CALC. TO DETERMINE NEED

FOR A

N2 PHASE SEPARATOR

D-ZERO ENGINEERING NOTE # 3823.115 -EN- 422

April 7, 1995

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RD/DØ Mech.

Approved : Kurt J Krempetz

Summary

A nitrogen phase separator is recommended on the liquid supply line at the helium refrigerator plant. This engineering note documents the calculations done to reach that conclusion.

Method

The steady state liquid nitrogen consumption rate for the refrigerator, VLPC and solenoid systems is about 30 gal/hr. The estimated heat leak for the piping run to the refrigerator location is 50 watts.

The calculated quality at the refrigerator was 0.032. Given this quality, a two phase flow model based on Lockhart-Martinelli and also incorporating Baker diagram nomenclature was run on TK solver.[1] The result of this program was that without the use of a phase separator we could expect a slug flow pattern with a volume fraction of gas of 65%. Based on this, I recommend that we use a phase separator to siphon off the gas before the nitrogen is sent to a standard saver type subcooler. Including the phase separator will help ensure proper operation of the subcooler . The subcooler will help us attempt to deliver single phase liquid to the nitrogen control valves.

The raw calculations follow.

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[1]This TK solver program was developed a few years ago and was verified during the solenoid design report era. See D-Zero engineering note 3823.111-EN-338.

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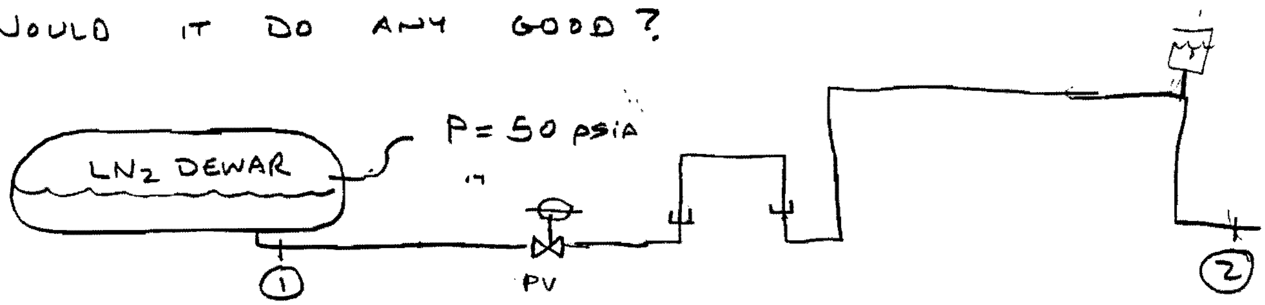
SUBJECT LN_2 PHASE SEPARATOR		NAME RUSS RUCINSKI	DATE 4-7-95	REVISION DATE

LOOK AT THE NEED FOR A LN_2 PHASE SEPARATOR.
WOULD IT DO ANY GOOD?



HEAT LOADS:

400 Ft FROM DEWAR TO REFRIG. × .1 W/Ft = 40 W
4 VALVES ≈ 1.5 W × 4 = 6 W

SAY 50 WATTS

EXPECTED FLOW RATE:

REFRIG. = 10 gal/hr
VLPC = 18 gal/hr
SOLENOID = 2 gal/hr
} 30 gal/hr

STATE (1)

P = 50 psia ≈ 3.5 ATM
SAT. LIQUID.
$h_f = -95.175$ J/g

STATE (2)

P = 45 PSIA ≈ 3 ATM
$h_f = -98.938$ J/g
$h_g = 84.167$ J/g

Volume = 1.339 cc/g

$$\dot{m} = 30\,\frac{gal}{hr} \times \frac{3785.4345\ cm^3}{1\ gal} \times \frac{1\ hr}{3600\,s} \times \frac{1}{1.339\ \frac{cm^3}{g}} = 23.56\ g/s$$

$$Q_{12} = \dot{m}(h_2 - h_1)$$

$$h_2 = \frac{Q_{12}}{\dot{m}} + h_1 = \frac{50\ J/s}{23.56\ g/s} + -95.175\ \frac{J}{g} = -93.053\ J/g$$

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$$\chi_2 = \frac{h_2 - h_f}{h_g - h_f} = \frac{-93.053 + 98.938}{84.167 + 98.938} = \underline{.03214}$$

$$\rho_{2l} = \frac{1}{1.323\ cc/g} = 0.75586\ \frac{g}{cm^3} \times \frac{1\ kg}{1000\ g} \times \frac{1000\ L}{1\ m^3} \times \frac{1000\ cm^3}{1\ L}$$

$$= 755.86\ kg/m^3$$

$$\rho_{2g} = \frac{1}{78.013\ cc/g} = .01282\ g/cm^3 = 12.82\ kg/m^3$$

PLUG NUMBERS INTO TK 2 PHASE FLOW MODEL.

PER; BAKER DIAGRAM WE ARE IN SLUG FLOW REGIME.

PER LOCKHART MARTINELL VOLUME FRACTIONS ARE:

$$V_L = 34.8\%$$

$$V_{gas} = 65.2\%$$

$$\Delta P_{100\ ft} = 0.24\ psi$$

1/2" SCH. 10 PIPE

I CONCLUDE WE SHOULD PUT A PHASE SEPARATOR BEFORE THE SUBCOOLER.

St	Input	Name	Output	Unit	Comment
					Lockhart Martinelli correlation for two phase pressure drop through an adiabatic horizontal pipe
					NITROGEN FOR REFRIG. UPGRADE
					Russ Rucinski 4/7/95
					Steady state flow conditions
					Demonstrates need for phase separator
		valid	'accurate		model validity
		flow	'turbulen		flow type
		f	.01692739		friction factor ; Must enter a guess
	.000005	epsilon		ft	pipe roughness
	.674	D		in	Pipe inside diameter, 1/2" sch. 10 pip
	100	L		ft	Length of pipe
		A	.00023019	m^2	cross sectional area
	23.56	mdot		g/s	Total mass flow rate
	755.86	rhoL		kg/m^3	Liquid density
	12.82	rhog		kg/m^3	density of the gas
	10.5	muL		μPa-s	Liquid viscosity
	6.12	muG		μPa-s	Gas viscosity
	.0062	sigmaL		N/m	Surface tension of liquid
	.0324	x			quality = mdotG/mdot
		ReL	161500		Reynold's # for liquid
		mdotL	.02279666	kg/s	Mass flow rate for liquid
		ReG	9277		Reynold's # for gas
		mdotG	.00076334	kg/s	Mass flow rate for gas
		dpdLL	6.4152443	Pa/m	ΔP/ΔL for the liquid
		dpdLTP	53.013559	Pa/m	ΔP/ΔL for the two phase
		deltaP	.23436336	psig	Total pressure drop for the pipe
		phiL	2.8746626		Lockhart-Martinelli parameter
		X	2.8025469		Lockhart-Martinelli parameter
	.25	m			L-M constant; Look up in Table 7.19
	.2	n			L-M constant; Look up in Table 7.19
	.316	CG			L-M constant; Look up in Table 7.19
	.184	CL			L-M constant; Look up in Table 7.19
	20	C			L-M constant; Look up in Table 7.19
		lambda	2.8445187		Baker diagram dimensionless parameter
	1.2	rhoair		kg/m^3	density of air
	998	rhoH2O		kg/m^3	density of water
		sigh	3.1030377		Baker diagram dimensionless parameter
	.073	sigmaH2		N/m	Surface tension of water
	1000	muH2O		μPa-s	viscosity of water
		BakerXa	264		X - axis value for Baker plot
		BakerYa	860	lbm/hr-ft	Y - axis value for Baker plot
		c	1		
		M	1		
		RsubL	.34786691		Volume fraction of liquid phase
		RsubG	.65213309		Volume fraction of gas phase
		GasVel	.39666047	m/s	Gas velocity
		LiqVel	.37665137	m/s	Liquid velocity

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Liquid	Vapor	C
Laminar	Laminar (*vv*)	5
Turbulent	Laminar (*tv*)	10
Laminar	Turbulent (*vt*)	12
→ Turbulent	Turbulent (*tt*)	20 ←

The friction factor is calculated from eqn. (3.59) for laminar flow, or from eqn. (3.60) or eqn. (3.61) for turbulent flow, using the Reynolds number Re_L.

For diabatic flow (flow with heat transfer to or from the system), the pressure gradient at each point within the tube is given by a modification of eqn. (7.56) (Martinelli and Nelson 1948):

$$(dp/dL)_{TP} = (1 - x)^{2-n}\phi_L^2(dp/dL)_0 \qquad (7.61)$$

where x is the local value of the fluid quality and $(dp/dL)_0$ is the pressure drop per unit length that would exist if the liquid were flowing alone at the total mass flow rate $(\dot{m}_L + \dot{m}_G)$.

Because the addition of heat causes a change in the fluid quality along the length of the tube, the total frictional pressure drop must be determined by numerical integration of eqn. (7.61), or

$$\Delta p_f = \int_0^L \left(\frac{dp}{dL}\right)_{TP} dL = \left(\frac{dp}{dL}\right)_0 \int_0^L (1 - x)^{2-n}\phi_L^2\, dL$$

If the axial heat flux is constant (as is frequently the case for cryogenic-fluid transfer lines), we may write

$$dx/dL = (x_2 - x_1)/L$$

Table 7.19. Lockhart-Martinelli correlation constants

Constant	Laminar $Re < 2000$	Turbulent $3000 < Re < 50\,000$	Turbulent $Re > 50\,000$
m (vapor)	1	0.25	0.20
n (liquid)	1	0.25	0.20
C_G (vapor)	64	0.316	0.184
C_L (liquid)	64	0.316	0.184

The four possible combinations are:
vv = laminar liquid, laminar vapor
vt = laminar liquid, turublent vapor
tv = turbulent liquid, laminar vapor
tt = turbulent liquid, turbulent vapor

$Re_{GAS} = 9277$

$Re_{LIQ} = 1615CC$

where subscript 1 refe
With this substitution,

$\Delta p_f =$

Because of the chan
bulk fluid velocity, the
as a result of the accel
momentum pressure d

where the momentum-

$\phi_M = \frac{(1 - x_2)^2}{R_{L2}} -$

The quantity R_L is th
denotes inlet condition
parameter is a function

The total pressure drop
tional and momentum

Example 7.8. Determine
phase hydrogen in a 150-m
mass flow rate of 4.00 kg/s
hydrogen is flowing at an a
the line is 160 m (525 ft).

For hydrogen at 354.6 k
find the following fluid pro

$\mu_L = 10.34\ \mu Pa$
$\mu_G = 1.367\ \mu Pa$
$\sigma_L = 1.120$ mN

The cross-sectional area for

$A = \frac{1}{4}\pi$

The total mass flow rate per

$\dot{m}/A =$

The Reynolds numbers for

$Re_L = (0.1627)(19$
$Re_G = (0.1627)(19$

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R. Rucinski
4-7-95

△ – LN_2 SUPPLY @ REFRIGERATOR WITH NO PHASE SEPARATOR

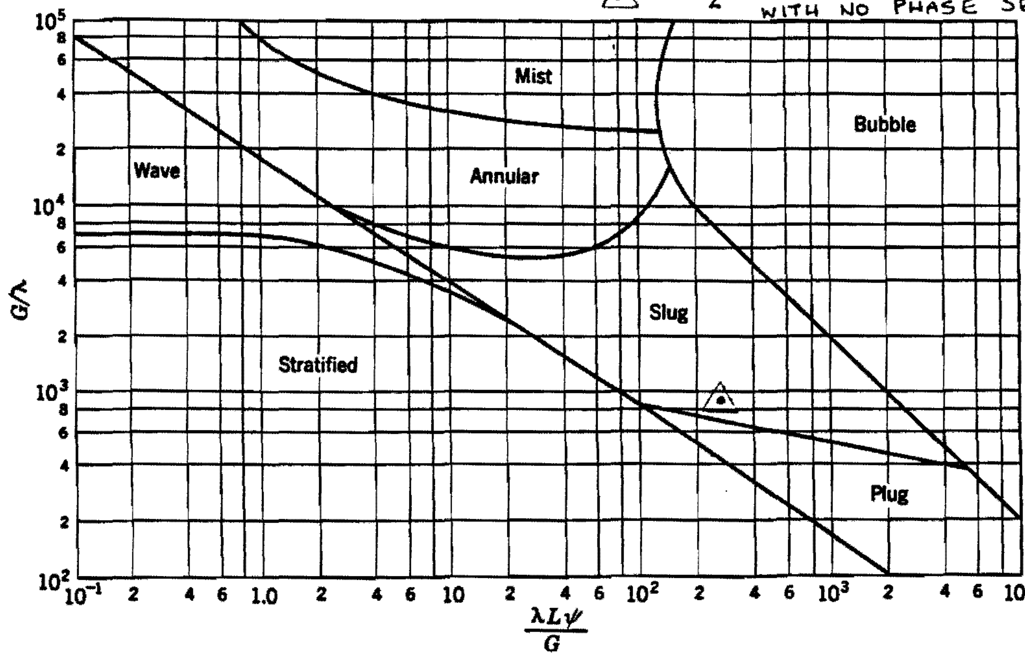

Fig. 7.26. Two-phase flow pattern regions according to Baker (1954).

$\lambda = [(\rho_G/\rho_a)(\rho_L/\rho_w)]^{1/2}$
$\psi = (\sigma_w/\sigma_L)[(\mu_L/\mu_w)(\rho_w/\rho_L)^2]^{1/3}$
L = liquid mass flow rate per unit area, $lb_m/hr\text{-}ft^2$
G = vapor mass flow rate per unit area, $lb_m/hr\text{-}ft^2$
ρ_G = vapor density
ρ_L = liquid density

ρ_a = density of air = 1.20 kg/m^3 = 0.075 lb_m/ft^3
ρ_w = density of water = 998 kg/m^3 = 62.3 lb_m/ft^3
σ_L = surface tension
σ_w = water surface tension = 0.073 N/m
μ_L = liquid viscosity
μ_w = water viscosity = 0.001 Pa-s = 1 centipoise

disintegrates into a mist or spray dispersed within the vapor phase; thus, *mist flow* results.

For low-quality flow, vapor bubbles are formed within the liquid phase, and *bubble flow* is achieved. For horizontal flow, the bubbles are usually concentrated in the upper portion of the tube; however, in vertical flow, the bubbles may be dispersed throughout the liquid. Increasing the vapor content of the flowing stream causes the bubbles to collect together into plugs of vapor that flow at intervals along the top of the pipe; hence, the name plug flow is applied to this flow pattern.

A complete disc
by Collier (1972).

7.21. Pressure dr

Two fundamental
drop in two-phase
phases are treatec
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considered to be
Martinelli (1949)
separated-flow mc
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drop for such cry
nitrogen (Shen anc
sider the Lockhart
The parameters
data according to

$(\Delta p/\Delta$

where $(\Delta p/\Delta L)_{TP}$ i
and $(\Delta p/\Delta L)_L$ and
would exist if the
alone. In the expre
and Re_G are calcul

Re_L

where D = tube i
μ = viscos
A = cross-
$\dot{m}$ = mass

The constants C_L
These parameters
the individual ph
The Lockhart-
expression:

where the constan
for liquid alone in