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GAS REMOVAL SYSTEMS
ASSOCIATED WITH DREDGE
PUMPS

by
P. Aarne Vesilind

John B. Herbich

Fritz Engineering Laboratory Report

No. 310.2

CIVIL ENGINEERING DEPARTMENT
FRITZ ENGINEERING LABORATORY
HYDRAULICS DIVISION

GAS REMOVAL SYSTEMS ASSOCIATED
WITH DREDGE PUMPS: LITERATURE SURVEY

Status Report No. 2

Prepared by

P. Aarne Vesilind
and
John B. Herbich

Prepared for
U. S. Army Engineer District, Philadelphia
Corps of Engineers
Philadelphia, Pennsylvania

April 1964
Bethlehem, Pennsylvania

Fritz Engineering Laboratory Report No. 310.2

P R E F A C E

The following report summarizes the studies performed during the months of March and April 1964, at the Hydraulics Division of Fritz Engineering Laboratory, under terms of Contract No. DA-36-109-CIVENG-64-72. Earlier studies were described in Fritz Laboratory Report No. 310.1 dated February 1964.

Dr. John B. Herbich is the project director and is assisted by Professor W. P. Isaacs, Dr. A. W. Brune, Mr. Hugh Murphy, Mr. Brian Van Weele and Mr. Aarne Vesilind. Professor W. J. Eney is head of the Civil Engineering Department and Fritz Laboratory and Dr. Lynn S. Beedle is the Director of Fritz Engineering Laboratory.

I N T R O D U C T I O N

The modern sea-going hopper dredge is the result of progressive development which has taken place during the past century. It has found increasing importance in improving harbors and seaways both in peace time and during war. The majority of hopper dredge in the United States are of the hydraulic suction type, equipped with special Machinery enabling them to dredge material from the ocean bed or channel bottom, discharge it into hoppers, transport it and dump it at disposal sites.

The heart of the dredge is the pump. This pump is of the centrifugal radial type and must be designed to withstand heavy wear and abrasion. While in operation, it may encounter a variety of mixtures made up of liquids, solids, and gases. No particular difficulty is encountered when the mixture is composed of solids and liquids, except when the density of the mixture becomes too high. This may occur if the drag is buried too deep in mud at the channel bottom, causing the pump to choke. This condition is remedied by lifting the drag arm.

When material containing a considerable amount of gas is encountered, the gas drawn into the suction line causes appreciable decrease in vacuum and volume of solids discharged. The remedy for this is identical to that of choking, and for many years the difference between choking and unloading due to gas was not recognized.

In recent years, however, gas removal equipment has been installed on many dredges and the efficiencies have been increased considerably. These systems are not totally effective and better methods must be developed to both

increase the amount of gases removed and lower the cost of such an operation.

As a result, the U. S. Army Corps of Engineers entered into a contract with Hydraulics Division of Lehigh University to conduct a study to learn more about the mechanics of a problem and recommend a more efficient method of gas removal. The following report is the second of a series of preliminary studies and is devoted principally to the survey of literature which may be of value in the later stages of the project.

P R E S E N T G A S R E M O V A L
S Y S T E M S O N D R E D G E S

Gas samples taken from dredged material indicate that the composition of the gas may be about 85% methane and 15% carbon dioxide.

The present gas removal systems are all basically alike. Accumulators are installed near the pump or at the highest point on the suction line and the gas is ejected from the accumulators either by vacuum pumps or water or steam ejectors. The accumulation of the gas is possible because most of it rises to the top of the suction pipe as it is being brought up from the bottom. The accumulators are about six foot high and have diameters approximating those of the suction lines.

L I T E R A T U R E S U R V E Y

The literature reviewed in this study can be divided into three general areas: general considerations of two-phase flow; gas accumulation devices; and gas ejectors.

I TWO-PHASE FLOW CONSIDERATIONS

A. GENERAL

If actual conditions are to be duplicated in the laboratory, gas has to be introduced into the suction line. This can be accomplished with an aspirator, injecting gas parallel to and in the middle of the pipe⁽¹⁹⁾.

Other investigators found that injecting the air vertically from the top of

the pipe and perpendicular to it resulted in a good distribution of bubbles⁽¹²⁾. It has been discovered that the bubble size and distribution is dependent on a number of conditions such as the size of the emitting orifice, the viscosity of the liquid, etc.⁽⁴⁾ Air bubble resorption has also been studied and analyzed⁽¹⁸⁾. The measurement of the flow rate in two-phase flow can be accomplished in two ways. One method requires measuring the flow of liquid and gas separately and then combining the two streams. Most of the experiments performed on two-phase flow used this method. The second method is to measure the two-phase flow itself by means of an orifice meter⁽¹³⁾⁽²⁴⁾. This method, however, would be cumbersome to use and would also change the nature of the flow.

The absorption of gases in liquids will also be of importance in the laboratory. This has been studied quite extensively and is presented in many text books. Two articles of interest are included in the abstracts⁽⁷⁾⁽¹⁷⁾.

B. PRESSURE DROPS IN HORIZONTAL PIPES

There are at present two basic methods by which two-phase flow can be analyzed. The first developed by Martinelli and co-workers uses a relatively empirical approach. The second method utilizes a single friction factor for both phases together and is hence known as the "homogeneous flow" approach.

1. THE MARTINELLI METHOD

With this method⁽¹²⁾⁽³⁵⁾⁽¹¹⁾ it is possible to obtain pressure drop accuracies to $\pm 30\%$ by using two coefficients, X and ϕ , which are derived

from the following premises: (a) the static pressure drop for the liquid phase is equal to the gaseous phase static pressure drop, (b) the volume of the pipe at any instant is equal to the volume of gas plus the volume of liquid in the pipe.

It was also recognized that both the gas and the liquid could flow in either the turbulent or laminar range. This resulted in four combinations. The fully turbulent flow is the only one considered in the abstracts since this condition will exist in testing. Sample calculations of this method are included in the abstracts⁽²⁾.

Even though at present the Martinelli method seems to be the most reliable, it does have some shortcomings. Some investigators feel that not all of the flow rate parameters, such as frictional losses, momentum changes, etc., are fully considered⁽⁹⁾.

Since the angle of the pipe carrying the two-phase flow is important in our investigation, it is noted that his method is quite applicable to pipes on a slant by merely correcting for the static head⁽³⁵⁾⁽¹¹⁾.

Many investigators have used the Martinelli correlation with varying degrees of success. Levy⁽³³⁾ gave some theoretical backing to the method and achieved results of +20% accuracy. Baker⁽¹⁾ and Van Wingen⁽⁴⁰⁾ applied this to oil pipelines and with some modifications achieved reasonable accuracy. Other investigators also found the method acceptable⁽²⁹⁾⁽⁴³⁾.

Since the analysis was derived for isothermal flow only, experiments have been performed with the idea of adapting it to non-isothermal flow⁽²²⁾
(27)(41)(23)(36). In most cases this met with success.

2. THE HOMOGENEOUS APPROACH

The basic premise in this method is the assumption that the velocity of the gas and the liquid is equal and hence one friction factor can be used for the entire two-phase flow. Since this condition seldom exists, the method is open to argument.

One of the best analysis of this method is presented by James and Silberman⁽¹⁰⁾. They also found that there seems to exist a secondary current in the pipe, upward in the middle and downward around the walls.

C. FLOW PATTERNS IN TWO-PHASE FLOW

A number of patterns or regimes can exist in two-phase flow, ranging from mist to bubble flow. These were first described in detail by Alves⁽²¹⁾ and are presented in graphical form in abstract⁽¹⁰⁾.

D. PRESSURE DROPS IN VERTICAL PIPES

Vertical two-phase flow has been studied by a number of investigators since it is of importance in distillation towers and other chemical engineering applications⁽⁴⁴⁾⁽¹⁴⁾⁽²⁸⁾⁽³⁶⁾.

II EVACUATORS

Two basic types of evacuators are presently in use; the ejectors and the vacuum pumps.

A. EJECTORS

The advantages and disadvantages of ejectors are presented in two articles⁽⁸⁾⁽⁶⁾. The main advantage seems to be that they are able to pump "dirty" gases and hence the gases drawn from the suction line do not have to be cleaned. A thorough investigation of ejector performance can be found in the Heat Exchange Institute Standards for Steam Ejectors⁽⁴⁵⁾.

Recently there has become available an automatic relief valve which prevents some "choke off" due to high vacuums. This valve, however, still requires ejectors for pumping of very gaseous liquids⁽²⁵⁾.

B. VACUUM PUMPS

A new design in vacuum pumps is reported⁽³²⁾ which decreases vibrations and hence increases efficiency and capacity.

Some work has been done on vapor-liquid separators⁽³⁸⁾ and might be applied to clean the gas more thoroughly before entering the vacuum pump. Experiments have also been performed on solid-gas two-phase flow⁽²⁰⁾ and they might be used toward the same end.

III ACCUMULATORS

Very little work has been done on the accumulation of gases in pipelines. This problem is mentioned at times in connection with water tunnels⁽³⁹⁾ and drilling operations⁽³⁴⁾⁽²⁶⁾ but these have proven to be of little value.

The most promising area under consideration at the present time is that of vortex type separators. The Nichols Engineering and Research Corporation has developed a number of these separators mainly for use in the paper

manufacturing industry⁽¹⁵⁾⁽⁵⁾⁽⁴²⁾⁽³⁾. The vortex separators now manufactured are designed to remove both grit and gas from the pulp. Another model under study, the "Foamtrap",⁽³⁾ is designed to separate only gas and foam.

This device operates on the principle of centrifugal force. The liquid containing the gas is pumped tangentially at high velocities into a cylinder. The centrifugal force drives the high density materials to the outside and they spiral downward. A gas core forms in the middle and can be evacuated by vacuum pumps.

Another possibility of gas accumulation was suggested by Silberman⁽¹⁸⁾. He proposed using vanes in elbows and forcing the high density materials to the underside, leaving air pockets on the convex side of the vanes.

C O N C L U S I O N

The literature reviewed in this study has been intended to contribute to the design of an acceptable gas removal system.

Abstracts and annotations of the articles reviewed during the report period are included in the appendix. Other articles were abstracted in the Fritz Engineering Laboratory Report No. 310.1.

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Bergelin, O. D. "FLOW OF GAS-LIQUID MIXTURES" Chemical Engineering May 1949, p. 105

The author reviews the progress in the analysis of gas-liquid mixtures. The various flow regimes are named. The Φ - X correlation of Martinelli is reviewed and it is suggested that this is the most reliable and convenient method for figuring the two-phase pressure drop.

The method is presented as follows:

- (1) Find the Reynolds Number for both gas and liquid.

If $R_n > 2000$ for both, the flow is fully turbulent.

- (2) Select the proper coefficients. For turbulent flow, $m=n=0.2$ and

$$C_L = C_g = 0.046$$

- (3) Calculate X using the following equation;

$$X = \sqrt{\frac{(R_n)_G^m}{(R_n)_L^n} \cdot \frac{C_L}{C_G} \cdot \left(\frac{W_L}{W_G}\right)^2 \cdot \frac{\rho_G}{\rho_L}}$$

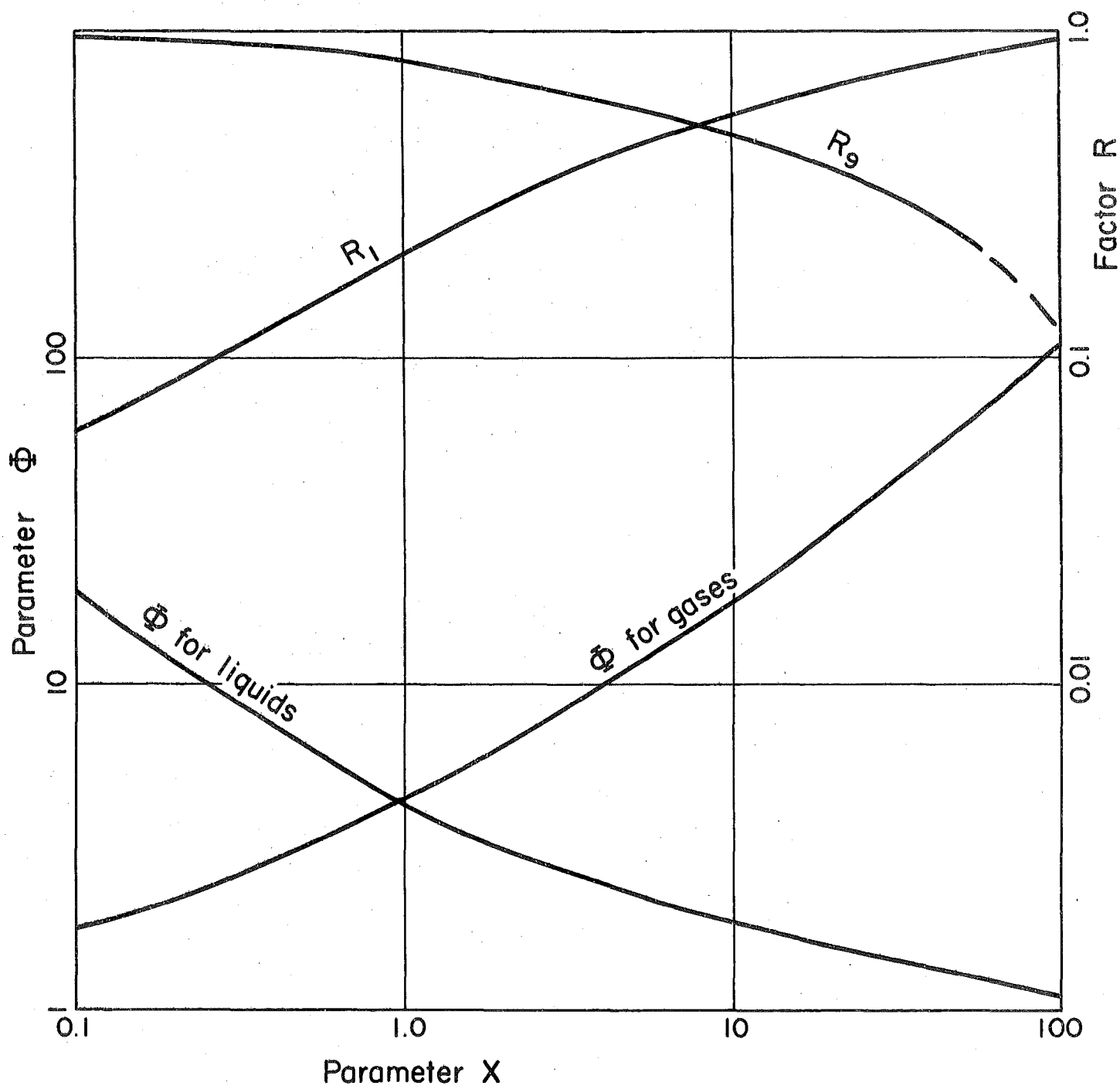
where W = mass rate of flow in lb/sec, and ρ = fluid density in lb/cu. ft.

- (4) From the figure presented on the next pages, find Φ .
- (5) Calculate the two-phase pressure drop using the following equation:

$$\Phi_{G \text{ or } L} = \sqrt{\left(\frac{\Delta P}{\Delta L}\right)_{TP} \div \left(\frac{\Delta P}{\Delta L}\right)_{G \text{ or } L}}$$

- (6) From the calculated value of X find the fraction of tube filled by one phase R.

The author mentions briefly Gas Lifts, Boiler Liquids, and Condensation.



Broadway, J. D. "THEORETICAL CONSIDERATIONS OF VORTEX SEPARATORS"
 Technical Association of the Pulp and Paper Industry 45:4 April 1962

The flow pattern of almost all vortex type separators are two dimensional. Fluid is injected tangentially and spirals downward. As the fluid finally migrates to the center it spirals up and through the central outlet. The basic design parameter is the orifice ratio, a constant used for fan design, and stated as

$$O = \frac{Q}{D^2 \sqrt{\frac{2gP}{\rho}}}$$

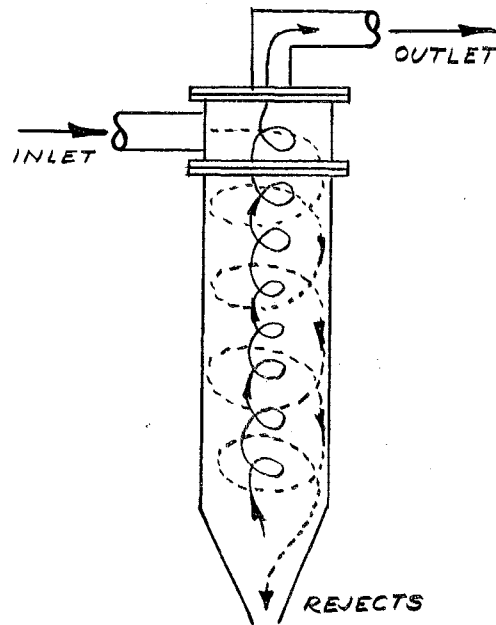
where O = orifice ratio

D = a standard dimension used
 for scaling, usually the
 separator barrel diameter

g = gravitational constant

ρ = fluid density

P = pressure drop across the
 device



Basic Vortex Separator

The parameter O can be varied by changing the ratio of the inlet area to the barrel area, hence changing the amount of pressure energy converted to velocity. It was found that for small values of O , the size of the system had to be increased but the better cleaning performance more than compensated the additional cost.

If frictional losses are considered it is common to express the vortex equation as

$$v = K_g r^n$$

where

v = velocity at any radius

r = radius

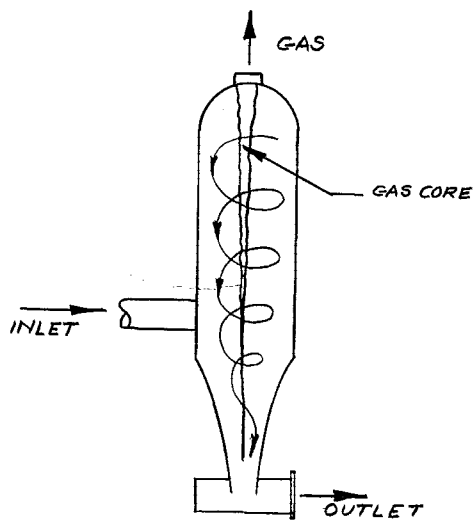
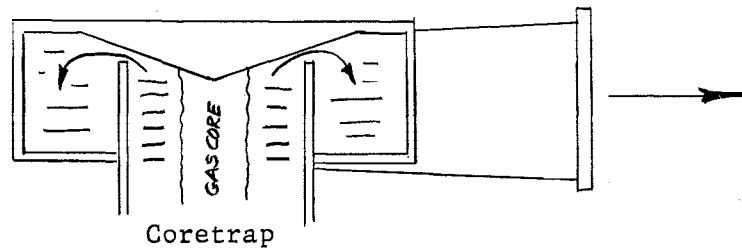
K_g = general vortex constant

n = constant between + 1 and - 1.

If $n = - 1$, a condition of no frictional loss is assumed (free vortex). At $n = + 1$, there is a complete frictional loss of energy, and a wheel type vortex results. The velocities in the vortex are important in determining the pressure drop across it. The pressure drop increases from the wheel type vortex ($n = +1$) to the free vortex ($n = -1$). This is due to the larger fluid velocities at the center as a result of reduced energy losses.

In the center of the vortex, there usually exists a space devoid of liquid due to the high centrifugal force which flings liquid from it. The gas core extends the length of the barrel and is withdrawn at the bottom by a vacuum pump. At the top of the barrel, the fluid is withdrawn and the gas is held in with the use of a "coretrap". The vortex is conveyed through the barrel and brought against the blunt cone where it will escape out the sides while the gas core is held inside. In order that the dissolved gases might be brought into the core, the inlet velocity must be relatively high.

At the conclusion of the paper, the author presents some examples of vortex separators manufactured by his corporation (Nichols Eng. & Res. Corp). Most of these are designed to remove grit and air. One, called the Foamtrap, is designed expressly for the removal of air. It has been patented by the author in collaboration with H. Freeman. In this device, the liquid is inserted at the top and removed at the bottom, resulting in a single vortex. The author states that this device is better for removing gases than the other, but does not give any more details.



Foamtrap

Donoghue, J. J. "AIR-BUBBLE GENERATOR" U. S. Navy Department, David Taylor Model Basin Report R-83 May 1943

Various methods of generating air bubbles in water are reviewed with the objective of finding some method which would produce uniform and reproducible screens of air bubbles. Two were chosen as meeting those requirements, the shear type and the dis-solution type.

The forces that act to remove the growing bubble from its attachment to an orifice determine its time of release, and therefore its size. By letting a jet of water flow past an air orifice the bubble size was controlled. It was noted that as the air flow is increased the size of the bubbles increase as long as the water velocity is zero. As the water flow is increased, with constant air flow, the size of the bubbles decreases and their number increases.

The dis-solution type bubble generator works on a principle of dissolving the air first under high pressure and then releasing it under atmospheric pressure. The physical properties of the liquid can be changed to vary the bubble size.

Three methods that did not meet the stated requirements are the simple orifice, the submerged nozzle and the porous media.

The simple orifice had many drawbacks some of which were; (a) the minimum diameter of the bubble was approximately 10 times the diameter of the orifice, (b) a small variation of relative submergence will produce non-uniform bubbles, (c) very little control over either size or number of bubbles.

The submerged venturi tube or nozzle apparatus was not acceptable because it could not produce a screen of bubbles over a rectangular area.

The porous media method had the same disadvantages as the simple orifice with even less chance of reproducibility.

The factors that determine the size of an air bubble formed in water by forcing air through a permeable surface are:

- (1) the diameter of the orifice
- (2) the rate of flow of gas
- (3) the proximity of other orifices
- (4) the interfacial forces in the liquid-solid boundaries
(electrolytic salt will vary the size of bubbles).
- (5) the viscosity
- (6) the induction time, time of adherence to solid.

Freeman, H. and Broadway, J. D., Consolidated Paper Corp., Ltd.
NEW METHODS FOR THE REMOVAL OF SOLIDS AND GASES FROM LIQUID
SUSPENSIONS WITH PARTICULAR REFERENCE TO PULP STOCK CONDITIONING.
Pulp and Paper Magazine of Canada, Vol. 54, No. 4, pp. 102-107, 1953

Gas in pulp stock exists as bubbles, gas in solution, and adsorbed onto fibers. To remove the bubbles, most of the total gas must be removed. A Vortrap was redesigned to remove both dirt and gas simultaneously; this was accomplished by increasing the fluid velocity within the separating tube so that a vacuum is produced and to which core the gases are displaced by centrifugal force. The gases are then removed by a vacuum pump.

The Vortrap, manufactured by Nichols Research and Engineering Corporation, has been used since 1932 as a classifier in paper mills. In 1938 H. Freeman noted a gaseous core in a glass Vortrap.

Gas in pulp stock has been determined as 3.1%, volumetric basis, with 0.25% being in the form of bubbles. The analysis of the gas is 6.5% CO₂, 25.5% O₂, and 68% unanalyzed.

At one installation a vacuum deaerator was used to remove 70% of all gases present; this included all gas in the form of bubbles. The deaerator consisted of an evacuated tank with baffle plates against which the stock was sprayed. The stock was removed by pumping from the tank, and the gases were removed by an extensive vacuum system.

A general expression for a vortex is $v r^n = k$, in which v is the tangential velocity at radius, r , and n and k are constants. The extremes of n are +1 and -1. The authors assumed n as zero; thus $v = k$, a condition of constant tangential velocity at all radii.

The force acting on the wall of the tube is:

$$F = \int_{R-D}^R \frac{2\pi \gamma L V^2}{g} dr = \frac{2\pi \gamma L D V^2}{g}$$

The force acts on the area of the inside wall and is related to the pressure.

$$F = 2\pi R L p.$$

$$\text{Consequently, } p = \frac{\gamma V^2 D}{g R}, \text{ or}$$

$$p/\gamma = \frac{2D}{R} \frac{V^2}{2g}$$

D = depth of liquid against the wall.

R = radius of Vortrap tube.

If the velocity head exceeds half the pressure head, a vacuum core is produced, the size of which can be predicted.

A Vortrap was redesigned to produce a sizable vacuum core, and the unit could then separate both gas and dirt. The inlet pressure and flow remained constant, but the velocity in the tube was increased and the pressure reduced.

An analysis of dirt separation was made considering particles separating in a viscous medium according to Stokes' law. The results were relatively substantiating, but the discharge of dirt was less owing to the apparent orientation of flat, clay particles.

A photograph is shown in the article of a glass Vortrap, 1-1/2 in. in diameter, with a vacuum core of 1/4 in.

In the Vortrap degasser, stock enters tangentially at the top and spirals down along the wall; most of the dirt, being adjacent to the wall, is removed at the bottom through a tangential outlet. The cleaned stock is turned upward and toward the center. Within the center of the upward rising stock a zone of reduced pressure occurs, and a vacuum develops into which the gases enter. Most of the gases are removed through a bottom center pipe connected to a vacuum pump and spray condenser; the remaining gases exit with the clean stock, are separated in a small vacuum tank and thence to the vacuum pump and condenser.

The performance of a 4-in. degassing Vortrap is as follows:

Pressure	15-35 psi
Flow	175 gpm (US)
Dirt removal, efficiency	50-85%
Percent rejects (bottom dirt)	2%
Gas removal, efficiency	80-90%

Fondrk, V. V. THE STREAM JET EJECTOR: A VERSATILE PUMP FOR HIGH VACUUM
American Vacuum Society Symposium-Transactions 1957

Steam Jet Ejectors have now been developed to the point where they are capable of operating at pressures from well over 50 microns of mercury absolute to well under 10 microns of mercury absolute. They can handle capacities from 0 to 250,000 cfm.

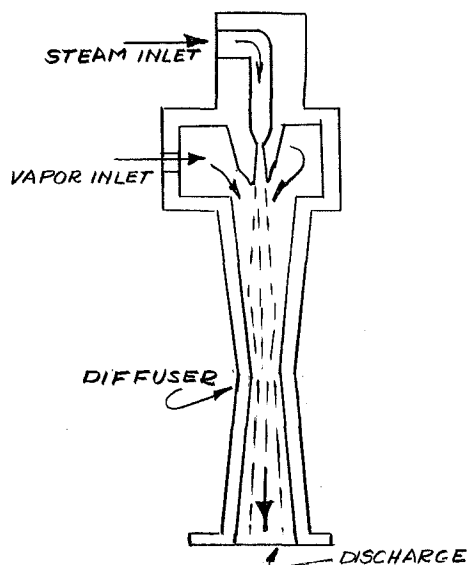


Figure 1 shows a cross section of a typical steam jet ejector. Compressed steam at high pressure is allowed to expand through a nozzle thus converting its potential or pressure energy into kinetic energy. As a consequence, low pressures, or vacuums are created in the suction chamber, thus sucking the unwanted gases away.

Figure 1 shows only a one stage ejector. If very low pressures are required in the suction chamber a typical installation may have four to six stages, each separated by a compressor; or if the unwanted gas is incompressible, by a condenser and compressor both.

Figure 2 shows a typical throughput curve for an ejector used to eject air. Here chamber suction pressure is plotted against capacity in weight of air removed per hour.

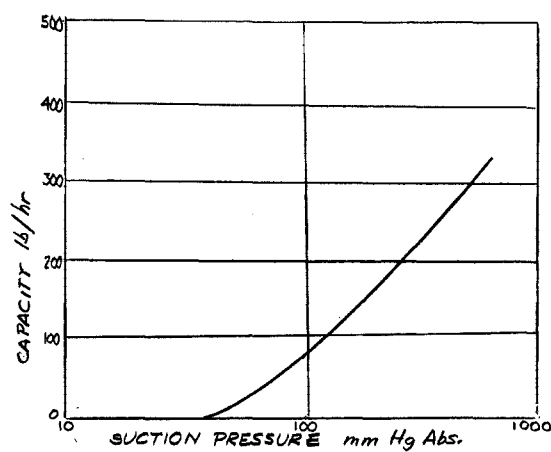


Figure 3 shows the same curve where capacity is plotted in volume of air removed per hour

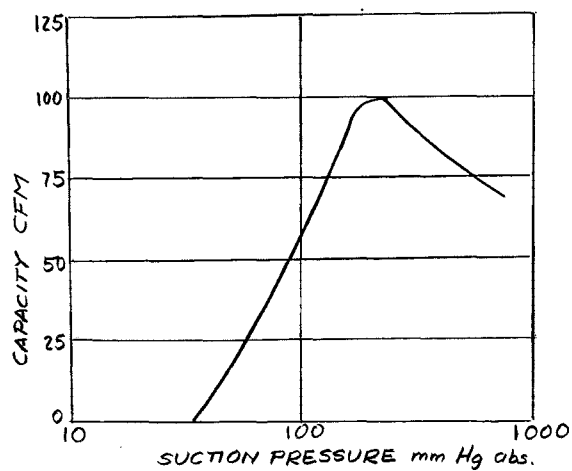


Figure 2 shows that the ejector is capable of pumping increased weight rates of flow as pressure increases; whereas, figure 3 shows that the volumetric capacity reaches a peak point and then drops off as pressure increases. This peak point is usually taken as the design point. The ejector can be designed so that the peak point occurs at any capacity and pressure required. However, in order to prevent condensation and backflow of the expanded steam into the suction chamber when the installation is closed, the shutoff pressure should be as high as possible.

A good point to note in the design of an ejector installation is that the condenser should be located at least 34 feet above the water level in the sump in order to permit drainage by gravity even if full vacuum exists in the suction chamber. This prevents water from draining into the suction chamber (which produces some unpleasant results) even if the power system fails.

One important attribute of the steam jet ejector is its capacity to eject "dirty" gases which contain particles of dirt or liquid carryover. Any particle capable of being carried in the stream of gases being evacuated will carry through the ejector without harm to the equipment and will be washed out in the first condenser. When the load gases (unwanted gases to be ejected) are corrosive, special corrosion resistant materials can be incorporated into the design of the ejector.

Initial costs for a steam jet ejector vary with the size of the ejector. As an example, an ejector with capacity of 10,000 cfm costs from

\$0.75 to \$1.20 per cfm capacity of 200,000 cfm only costs \$0.15 to \$0.25 cfm of capacity. When steam and water are already available, installation costs approximately 40% of the first cost of the ejector.

Houghton, G., McLean, A., and Ritchie, P.

ABSORPTION OF CARBON DIOXIDE IN WATER UNDER PRESSURE USING A GAS BUBBLE COLUMN Chemical Engineering Science Vol. 7, n 1-2, 1957

A gas bubble column is simply a hollow upright tube, or pipe in which water under pressure is made to flow downward from the top of the column while gas, after first diffusing through a perforated plate on the bottom of the column, flows upward. As the bubbles of carbon dioxide flow upward through the water they dissolve and are absorbed by the water. In this manner, absorption rates as high as 90% have been obtained; almost three times the rate of the more usual "packed tower" type of water scrubber.

Any given bubble will rise with a certain terminal velocity of rise in water; and if the water is given a downward velocity equal to this terminal velocity of bubble rise, the bubble will remain stationary with respect to the column. Further increase in liquid velocity will carry the bubble in the same direction as the liquid. Hence, too great a liquid velocity will carry bubbles out of the column without being absorbed. This process is known as carryover. For the porous plates used (having porosity ranging from 72 μ to 1150 μ) it was found that the critical relative velocity between liquid and gas velocities was 0.3 ft. per sec.

It was found that for a constant liquid velocity the upward gas velocity had a great effect on the efficiency of gas absorption. At constant liquid velocity the absorption efficiency increases with rising gas velocity until a constant absorption efficiency is obtained, independent of further increase in gas velocity. For the cases tested it was found that this constant

absorption efficiency is reached at a gas velocity slightly greater than 0.17 fps. The region where the absorption efficiency increases with u_g , (gas velocity) is known as the "rising hold up" region and it was found that the absorption efficiency is proportional to $u_g^{0.61}$; whereas, the region where the efficiency is constant is known as the "constant hold up" region. This phenomena is explained by the fact that the bubble size is constant regardless of gas velocity. Thus, as gas velocity increases, the number, not the size, of bubbles increases, thus increasing the surface area for absorption, and thus increasing the efficiency. However, as the gas velocity increases to a point where $u_g > 0.17$ fps the bubbles change in size to very large, or "plug" bubbles. Thereafter the surface area remains constant, as does the absorption efficiency.

In addition to the above it was found that the absorption efficiency is proportional to $p^{-0.33}$ where p is the partial pressure of the carbon dioxide. As can be expected, absorption rate increases with temperature.

"How Ejectors Are Developing" Ed. Staff, Chemical Engineering May 1949 p. 136

This article attempts to define the limits within which the ejector may prove to be the best all-around vacuum producer.

The advantages of ejectors are:

- (1) Low first cost,
- (2) Simplicity,
- (3) No moving parts,
- (4) Light weight, and
- (5) Easy maintenance.

The disadvantages are:

- (1) Usually a fixed load machine maintaining a fixed suction pressure,
- (2) High water requirements,
- (3) Poor steam economy at pressures below 100 psig,
- (4) If operated at loads lower than design loads, instability and erratic performance results.

Some conditions which favor ejector use over other devices are:

- (1) Availability of steam pressures greater than 100 psig,
- (2) Availability of water,
- (3) The need for suction pressures between 5 inches and 100 microns Hg.

A series of tables is presented showing ejector capacities. If the suction pressure is fixed, and the motive steam pressure is varied, the pound per hour of motive steam required to compress "n" lb/hr. of air can be measured. This measurement for various types of ejectors, with or without intercondensers and/or multiple stages, is presented in these tables. The net cfm displacement required if a mechanical pump was used to produce the same result is presented for comparison. It is seen that for low suction pressures, the ejector has a great advantage over the pump.

It is shown that by increasing the number of stages and installing intercondensers, the efficiency of the ejectors can be increased, but at the cost of higher operating expense.

The article also mentions briefly the use of ejectors as thermo-compressors and boosters.

Isbin, H. S., Moen, R. H., Mosher, D. R., "TWO-PHASE PRESSURE DROP",
Atomic Energy Commission Unpublished Report - 2994 November 1954.

The discussion on two-phase flow presented in this report refers to the simultaneous and co-current flow of gas-liquid mixtures. The purpose of this report is to summarize the current knowledge and limitations of the present (1954) methods of estimating two-phase pressure drops. The following topics are considered: Present status of two-phase calculation methods, flow patterns, flow types, flow models, and flow stability. An extensive literature survey is included.

PRESENT STATUS OF TWO-PHASE PRESSURE DROP CALCULATION METHODS.

In many cases it is possible to show calculated two-phase pressure drop values to be within 30% of experimental values. One such example is the Martinelli method, but this covers the ranges for steady flow only. It is recognized also that total flow rate parameters are not adequately provided for in this correlation. Another deficiency in the Martinelli approach is that frictional losses, momentum changes and head effects are not fully considered. In the annular flow region, for example, gravity effects play an important role, the water layer being thicker on the bottom of the pipe than on the top.

Another method of finding pressure drops is the development of one friction factor (a combination of f_G and f_L) to cover two phase flow. One of the basic assumptions is that the gas and liquid are moving at the

same velocity. This is not always true. No really good methods of "homogeneous flow" types have yet to be advanced.

FLOW PATTERNS.

The mode of flow for each phase of a liquid-gas flow is determined by the shape of the confining conduit, gravitational force, interphase forces and intraphase forces. The interplay of these forces leads to a number of possible cross-sectional and longitudinal profiles for flow. The behavior of the system depends on which of these flow patterns occurs. For horizontal flow, the definitions advanced by Alves are presented. Other authors were able to duplicate these and are fairly well agreed on the definitions, but not on the most advantageous way of presenting results. In vertical flow pipes, the same definitions can be used with minor revisions.

FLOW TYPES.

Flow types are usually designed on the basis of whether laminar or turbulent flow would exist if the phase under consideration were flowing alone in the pipe. Turbulent flow is said to exist at Reynolds Numbers greater than 2000; and at less than 1000, laminar flow is said to exist. Obviously there are then four possible flow types; turbulent-turbulent; turbulent-viscous; viscous-turbulent; and viscous-viscous; describing first the gas phase and the second the liquid phase.

FLOW MODELS

A variety of physical models have been used to define two-phase flow phenomena. Two of the most used are the Martinelli model and the models assuming homogeneous flow.

Martinelli Model

(This model is presented in another abstract of this report but a few notations will be repeated here).

The basic assumptions of the model are (1) the static pressure drop for the gaseous phase is equal to the liquid phase static pressure drop; (2) the volume of the pipe at any instant is equal to the sum of the volume of the gas plus that of the liquid.

As shown in the abstract of the Martinelli article, the factor β was allowed to equal unity. This resulted in the 30% error. It was later recognized that it was better to allow both α and β to vary. In the latest Martinelli paper, the correlation was extended to include condition in which the quality would vary directly with the low rate, as in boiling water. It was discovered that the prediction overestimated the experimental tests. The difficulty was solved when the factor was reduced to the atmospheric pressure conditions. This indicated an unexpected dependency of ϕ to pressure. The method is then applied to steam-water mixtures.

There have been a number of modifications and comparisons of the Martinelli Correlation, among them Levy who gave theoretical support to

Martinelli's choice of parameters. Gazley's analytical development of , however, is about 25% lower than the empirical function. Others have obtained experimental correlations and found Martinelli's values to be acceptable. Among these authors are Bonilla, H. A. Johnson, and VanWinger. Still others have found the method to over estimate the pressure drop, among those are Baker, Alves, Untermeyer, and Schuler.

Friction Factor Models

The basis of the classification of the Friction Factor Models is the use of a single friction factor for the mixed flow. One of the more widely used methods is that of "Homogeneous Flow". The basic premise here is the assumptions of equal gas and liquid velocity and of thermodynamic equilibrium between phases (vapor-liquid equilibrium). Even though the first assumption is seldom fulfilled, many useful results have been derived with it. If this assumption does not hold, the friction factor calculated can be extremely small or even negative.

The friction factor is usually derived by using the energy balance equation, the equation of momentum, and the continuity equation. Many complex relationships have been developed from these basic equation. The experimental data verifies these equations to 50 per cent.

Other types of friction factor models have been attempted. Bergelin and Gazley observed that for both horizontal and vertical flow an increase in the liquid flow results in an increase in pressure drop. This was attributed to the "rough-wall" effect.

Huntington and co-workers developed an expression for two phase friction factor which yielded results of 17 per cent accuracy. Their friction factor was developed empirically.

Mixed Models

Mixed models are characterized by having features based both on the Martinelli and homogeneous models, even though they are contradictory. The usefulness of these models has yet to be established.

FLOW STABILITY

Two-phase flow may become unstable during transition between flow patterns which result in a large pressure fluctuation. The instability is usually associated with the transition from bubbly to stratified flow and the transition from wavy to annular flow.

LITERATURE SURVEY

As stated earlier, an extensive literature survey is included in this report. It includes almost all of the significant papers on two-phase flow up to the year 1954.

Lockhart, R. W., Martinelli, R. C. "PROPOSED CORRELATION OF DATA FOR ISOTHERMAL TWO-PHASE, TWO-COMPONENT FLOW IN PIPES" Chemical Engineering Progress 45:1 p. 39-48, January 1949

The basic postulates upon which is based the analysis of pressure drop resulting from the simultaneous flow of a liquid and gas are:

- (1) Static pressure drop is equal for both the gas and liquid phase,
- (2) The volume of gas plus volume of liquid must equal volume of pipe.

Working with these postulates, the authors have developed a method by which the pressure drop can be predicted for both laminar and turbulent flow. Using the first postulate and the Fanning equation for pressure drops, the authors derived the following relationships: (subscripts l and g refer to liquid and gas respectively; the remainder of the symbols are defined at the end of this abstract).

$$\left[\frac{(\Delta P/\Delta L)_{TP}}{(\Delta P/\Delta L)_l} \right] = \alpha \frac{n-2}{2} \left(\frac{D_p}{D_l} \right)^{\frac{5-n}{2}} = \Phi_l$$

$$\left[\frac{(\Delta P/\Delta L)_{TP}}{(\Delta P/\Delta L)_g} \right] = \beta \frac{m-2}{2} \left(\frac{D_p}{D_g} \right)^{\frac{5-m}{2}} = \Phi_g$$

The application of the second postulate resulted in the following:

$$R_g = 1 - \alpha \left(\frac{D_l}{D_p} \right)^2 \quad R_l = 1 - \beta \left(\frac{D_g}{D_p} \right)^2$$

In all these equations, four variables have appeared, namely,

$$\frac{D_l}{D_p}, \frac{D_g}{D_p}, \alpha \text{ and } \beta$$

which can be expressed in terms of four experimentally determined variables;

$$\Phi_l, \Phi_g, R_l \text{ and } R_g.$$

One more variable is apparent in the derivation, and this can be expressed as

$$\frac{R_{gp}^m}{R_{lp}^n} \frac{C_l}{C_g} \left(\frac{W_l}{W_g} \right)^2 \frac{\rho_g}{\rho_l} = X^2$$

This variable can be shown to equal the ratio of liquid pressure drop to gas pressure drop, assuming each phase flows separately;

$$X^2 = \frac{(\Delta P/\Delta L)_e}{(\Delta P/\Delta L)_g}$$

It is now assumed that the four variables, Φ_g , Φ_e , R_g and R_l are all functions of the parameter X. In this paper, the authors propose to evaluate the correct form of X for all types of flows by the substitution of the appropriate exponents n and m and the constants C_1 and C_g . For fully turbulent flow, ($R_e > 2000$) the following values are given:

$n = m = 0.2$ and $C_1 = C_g = 0.046$. All runs were made with either horizontal pipe or sloping pipes corrected for static head.

Since $\Phi_g = X \Phi_e$, for fully turbulent flow X was found to be equal to

$$\left(\frac{W_l}{W_g}\right)^{1.8} \left(\frac{\rho_g}{\rho_l}\right) \left(\frac{\mu_l}{\mu_g}\right)^{0.2}$$

This relationship is presented in graphical form on page (5a) of this report, and these curves can be used to predict pressure drops in two-phase flow.

In a discussion by Carl Gazley and O. P. Bergelin the curves presented by Lockhart and Martinelli are verified to +20% and -30%.

Notation:

- D = hydraulic diameter, gas or liquid
- D_p = inside diameter of pipe
- W = weight rate of flow, lb/sec.
- ρ = weight density lb/cu. ft.
- μ = absolute viscosity
- P = difference in static pressure in length ΔL
- C = constant in Blasius equation
- m&n = exponents in Blasius equation
- R_e = Reynolds number = $\frac{4W}{\pi D \mu}$
- R = fraction of tube filled by gas or liquid

Subscripts:

- TP = two phase
- l = liquid
- g = gas
- P = pipe

Martinelli, R. C., Boelter, L. M. K., Taylor, T. H. M., Thomse, E. G. and
Morrin, E. H.

"ISOTHERMAL PRESSURE DROP FOR TWO-PHASE COMPONENT FLOW IN A HORIZONTAL
PIPE" Transactions A. S. M. E. February, 1944, p. 139

Tests conducted on 1 inch and 1/2 inch pipes, using air and various liquids such as water, benzine, kerosene, etc. The air was introduced into the pipe line a few feet in front of the test section. It was found that for minimum slugging, the air had to be introduced into the upper section of the pipeline and perpendicular to it.

Pressure drops in the test section were measured and plotted against the air flow rates. Some of the general trends evidenced from these plots are:

- 1 The static pressure drop for two-phase flow is always greater than the pressure drop for each phase flowing alone.
- 2 When air flow approaches zero, the pressure drop due to pure liquid is approached.
- 3 Flow of both air and liquid may be turbulent or laminar.

In this abstract, only the fully turbulent flow conditions are described.

There are two basic postulates on which the analysis of the results was based:

- I. The static pressure drop for the liquid phase must equal the static pressure drop for the gaseous phase.

II. The volume occupied by the liquid plus the volume occupied by the gas at any instant must equal the total volume of the pipe.

These postulates lead to the following equations:

$$\left(\frac{\Delta P}{\Delta L}\right)_{TP} = f_1 \frac{\gamma_1 V_1^2}{D_1 2g} \quad \text{and} \quad \left(\frac{\Delta P}{\Delta L}\right)_{TP} = f_g \frac{\gamma_g V_g^2}{D_g 2g} \quad (1) \text{ and } (2)$$

The subscripts g and l refer to the gas and liquid phase respectively. The hydraulic diameter D_1 and D_g will always be less than the pipe diameter D_p .

For the cross sectional area, the following relationships may be written:

$$A_1 = \alpha \left(\frac{\pi}{4} D_1^2\right) \quad \text{and} \quad A_g = \beta \left(\frac{\pi}{4} D_g^2\right) \quad (3) \text{ and } (4)$$

where α and β are the ratios of the actual cross-sectional area of the flow to the area of a circle of diameter D_1 and D_g respectively. In the following analysis β is assumed to be unity and α will be an unknown. (This approximation could be refined by successive trials assuming both α and β as unknown.)

The friction factor may be written in the general Blasius form as

$$f_1 = \frac{\left(\frac{\pi}{4}\right)^n C_1}{\left(\frac{W_1}{\alpha D_1 \mu_1 g}\right)^n} \quad \text{and} \quad f_g = \frac{\left(\frac{\pi}{4}\right)^m C_g}{\left(\frac{W_g}{D_g \mu_g g}\right)^m} \quad (5) \text{ and } (6)$$

For turbulent flow, $C_1 = C_g = 0.184$ and $m = n = 0.2$ (For laminar flow $C_1 =$

$C_g = 64$ and $m = n = 1$).

Substituting the values for turbulent flow in equations (1) and (2) and with further manipulation, it can be shown that

$$D_g = \frac{D_p}{\sqrt{1 + \alpha^{1.4} \left(\frac{\gamma_g}{\gamma_l}\right)^{0.416} \left(\frac{\mu_l}{\mu_g}\right)^{0.083} \left(\frac{We_l}{We_g}\right)^{0.75}}} \quad (7)$$

and further that

$$\left(\frac{\Delta P}{\Delta L}\right)_{TP} = \left(\frac{\Delta P}{\Delta L}\right)_g \left(\frac{D_p}{D_g}\right)^{4.8} \quad (8)$$

hence

$$\left(\frac{\Delta P}{\Delta L}\right)_{TP} = \left(\frac{\Delta P}{\Delta L}\right)_g \left[1 + \alpha^{1/4} \left(\frac{\mu_l}{\mu_g}\right)^{0.083} \left(\frac{\gamma_g}{\gamma_l}\right)^{0.416} \left(\frac{We_l}{We_g}\right)^{0.75} \right]^{2.4} \quad (9)$$

The preceding equation is the basic equation used in calculating the pressure drop. Only the term α remains to be found.

For simplicity, let

$$X = \left(\frac{\mu_l}{\mu_g}\right)^{0.111} \left(\frac{\gamma_g}{\gamma_l}\right)^{0.555} \left(\frac{We_l}{We_g}\right) \quad (10)$$

Plotting $\alpha^{1/4}$ vs. $X^{3/4}$ yields a smooth line with very little scatter.

Since $\alpha^{1/4}$ is now a function of X , equation (9) reveals that

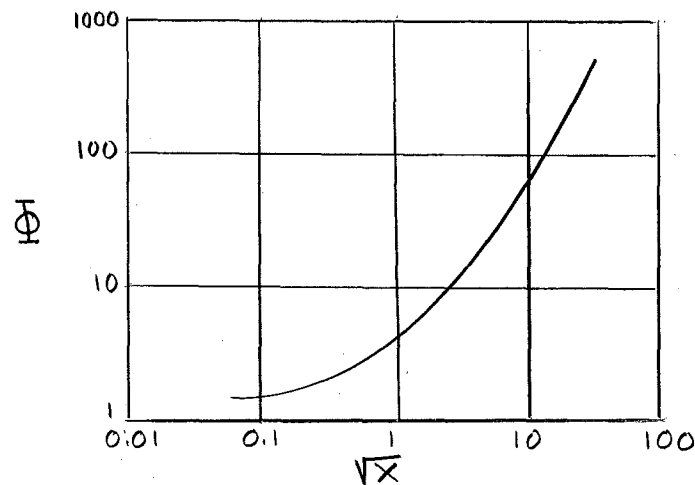
$$\begin{aligned} \left(\frac{\Delta P}{\Delta L}\right)_{TP} &= \left(\frac{\Delta P}{\Delta L}\right)_g \left[1 + \alpha^{1/4} X^{3/4} \right]^{2.4} \\ &= \left(\frac{\Delta P}{\Delta L}\right)_g \left[1 + f(X) (X)^{3/4} \right]^{2.4} \end{aligned} \quad (11)$$

and for each magnitude of X a value of $\alpha^{1/4}$ was established. Then

$$\left(\frac{\Delta P}{\Delta L}\right)_{TP} = \left(\frac{\Delta P}{\Delta L}\right)_g \Phi^2$$

where $\Phi = \left[1 + \alpha^{1/4} X^{3/4} \right]^{1.2} \quad (12)$

The following graph was then plotted showing $\bar{\phi}$ as a function of two-phase flow modulus \sqrt{X} for flow mechanism in which both liquid and gas are in a turbulent motion.



This figure was utilized to predict some pressure drops in the test apparatus and the predicted results were plotted against the actual. This plot showed good agreement.

EXAMPLE OF METHOD

It is desired to estimate the pressure drop in 100 ft. of 2 inch pipe in which air is flowing at a rate of 0.50 lb per sec., and water is flowing at a rate of 6 lb. per sec. The temperature of the fluids is 77° F and the average pressure in the line is 100 psia.

(a) Determination of properties and type of flow:

$$W_g = 0.50 \text{ lb. per sec.}$$

$$\gamma_g = 0.502 \text{ lb. per cu. ft. at 100 psia and } 77^\circ \text{ F}$$

$$\mu_g = 0.388 \times 10^{-5} \text{ lb.-sec. per sq. ft. at 100 psia and } 77^\circ \text{ F}$$

Reynolds modulus for gas considered as a single phase = $R_g = \frac{4}{\pi}$.

$\frac{W_g}{D_p \mu_g} = 295,000$. Likewise, the Reynolds modulus for liquid = 73,500 .

Equation (9) may be utilized due to fully turbulent flow.

(b) Calculation of pressure drop due to air alone: The usual calculation reveals that for $R = 295,000$ the friction factor is 0.018 and the pressure drop is 1.03 psi per 100 ft. of pipe.

(c) Calculation of X:

Using equation (10), it is seen that $X = 1.28$

(d) Calculation of Φ :

$\sqrt{X} = 1.12$ and hence $\Phi = 4.6$ (from the figure)

(e) Calculation of two-phase pressure drop:

$$\left(\frac{\Delta P}{\Delta L}\right)_{TP} = (4.6)^2 (1.03) = \underline{21.8 \text{ psi per 100 ft. of pipe.}}$$

Murdock, J. M. "TWO-PHASE FLOW MEASUREMENT WITH ORIFICES" Transactions ASME
Dec. 1962, Series D, p. 419

This paper presents a practical method for computing two-phase flow rates through standard orifice meters to a tolerance of 1.5%. Numerous data sources were used and various liquid-gas combinations tried. By plotting

$\left(\frac{\Delta P_{TP}}{\Delta P_G}\right)^{1/2}$ vs. $\left(\frac{\Delta P_L}{\Delta P_G}\right)^{1/2}$, a straight line resulted yielding the equation $\left(\frac{P_{TP}}{P_G}\right) = 1.26 \left(\frac{P_L}{P_G}\right) + 1$. Using this equation and the

actual equation of flow through an orifice meter, the following relationship was derived for the weight rate of two phase flow (lb/hour):

$$w_h = \frac{359 K_G Y_G F_a d^2 \sqrt{h_{wTP} G}}{(1-y) + 1.26 y \frac{K_G Y_G}{K_L} \sqrt{\frac{\gamma_G}{\gamma_L}}} \quad (1)$$

This equation applied only when:

- (1) $0.25 < d/D < 0.50$
- (2) Minimum Reynolds number for gas=10,000 and liquid = 50
- (3) Minimum volume ratio of gas to liquid = 100:1
- (4) Maximum liquid weight fraction = 90%

Notation:

Subscript G applies to flow of gas only

Subscript L applies to flow of liquid only

Subscript TP applies to two-phase flow

K = flow coefficient

d = orifice diameter

γ = specific weight

F = factor to account for thermal expansion

y = ratio of weight of liquid, two-phase to total

Y = expansion factor

D = inside pipe diameter

h_w = effective differential head

Nicklin, D. J., Wiles, J. O., and Davidson, J. F. "TWO-PHASE FLOW IN VERTICAL TUBES" Institute of Chemical Engineers-Transaction, Vol. 40 n¹, 1962

A number of flow patterns have been observed when gas and liquid flow together up a vertical tube, and one of the more common of these patterns is the so called "slug flow" which is characterized by very large, round nosed bubbles, or slugs of gas, which move at a velocity greater than the average velocity of the surrounding liquid.

It has been found that such slugs move at a velocity relative to the surrounding liquid which corresponds to that predicted by the Dumitrescu Theory:

$$\mu_o = 0.35 (g D)^{1/2}$$

where μ_o = relative velocity of the rising slug of gas.

g = acceleration of gravity.

D = the tube diameter.

The absolute velocity can be taken as the sum of the Dumitrescu velocity, μ_o , plus a component due to the motion of the liquid. For upward flow of water it was found that this component is 1.2 times the average liquid velocity when the Reynolds number is greater than 8000.

$$\text{thus: } \mu_s = 1.2 \mu_2 + 0.35 (g D)^{1/2}$$

where μ_s = absolute velocity of the gas slug

μ_2 = average velocity of the liquid

Price, R. V. "VORTRAP-VORJECT-VORVAC" Technical Association of the Pulp and Paper Industry 42:12 December 1959

The author discusses three cyclone classifiers manufactured by Nichols Engineering and Research Corporation and attempts to clarify the differences inherent in each piece of equipment and also to specify the possibilities of application.

The first, the Vortrap, is applied to removing sand, grit, and foreign material from all types of pulp and papermaking stock. It is not designed to remove gases. The second, the Vorject, is merely a modification and increases the efficiency but still does not remove air.

The Vorvac, however, is designed to remove grit and dirt as well as air. The liquid is forced through a nozzle type inlet which converts pressure energy into velocity. This nozzle enters tangentially into a cylinder, causing the liquid to follow the wall in a helical path. The dirt is carried close to the wall and is swept to the bottom of the cylinder. The cone turns back the cleaned stock which moves upward and exits at the top. The gases come out of solution as the result of the vacuum produced in the middle core. The headpiece is designed to trap the gas core so that no gas is removed off the top. A vacuum pump draws the gases through the cone.

The Vorvac is available in two sizes, the largest being 24 inches in diameter and a capacity of 1125 g. p. m. By installing units in parallel, any flow requirement can be handled.

Silberman, E. "AIR BUBBLE RESORPTION" St. Anthony Falls Hydraulic Laboratory
 Technical Report No. 1, Series B, 1949.

In the test sections of water tunnels, bubbles are sometimes formed and must be resorbed by the water before it is recirculated since the air content must be constant. An experimental apparatus was built which was able to produce these bubbles under various pressures, temperatures, and velocities and the resorption of the bubbles was measured.

The equation

$$T = \frac{L}{K_1 \beta^D (1 - \alpha \frac{C}{C_G}) R_e^n} R_1$$

was proposed and verified as the basic equation governing gas bubble resorption in turbulent liquid.

In this equation T = time required for complete resorption of gas bubbles (in seconds) of initial radius R_1 (in feet), β = solubility in gas volume per unit volume of liquid at a temperature t and pressure p , α = relative saturation of the liquid ($\frac{C_L}{C_G}$), C = concentration of gas dissolved in liquid ($\frac{\text{lb}}{\text{ft}^3}$), $C_G = C$ if liquid is saturated, C_L = actual value of C in interior of liquid, C_g = difference between C at edge of sublayer and C_L , L = linear dimension measuring scale of the turbulence, taken as the boundary dimension of R_e , R_e = Reynolds number, D = specific

coefficient of diffusion (about 1.95×10^{-8} ft³/sec. for air and water). The constants K_1 and n were evaluated as 0.0162 and 1.25 respectively and $C_g/C_G = 0.18$ at high turbulence.

Mention was made of an idea for removing bubbles in the downstream section of the tunnel. It was suggested that bubbles be bled off by external suction applied at slots on the convex surface of the turning vanes downstream from the diffuser. Preliminary tests indicated that about 5% of the flow would have to be sucked off to withdraw a very large part (but not all) of the bubbles. This idea, however, in the authors words "was left dormant".

Soo, S. L., and Regalbuto, J. A. "CONCENTRATION DISTRIBUTION IN TWO-PHASE PIPE FLOW" Canadian Journal of Chemical Engineering, Vol. 38 n 5, Oct. 1960, pp. 160-166.

Experiments were carried out by the authors in an effort to determine the manner in which the concentration of solid particles varies across the diameter of a pipe filled with flowing gas. Thus, the two phases were the solid particles and the gas. The results obtained are only true for fully turbulent flow and for such a small number of solid particles that gravity effects could be disregarded.

In the experimental approach to the problem, small glass beads were introduced into a 3 inch pipe carrying turbulent flowing air. Fifteen feet downstream the concentration of the solid particles along points across the diameter were measured by means of 0.062" probe which was moved along the diameter, and various electronic particle counters.

Except for the effects of gravity (which can be very large for heavier solid particles or larger amount of particles) the experimental results correspond reasonably to the results of a theoretical investigation; which, after many assumptions and mathematical operations boils down to:

$$C = C_0 J_0 \left(r \sqrt{\frac{f/D}{u}} \right)$$

where

C = conc. of solid particles at any point along the diameter

C_0 = conc. at the center of the pipe

J_0 = the Bessel function of the first kind and of zero order

r = distance from pipe center to the point where C is required

f = acceleration due to transport of particles across the turbulent velocity field of the stream. Thus f relates the fluid drag and inertia of solids and is, in general, a function of state parameters of the system and radial coordinate, r ,

D = the particle diffusivity (due to stream diffusivity)

u = mean velocity of the stream

The important parameter to be derived from the equation on the preceding page is:

$$fR^2/Du \quad \text{where } R = \text{pipe radius}$$

The greater this parameter (e. g. - the greater u becomes) the more uniform the particle concentration.

The authors conclude that for cases where gravitational effects are very small (small particles and light loading) the concentration distribution approximates the turbulent velocity profile across the pipe cross-section. Also, the assumption of a constant concentration distribution appears to be valid for the conditions where gravity effects are small and the mean stream velocity is great.

A P P E N D I X B

A N N O T A T I O N S

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Alves, George E., "CONCURRENT LIQUID-GAS FLOW IN A PIPE-LINE CONTACTOR"
Chemical Engineering Progress, V 50, n9, Sept. 1954, p. 449-456

This report contains the results of an experimental investigation on the isothermal flow of water-air and oil-air mixtures in a 1-in. concurrent pipe-line contactor. The investigation was undertaken to provide basic data on flow pattern, pressure drop, and liquid holdup for the flow of liquid-gas mixtures in this type of equipment. Pressure-drop data for the flow of liquid-gas mixtures have been extended to include the return bends and inlet mixing tee. It is interesting to note that the flow through the pipe is finally separated by a cyclone type of separator.

2b

Begell, W., and Hoopes, J. W., "ACCELERATION OF PRESSURE DROPS IN TWO-PHASE FLOW" Atomic Energy Commission Tech. Inf. Service CU-18-54-At-dP-Ch.E.
April 1954

Two-phase water-steam mixtures are used at high temperatures to produce nomographs for the rapid computation of the parameters used in the Lockhart-Martinelli correlation. The lowest temperature used is over 100° F and therefore the graphs are of little value in connection with the current project, for high project, for high pressure and temperature two-phase flows, however, this paper may be very valuable.

Chenoweth, J. M., and Martin, M. W. "PRESSURE DROP OF GAS-LIQUID MIXTURES IN HORIZONTAL PIPES" The Petroleum Engineer, Vol. 28, n4, April 1956, p. C42-C45.

This investigation was undertaken to check the Lockhart and Martinelli correlation with pressure-drop data for two-phase flow in larger pipes and at higher pressures.

Christolm, D., "THE FLOW OF STEAM/WATER MIXTURES THROUGH SHARP-EDGE ORIFICES" Eng. and Boiler House Review, Vol. 73; p. 253, 1953

Steam-water mixtures were used to find the flow rates for two-component flow through orifices.

The equation

$$\frac{W_{LN}}{W_L} - \frac{1}{C_L} = 2.1 Y^{0.825}$$

where

$$Y = \frac{g}{1-g} \frac{W_{LN}}{W_{VN}}$$

g = dryness fraction

W_L = liquid weight flow rate during two-phase flow

W_{LN} = liquid weight flow rate during single phase flow

W_V = vapor weight flow rate during two-phase flow

W_{VN} = vapor weight flow rate during single phase flow

C_L = liquid contraction coefficient

has been shown to predict flow rates within $\pm 10\%$ of experimental values for flow through sharp edged orifices.

Erwin, R. W., "NEW VACUUM-TYPE DEGASSER REDUCES DEEP DRILLING HAZARDS"
World Oil, Vol. 136, June 1953, p. 125, 127, and 130.

A new cascade-vacuum type degasser with a 25-inch mercury rating, controls and balances mud weight to prevent the shift from lost circulation to blow out conditions, normally a deep drilling hazard. The mud from the well is cascaded over baffles and subjected to high vacuum to draw off entrained gases. The author states that the gaseous fluffiness is eliminated from muds which enables the mud to be handled more easily by pumps.

7b

Fried, Lawrence, "PRESSURE DROP AND HEAT TRANSFER FOR TWO-PHASE, TWO-COMPONENT FLOW" Chemical Engineering Progress Symposium Series No. 9
Vol. 50, 1954, p. 47

When two-phase flow is in the isothermal condition, the Martinelli Φ - X correlation can be used to measure pressure drop. It was found that for non-isothermal flow in which the mixture is heated in the test section, the values of Φ are considerably higher. This is due to the kinetic energy changes. When it was corrected for these changes, the Martinelli correlation was found applicable.

Galegar, W. C., Stovall, W. B., and Huntington, R. L. "REPORT ON TWO-PHASE VERTICAL FLOW" Pipe Line Industry, Vol. 4, n2, Feb. 1956, p. 38-40

This investigation presents experimental data on the performance of kerosene-air and water-air systems in two-phase vertical upward flow using two test sections of different size but having the same ratio of diameter to height. A tentative correlation is presented in conjunction with visual data on the flow patterns involved.

Johnson, H. A., and Abov-Sabe, A. H., "HEAT TRANSFER AND PRESSURE DROP FOR TURBULENT FLOW OF AIR-WATER MIXTURES IN A HORIZONTAL PIPE" Transactions ASME Vol. 74, 997, 1952

The static pressure drop and heat transfer for two-phase, two-component flow of air and water was measured on a 15 foot horizontal pipe. The Martinelli method was used in predicting the results and $\pm 30\%$ accuracy was obtained.

Kalinske, A. A., and Robertson, J. M., "AIR ENTRAINMENT; CLOSED CONDUIT FLOW"
Transactions ASCE 1943 p. 1435

When an air pocket is formed at the summit of a pipeline, the conduit flows only partially full. This air pocket can be removed either by a relief valve or by a hydraulic jump which follows the air pocket. This paper deals with the ability of the jump to carry off the air in the form of bubbles.

A long transparent tube was constructed which was so arranged that the slope could be varied from horizontal to 30 per cent. Entrance and exit structures controlled the flow so that the depth and velocity could be controlled separately. It was found that at some slope and depth, there was a critical water discharge at which the jump could no longer carry all the air being pumped in. This critical condition, for any slope of the pipe and for any relative depth of flow in the air pocket, depends on the value of the Froude number of the flow ahead of the jump. Below the critical value the flow beyond the jump will not be able to handle all the air entrained by the jump and thus the air removed will not be a function of the jump characteristics but rather on the hydraulic features of the flow beyond the jump.

In other words, the rate of air entrainment by a hydraulic jump depends largely on the water discharge and the Froude number of the flow ahead of the jump.

Lorenz, A., "NEW DESIGN OF MECHANICAL VACUUM PUMPS" American Vacuum Society Symposium-Transactions 1958

The most widely used vacuum pumps are oil-sealed rotary piston pumps of the sliding vane type. There are various designs for low and medium capacities. For larger capacities, pumps of the rotary plunger type are generally preferred because their inherently sturdy construction insures reliable operation even under adverse conditions.

Until recently, a serious disadvantage of rotary plunger pumps was the fact that they were frequently unbalanced, and consequently, vibrations have been a problem. Recently, however, a new design has been developed which has completely eliminated this problem. By mounting two different sizes of pistons, along with their eccentrics, and the drive gear whose mass is distributed unsymethically, all on the same drive shaft, the system consisting of the three differently distributed masses can be so arranged and adjusted so that vibration-less operation is obtained in the rotary plunger type.

Models with capacities ranging from 13 to 105 cfm have already been built according to this principle of design.

Levy, S., "THEORY OF PRESSURE DROP AND HEAT TRANSFER FOR ANNULAR STEADY STATE TWO-PHASE TWO-COMPONENT FLOW IN PIPES" Ohio State U. Eng. Exp. Sta. Bulletin #149 Proc. of Second Midwest Conference on Fluid Mech. 1952

The author presents a theoretical approach to the two-phase problem and correlates his results with those of Martinelli, obtaining fair agreement (+ 20%). He also justified the independence of Φ and R from X. (X = two-phase flow modulus, R = fraction of tube filled, % gas or liquid, Φ = dimensionless two-phase pressure drop).

Marks, Robert H., "VACUUM DEGASIFICATION SHOWS ITS VERSATILITY" Power Engineering, Vol. 59, n8, August 1955, p. 92-95

The types of degasifiers described are basically packed columns used to carry out mass transfer between liquid and gas under vacuum conditions. The article included applications of a vacuum degasifier in treatment of: chemical plant cooling water, water for oil field flooding operations, boiler feedwater makeup for a public utility power plant, and cosmotron cooling water.

Martinelli, R. C., Putnam, J. A., and Lockhart, R. W., "TWO-PHASE, TWO-COMPONENT FLOW IN THE VISCOUS REGION", Transactions A. I. Ch. E. Vol. 42, p. 681, 1946

This paper is a supplement to the paper presented by Martinelli et al in the 1944 Transactions of A. S. M. E. in which the ϕ - X correlation was first discussed and experimental data was presented for conditions where the flow was fully turbulent (Reynolds No. > 2000 for both gas and liquid). Experimental data for fully viscous flow is now presented and the ϕ - X correlation is again used.

The angle of the test section was variable and it was found that the pressure drop in fully viscous flow is independent of the angle of the tube. This conclusion, however, "should not be extrapolated to other systems until further experiments are performed."

McAdams, W. H., Woods, W. K., Bryan, R. L., "VAPORIZATION INSIDE HORIZONTAL TUBES" Transactions ASME 1948 p. 545

This paper reports an investigation carried out to determine the changes in the coefficient of heat transfer for the evaporation of a liquid flowing inside a heated horizontal tube. The temperature differences used would be too great for the paper to be of any value in this project. The authors state, however, that in subsequent papers they will describe their method of obtaining the pressure drop, which would be of interest. This paper would then serve to more fully explain the subsequent papers.

Moore, D. W. "THE RISE OF A GAS BUBBLE IN A VISCOUS LIQUID" Journal of Fluid Mechanics, Vol. 6, Part I, 1959

The rise of a gas bubble in viscous liquids and at high Reynolds Numbers is theoretically analyzed. It is shown that the drag coefficient of a spherical bubble is $32/R$ where R is the Reynolds number (based on diameter) of the motion of the rising bubble. Equating the drag force to the bouyant force of the bubble; the bubble diameter and velocity can be computed.

Mathematically extending his analysis to non-spherical bubbles, the author has developed similar expressions for non-spherical bubbles. Unfortunately, the author has not been able to compare these results with direct experimental observations.

Reid, John L., "VAPOR-LIQUID SEPARATORS" Plant Engineering, Vol. 10, n7, July 1956, p. 106-107, 182, 184.

A wire mesh separator is used mainly for obtaining a desirable vapor rather than liquid. The principle is this: As the vapor disengages from the liquid, it carries with it fine liquid droplets. When the vapor stream passes through the fine wire mesh, the liquid droplets impinge on the wire surfaces, coalesce into large drops, and fall off. The vapor is now dry and free from entrained liquid.

Ripken, J. F., "DESIGN STUDIES FOR A CLOSED JET WATER TUNNEL" St. Anthony Falls Hydraulic Laboratory Technical Report No. 9, Series B, 1951

Extensive studies on a model water tunnel are presented. The problem of air entrainment, however, is dismissed by the statement that "no special provision for air content control was made in the basic flow circuit except for three air collection domes to assist in pre-test bubble removal".

Van Wingen, N., "PRESSURE DROP FOR OIL-GAS MIXTURES IN HORIZONTAL FLOW LINES" World Oil October 1949, p. 156

The author adapts the method of Martinelli and co-workers to the problems of transporting oil and gas from a well to a tank farm. Pressure drops are taken for a number of oil lines and this data is compared to the expected results using the Martinelli method. Generally the results are acceptable, even though considerable scatter is present. This scatter, however, can be attributed to the poor control of variables such as temperature, viscosity, etc.

Weiss, D. H. "PRESSURE DROP IN TWO-PHASE FLOW" Argonne National Laboratory-4916, October 1952.

Pressure drop was determined for high pressures and temperatures. The range of pressures was from 20 to 1400 psia and heat fluxes from 100,000 to 500,000 Btu/hr/lb². The Martinelli method of prediction of pressure drops was used and resulted in fairly close agreement. (± 30%).

22b

Wenberg, H. B. "NICOLET'S EXPERIENCE WITH THE VORVAC SYSTEM" Paper Trade Journal Reprint obtained from Nichols Engineering And Research Corp., New York

The author is a manager of a paper manufacturing concern and relates his experiences using the Vorvac system. He claims almost complete absence of air bubbles in the manufactured paper and attributes this to the installation of the Vorvac system.

23b

White, Philip D., and Huntington, R. L., "HORIZONTAL CO-CURRENT TWO-PHASE FLOW OF FLUIDS IN PIPE LINES" The Petroleum Engineer, Vol. 27, August 1955, p. 40-45

This research project was under taken to study the visually observed flow type in general and pressure drops resulting from the stable types of two-phase flow over a wide range of variables. A tentative correlation is presented that shows the type of flow that would exist under a given set of mass flow rates. An empirical correlation is presented, which allows the two-phase pressure drop to be predicted if the flow rates, physical properties, and pipe diameter are known. Also a tentative correlation is presented to predict pressure drops resulting from ripple type flow. The Martinelli method is used and found to be fairly accurate.

Zmola, P. C., Bailey, R. V. Taylor, F. M., Planchet, R. J., "TRANSPORT OF GASES THROUGH LIQUID-GAS MIXTURES" Oak Ridge National Laboratory and Tulane University Publication December 1955.

A vertical column of liquid was used and air was bubbled through the liquid. The flow rate of air and the detention time in the liquid was measured. It was found that even in still liquids the gas would first be in bubble form and as the gas flow is increased, would form slugs. This same phenomenon is also noticed in horizontal tubes.

STANDARDS FOR STEAM JET EJECTORS, Erd. EDITION 1956, HEAT EXCHANGE INSTITUTE
New York

Describes in detail the different types of ejectors along with methods and materials of their construction. Essential design information is given in the way of graphs, data, and examples. Standard methods of operation and basic specifications for performance tests are explained.