufficiently high.

Multicomponent systems

To evaluate the application of this method to multicomponent systems, consider a system in which the pump handles water from a synthesis gas separator. The water, at 840 psig and 104°F, is saturated with gas of the water-free composition that is shown in Table 2.

The difference from the ammonia wash system is that the synthesis gas contains more than one major component. Therefore, a single-component approximation can not be successfully applied.

A similar example was discussed by Tsai in Reference [2]. There, the average molecular weight and average compressibility factor of the dissolved gas mixture were used to obtain the specific volume of gas for calculating the solubility factor and the effective vapor pressure. When that approach is used to calculate the effective vapor pressure in the present

case, the result obtained is 738 psia. This value is quite close to the 742 psia obtained by Tsai, because the two systems are very close in composition.

The accuracy of these results can be checked by running a process simulation, in which the saturated liquid is flashed to a lower pressure. The pressure is adjusted until the volume of the flashed vapor equals to 2.5% of the total volume. The final pressure proves to be 380 psia. In other words, when the approach in this article is applied based upon average molecular weight and average compressibility factor, the calculated effective vapor pressure is almost twice the true value (obtained by process simulation) for the synthesis gas mixture.

Various averaging methods have been examined in an effort to bring the calculated value close to the simulation result. The following approach results in the closest value, 363 psia:

1. Use the solubility and density of each individual major component (CO_2 and H_2) to calculate the effective vapor pressure of that component. Ignore the minor components of the mixture.
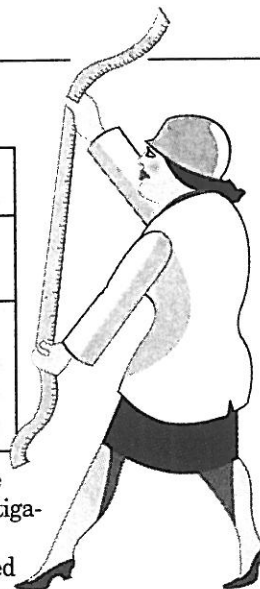
2. Calculate the molar-average effective vapor pressure of the major components (CO_2 and H_2) by summing the products of component effective vapor pressure (from Step 1) and component mole fraction.

This works well for the synthesis gas system. But it might not work for other ones. Therefore, when dealing with multicomponent systems, process simulation is recommended for obtaining the effective vapor pressure.

Don't ignore liquids

The caution not to overlook dissolved gases can analogously be applied to dissolved liquids in liquid-liquid systems of low miscibility. Consider, for instance, the case when steam stripping removes light components from a liquid hydrocarbon stream in a distillation column, and the bottoms product is accordingly saturated with water. The effects of dissolved water on product-pumping NPSHA have been studied for a low-pressure, steam-stripping fractionator and for an intermediate-pressure steam stripper. The results are summarized in Table 1.

The effect of dissolved water on pumping a steam-stripped liquid (of lower volatility than that of water) is very similar to the effect of dissolved gas on pump-

ing a gas-blanketed liquid. When the system pressure is low, using either the effective vapor pressure or the operating pressure as vapor pressure in calculations of NPSHA and vessel-support height will obtain close results. However, when the pressure is high, using operating pressure as vapor pressure in NPSHA calculation will result in a very conservative design and is not recommended.

In the example concerning the intermediate-pressure (113.23 psia) steam stripper, about 6 ft of column skirt height can be saved if the effective vapor pressure in lieu of operating pressure is used in the design calculation. In general, the difference between operating pressure and effective vapor pressure for a steam stripping unit is not so great as that for a blanket gas system operated at same pressure level.

Summing it up

A centrifugal pump generally can tolerate 2-3% by volume of flashed vapor at the impeller eye. To take advantage of this tolerance when pumping a liquid saturated with low-solubility gas, an effective vapor pressure lower than the total vapor pressure (operating pressure) can be used to calculate the NPSHA and vessel-support-height requirement. This effort in some cases can result in a significant cost reduction.

This effective vapor pressure can be determined without trial and error by the method outlined in this article. For many situations, it can instead be read directly from Figures 1 or 2.

Water-air systems: In a water-air system, pumping water from a level above the pump will not incur any NPSH problem, and the dissolved air can be ignored. However, if water is pumped from a level below the pump, the effect of dissolved air on pump lifting capability should be investigated. Figure 3 can be used to obtain the effective vapor pressure for correct NPSHA and pump-lifting-height calculations.

One-component systems: For blanket-gas systems or other one-component-gas situations, the effective vapor pressure corresponding to 2.5-3.0% of allowable vapor volume of liquid saturated with the gas is generally higher than 95% of the operating pressure. For a low-pressure system, the difference between the process pressure and the effective vapor pressure is in-

significant, and the process pressure can be used in NPSHA calculation without significant cost impact.

For an intermediate- to high-pressure (say higher than 100 psig) system, the absolute difference between the operating pressure and the effective vapor pressure becomes significant, even though the percentage difference between the two remains small. In this case, the effective vapor pressure should be used in NPSHA calculations to prevent providing unnecessary vessel height.

If test data are not available, it is recommended that the engineer assume the

NOMENCLATURE

A	Coefficient in Eq. (9) as defined in Eq. (10)
a	Saturation factor [(dissolved gas at unsaturated condition)/(dissolved gas at saturated condition)]
B	Coefficient in Eq. (9) as defined in Eq. (11)
b	Saturation coefficient defined in Eq. (5)
C	Coefficient in Eq. (9) as defined in Eq. (12)
f	Volume fraction of vapor at pump eye [(volume occupied by flashed gas and vaporized liquid)/(total volume)]
N	Parameter defined in Eq. (6)
NPSH	Net positive suction head, ft of liquid
NPSHA	Net positive suction head available, ft of liquid
NPSHR	Net positive suction head required, ft of liquid
P	Pressure, psia
P_e	Effective vapor pressure corresponding to an allowable f, psia
P_o	Operating pressure of pump suction drum, psia
P_v	Vapor pressure of liquid (without dissolved gas), psia
R	Ratio of P_v and P_o
S	Solubility factor defined in Eq. (2)
W_a	Weight fraction of dissolved gas at total pressure of P_o
W_o	Weight fraction of dissolved saturation gas at total pressure of P_o
y	Ratio of P_e and P_o
ρ_{Go}	Density of gas at pump suction drum pressure (P_o) and temperature, lb/ft³
ρ_L	Liquid density, lb/ft³

liquid to be saturated with the blanket gas ($a = 1$).

Highly soluble gas present: For ammonia wash and similar systems, if the gas entering a pump contains mainly a low-solubility component and only a small amount of high-solubility constituent (such as ammonia), the latter will stay in solution and will not flash unless the pressure is drastically reduced. For this kind of system, the effective vapor pressure is likely to be much lower than the operating pressure, and the pump NPSH is generally not a problem. The effective vapor pressure can be obtained by considering only the dominating low-solubility inert component in the gas and ignoring the existence of other components.

Multicomponent systems: Since Penney's equations [1] (upon which the method outlined in this article is based) are derived for a single-component gas, application to liquid saturated with a gas mixture containing several major components (e.g., synthesis gas) is not reliable. Process simulation is the only trustworthy method for obtaining the effective vapor pressure of a multicomponent system.

Application to liquids: Systems in which water mixes sparingly with a lower-volatility liquid, such as a steam-stripping system, are quite similar to a blanket gas system. The effective vapor pressure corresponding to 2.5-3.0% of allowable vapor volume of liquid at bottom of the steam stripper is generally higher than 98% of the operating pressure. It is practical to use the operating pressure as vapor pressure in NPSHA calculation, except for high-pressure systems. ∎

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