Institutionen för vattenbyggnad
Chalmers Tekniska Högskola

Department of Hydraulics
Chalmers University of Technology

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A MODEL FOR CALCULATING THE PUMPING COST OF INDUSTRIAL SLURRIES

Anders Sellgren

Report

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ABSTRACT

A computational model of the operation of centrifugal pumps in slurry service has been developed and used in systematic studies on the influence of different operating conditions on the total pumping costs. The model can, of course, be used for calculating the pumping costs of arbitrary liquids, as well as the flow of air in fans.

The dominant influence of the energy cost in many applications is illustrated by the pumping of mine tailings. In the pumping of ore in autogenous milling operations, the great influence of wear on the pumping cost is discussed.

In a hydraulic hoisting system the grade of size reduction underground is a key parameter in the total economy. The particle size has a great influence on the choice of pumping equipment. The model is used to compare series installation of centrifugal pumps and pumping in one stage with positive displacement pumps of the plunger type. The two pumping systems are compared schematically in an example of hydraulic hoisting of iron ore a vertical distance of about 200 m.

The comparisons discussed in this study, indicate there is an economic potential in slurry transportation by positive displacement pumps of coarser particles than those transported by such pumps today. The fact that a positive displacement pump has a considerably higher efficiency should be weighed against higher initial and maintenance costs.

The total influence on the costs of different maintenance strategies and the reliability of the systems have not been considered in the model. Further development of the model should include these factors. In an Appendix, a new Swedish positive displacement pump is also described, the feasibility of which has recently been demonstrated in a prototype installation at a Swedish mine.

FOREWORD

The present report is part of an ongoing project entitled: "Hydraulic Transportation of Ores and Industrial Minerals from Underground Mines", which is financially supported by the National Swedish Board for Technical Development (STU).

When centrifugal pumps are used, the pumping cost is related to the operating conditions in a complex way, because the performance is affected by the solids in the slurry. The influence of solids was recently investigated as part of the author´s doctoral dissertation at the Department of Hydraulics, Chalmers University of Technology, Göteborg, Sweden (Sellgren 1979). That study provided the necessary background for the work presented here. A paper dealing with the performance of centrifugal pumps in slurry service was presented at the conference "Hydrotransport 6" in England in October 1979. The conference paper is included in this report as an Appendix.

The topic of this report is to describe the model developed and use it in some characteristic applications. The use of the model in a hydraulic hoisting application, which is a central result of this work, was first presented at: "The 5th International Technical Conference on Slurry Transportation 1980" (Sellgren 1980).

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INTRODUCTION

Escalating energy cost has made it necessary to reduce the consumption of energy in all types of pumping operations. The importance of using relevant efficiency concepts in calculating the energy cost of pumping liquids by centrifugal pumps was recently demonstrated by Gustavsson (1979). The efficiency should be related to the consumption of energy per year for the whole unit including not only pump, motor, and control equipment but also variations of the flow with time. In order to achieve an energy-efficient operation one must know the operating conditions in detail, because the efficiency of a centrifugal pump is highly dependent on these conditions.

In slurry handling, the situation is further complicated because the centrifugal pump characteristics and the wear are affected by the solids in the slurry. Alternative types of pumps have been considered in an attempt to achieve a more energy efficient operation.

In short-distance, in-plant slurry transportation systems, the pressure required normally is below 0.5 MPa. In these applications centrifugal pumps usually are the only economical alternative. In high-pressure, long-distance slurry pipeline transportation systems, positive displacement pumps operating at pressures of 8-16 MPa are used. Pumping with positive displacement pumps limits the maximum particle size to about 2 mm due to narrow valve passages, but no such limitations exist with centrifugal pumps. Recently, pumping at working pressures of over 5 MPa with centrifugal pumps has been reported, and, therefore, an interesting intermediate area exists, in which both types of pumps are technically feasible.

In hydraulic hoisting applications the working pressures generally are moderate to high. As the flow rate normally is below, say, 0.1 m^3/s, positive displacement pumps are also an economically interesting alternative.

OBJECTIVES AND SCOPE OF THIS STUDY

The objective was to develop a computational model for the determination of the most economical hydraulic design of a system based on centrifugal slurry pumps. A technical and economic comparison of hydraulic hoisting in systems based on series installation of centrifugal pumps and positive displacement pumps, respectively, was then to be made.

Positive displacement pumps in slurry service normally work with a constant efficiency independent of the operating conditions. Therefore, it is relatively simple to determine the cost of energy. However, in this study, it was not possible to obtain reliable information about the cost of maintenance when pumping coarser particles in positive displacement pumps.

When a large number of centrifugal pumps are required the reliability is adversely affected. The total effect of different maintenance strategies and the reliability of the system are not considered in the model. However, a further development of the model could include these factors in a systematic way.

PERFORMANCE OF CENTRIFUGAL SLURRY PUMPS

Clear-water head and efficiency of centrifugal pumps are generally lowered by the presence of solids, especially if the solids are coarse and heavy. The relative reduction of the clear-water head and efficiency at a constant flow rate and rotary speed may be defined by the following two ratios:

the head ratio: H/H_o

the efficiency ratio: η/η_o

where H = head developed in slurry service, meters of slurry

H_o = head developed in water service, meters of water

η = pump efficiency in slurry service

η_o = pump efficiency in water service

The influence of solids on the performance of a centrifugal pump has recently been investigated (see the Appendix A). The clear-water head and efficiency were found to be lowered about 30% by the iron ore shown in Figure 1 when pumped at a concentration by weight of 50%.

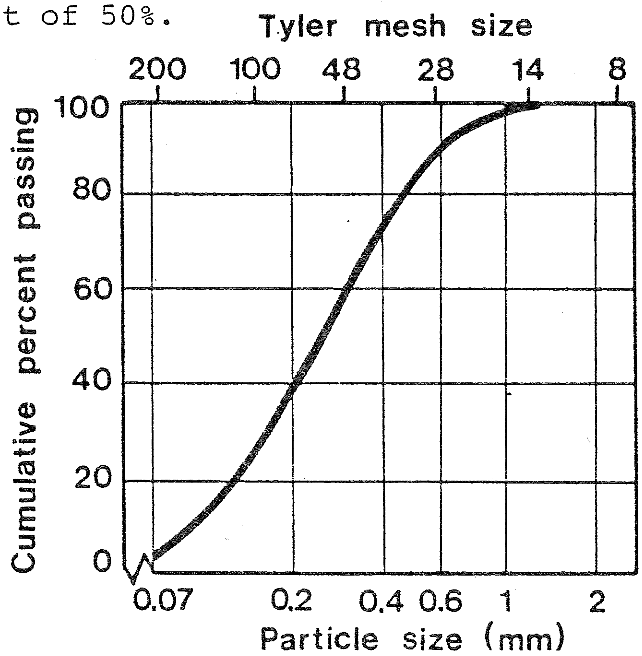

Figure 1. Particle size distribution of an iron ore (density = 4003 kg/m^3) used in the experimental study reported in the Appendix A.

The relative reduction, H/H_o was related to a function of the drag coefficient of the particles, the relative density of solids, and the delivered concentration of particles. The drag coefficient was calculated from weighted values of particle size and particle settling velocity (see the Appendix A, page A-10).

The drop in efficiency, η/η_o, was equivalent to the drop in head up to concentrations by volume of 20-25%. With higher concentrations, the drop in efficiency became greater than the drop in head. For these high concentrations the drop in efficiency was limited by the following relationship

$$1 - C_w \leqslant \eta/\eta_o \leqslant H/H_o \tag{1}$$

where C_w is the solids concentration by weight.

The difference in the drop in head and efficiency for higher concentrations is discussed in more detail in the Appendix A and in the dissertation (Sellgren 1979). This behavior has not been confirmed by other investigators; however, few published results exist of the effect of highly concentrated slurries on the pump performance. The larger drop in efficiency found here may be related to the individual pump size and geometry and to the domain of operation.

DESCRIPTION OF THE MODEL

Modeling of hydraulic data

The clear-water characteristics of centrifugal slurry pumps are normally presented in graphical form on charts delivered by the manufacturer. Typical curves are shown in Figure 2.

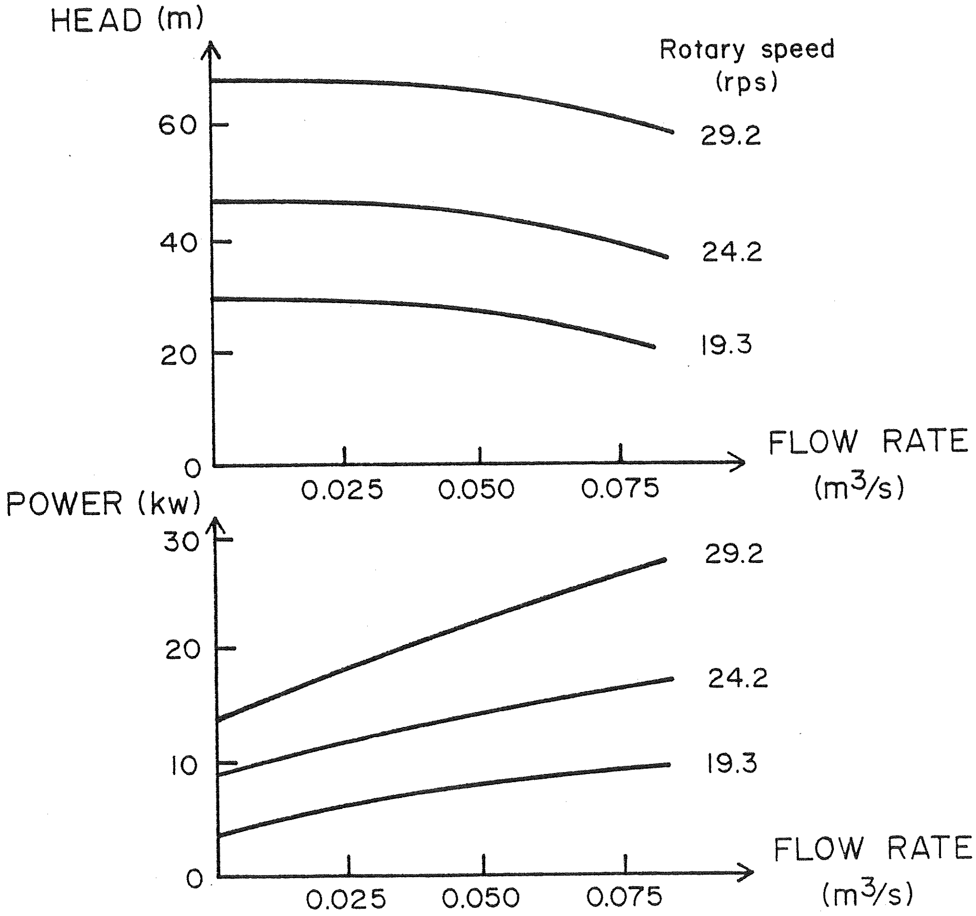

Figure 2. Typical performance curves of a centrifugal slurry pump.

The efficiency η_o of a pump is related to the head H_o, the power requirement P_o, the flow rate Q_o, and the water density ρ_o, by the following relationship:

$$\eta_o = \frac{P_o}{\rho_o g\ H_o\ Q_o} \qquad (2)$$

The clear-water head and efficiency relationships can thus be expressed by the following general equations:

$$H_o = \emptyset\ (Q_o,\ N) \tag{3}$$

$$\eta_o = \emptyset\ (Q_o,\ N) \tag{4}$$

where $\emptyset$ denotes "a function of" and N is the pump rotary speed.

A numerical procedure was developed in order to obtain the head and efficiency for an arbitrary set of flow rates and rotary speeds in Eqs. (3) and (4). The head and power requirement curves of a centrifugal slurry pump are approximately linear for large intervals of the flow rate. It was also found that the head and power requirement varied linearly with the rotary speed for a constant flow rate within relatively large intervals of speed.

Based on the pump performance curves, the clear-water head and power relationships were therefore linearized in steps (Figure 3). These relationships were then transformed into Eqs. (3) and (4) and stored in a computer memory for every type of pump considered.

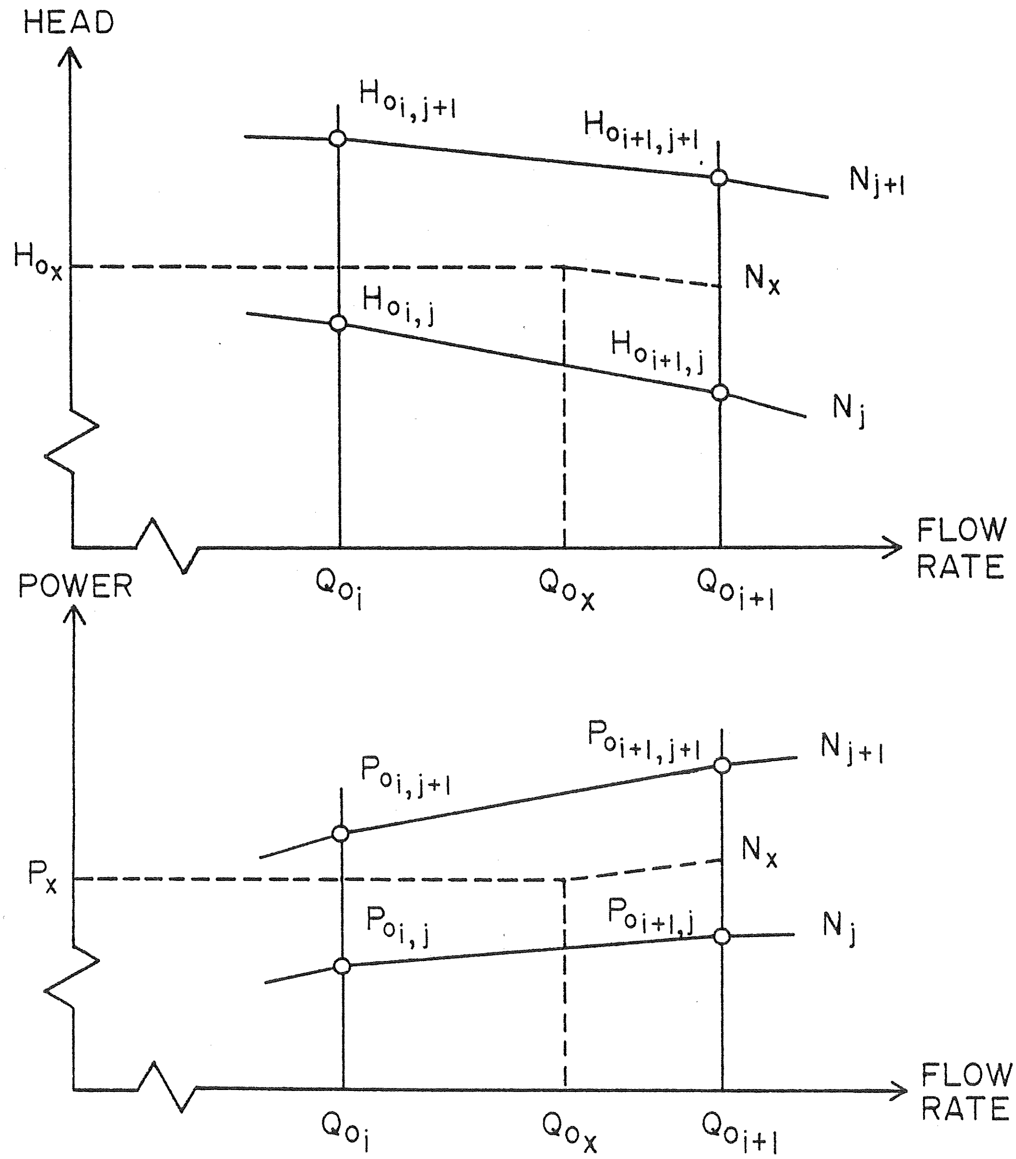

Figure 3. Linearization in steps of the clear-water head and power relationships.

The discret values $H_{o_{i,j}}$, $H_{o_{i+1,j}}$, $H_{o_{i,j+1}}$, $H_{o_{i+1,j+1}}$, $P_{o_{i,j}}$, $P_{o_{i+1,j}}$, $P_{o_{i,j+1}}$, and $P_{o_{i+1,j+1}}$ are input values.

The head and power requirements within an interval Q_i, Q_{i+1} were obtained for an arbitrary flow rate Q_x and speed N_x by the following interpolation relationships:

$$H_{o_x} = (H_{o_{i,j}} - H_{o_{i,j}})\ X - \left[\frac{H_{o_{i,j}} - H_{o_{i+1,j}}}{Q_{o_{i+1}} - Q_{o_i}}\right] Q_{o_x} - \left[\frac{H_{o_{i,j+1}} - H_{o_{i+1,j+1}}}{Q_{o_{i+1}} - Q_{o_i}} - \frac{H_{o_{i,j}} - H_{o_{i+1,j}}}{Q_{o_{i+1}} - Q_{o_i}}\right] X\ Q_{o_x} \tag{5}$$

where $X = \dfrac{N_{j+1} - N_x}{N_{j+1} - N_j}$

$$P_{o_x} = P_{o_{i,j}} + (P_{o_{i,j+1}} - P_{o_{i,j}})\ X + \left[\frac{P_{o_{i+1,j}} - P_{o_{i,j}}}{Q_{o_{i+1}} - Q_{o_i}}\right] Q_{o_x} + \left[\frac{P_{o_{i+1,j+1}} - P_{o_{i,j+1}}}{Q_{o_{i+1}} - Q_{o_i}} - \frac{P_{o_{i+1,j}} - P_{o_{i,j}}}{Q_{o_i} - Q_{o_i}}\right] X\ Q_{o_x} \tag{6}$$

The operating domain of interest and the numerical accuracy determined the spacing between the discrete values, i.e., the maximum values of the indices i and j denoted I and J, respectively. In practice, system considerations limit the operational domain of interest, and it was often sufficient to choose one interval of flow rates (Figure 4).

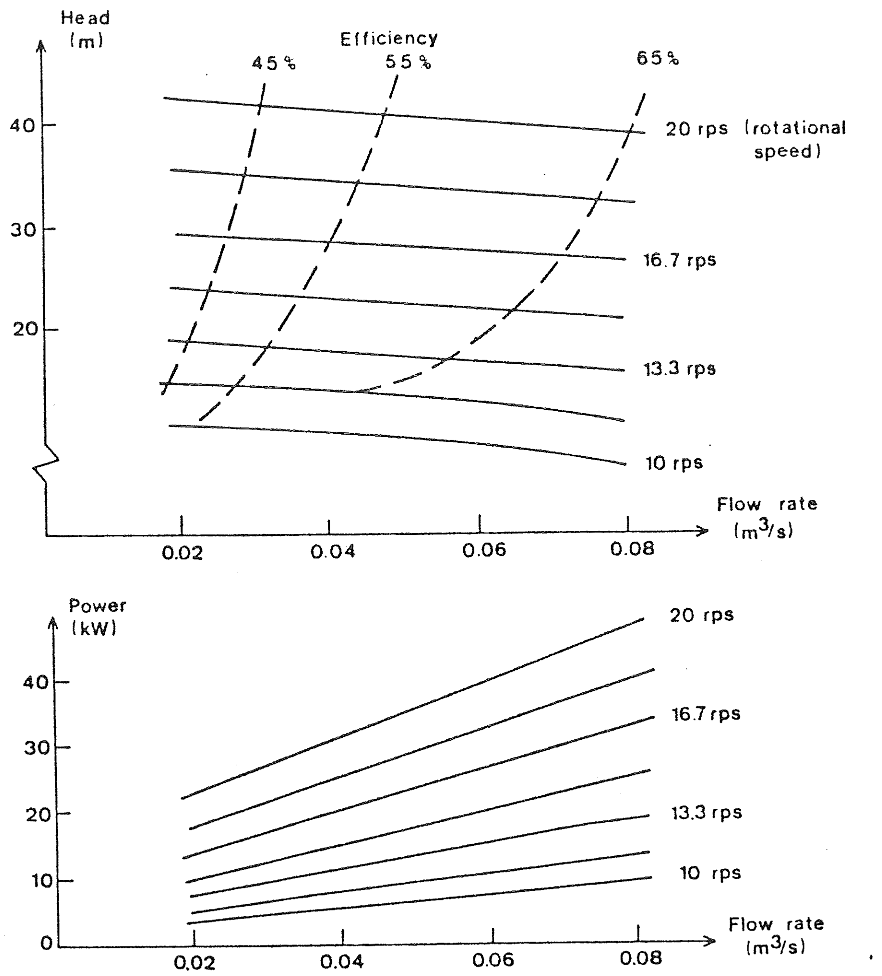

Figure 4. Clear-water head, power, and efficiency curves of the centrifugal pump used in the experimental study reported in the Appendix A. The head and power relationships were linear in a large flow rate interval. The unit was a 0.15 m by 0.15 m hydroseal rubber-lined pump with a 3-vane impeller, 0.43 m in diameter.

For a specific solid, the drag coefficient and the relative density of solids were constants, and the drop in head can be expressed by the following relationship (for details see Eqs. (19) and (23) in the Appendix A, page A-10).

$$H/H_o = 1 - c_1 \, C_w^{\,c_2} \tag{7}$$

where c_1 and c_2 are constants and C_w is the concentration by weight. The head and efficiency in slurry handling were then obtained from the general Eqs. (3), (4), and (7), i.e.,

$$H = \emptyset \, (Q, N, C_w) \tag{8}$$

$$\eta = \phi\ (Q,\ N,\ C_w) \tag{9}$$

where Q is the flow rate of the slurry. Numerically, Eq.(7) was used together with Eqs. (5) and (6).

The flow rate of the slurry was related to the mass flow rate of solids Q_s, the solids density ρ_s, and the concentration by volume C, by the relationship:

$$Q = \frac{Q_s}{\rho_s\ C} \tag{10}$$

where C was expressed in terms of C_w and the relative density of solids, $s_s = \rho_s/\rho_o$

$$C = \frac{C_w}{s_s - C_w\ (s_s-1)} \tag{11}$$

In the model the head and efficiency were normally calculated based on,

$$H/H_o = \eta/\eta_o$$

in Eq. (1), which holds true if the solids concentration is not too high. For example the relative reduction of the clear-water head and efficiency was approximately equal up to concentrations by weight of 60% for the iron ore shown in Figure 1. For very high concentrations, the model also calculates operating and cost data based on the other limit introduced in Eq. (1),

$$\eta/\eta_o = 1 - C_w$$

Modeling of the pumping costs

In the computer model, the total pumping cost per tonne of solids is computed for every set of hydraulic data generated. The costs of capital, energy, and maintenance are based on the following assumptions.

<u>Investments</u> (pumps, motors, and auxiliary equipment): In series installation, one unit is equipped with a variable-speed drive.

Annual costs are based on 6000h of operation per year. Furthermore, a life of 10 years and an interest of 16% are assumed.

<u>Energy</u>: The cost of energy is based on 0.15 Sw.Cr.[1] per kWh and 6000h of operation a year.

<u>Maintenance</u>: Background data of maintenance cost were obtained from industrial experience, mainly in Swedish ore-dressing plants. In typical in-plant service, the cost of maintenance (material and labor) ranged from 0.01 - 0.05 Sw.Cr. per pump and tonne. Furthermore, the final operating speed selected normally could be related to an average pump head for the pump and flow rates considered.

Based on this experience, the cost of wear was estimated by a simplified approach in the model. The speed corresponding to an average pump head in the operational domain considered, referred to as a reference speed ($N_{Ref.}$), was related to the year-round cost of maintenance. In the model, wear is assumed to change by the third power of pump speed. The cost of wear for an arbitrary speed N, is then estimated with the following relationship,

$$\left[\text{Cost of wear per pump}\right] = \left[\text{Cost of wear per pump at } N_{Ref.}\right] \left[\frac{N}{N_{Ref.}}\right]^3$$

1) One Sw.Cr. was about 0.22 US$ in January 1980.

Input and output

The input data to the model were given in the following form,

Pump data

Pump type

Input flow rates: Q_{o_1} ..... Q_{o_i} ..... $Q_{o_{I+1}}$

" rotary speeds: N_1 ..... N_j ..... N_J

" pump heads: $H_{o_{1,1}}$ ..... $H_{o_{i,j}}$ ..... $H_{o_{I+1,J+1}}$

" values of power: $P_{o_{1,1}}$ ..... $P_{o_{i,j}}$ ..... $P_{o_{I+1,J+1}}$

System data

Total head (m)

Maximum number of pumps

Solids throughput (tonnes/h)

Cost data

Pump costs (Sw.Cr.)

Motor costs (Sw.Cr./kW)

Variable speed drive (Sw.Cr./kW)

Time of operation per year (h)

Economic lifetime (years)

Interest (%)

Energy cost (Sw.Cr./kWh)

Reference speed (rps)

Cost of wear at reference speed (Sw.Cr./tonne)

Solids data

Mineral type

Solids density (Kg/m^3)

Weighted drag coefficient, $\bar{C}_D$ (for calculation of $\bar{C}_D$ see Eq. (20) in the Appendix A, page A8)

Solids concentrations by weight (%)

The output was given in the form of tables of operating data and estimated costs:

Operating data

Number of pumps:	-----------------
Pump rotary speed:	-----------------
Pump head/unit:	-----------------
Pump efficiency:	-----------------
Power requirement/unit:	-----------------

Cost data

(all costs are given in Sw.Cr./tonne)

Capital cost:	-----------------
Energy cost:	-----------------
Maintenance cost:	-----------------
Total pumping cost:	-----------------

Informative input data were tabulated together with the operating and cost data. It was also possible to obtain a graphical representation of the pumping costs versus the concentration by weight for a selected number of pumps. The computer programs used in the model are now on file with the Department of Water Resources Engineering at the University of Luleå, Sweden.

DISCUSSION OF RESULTS

With given constant input values of total head, mass flow rate, and selected solids concentrations, the model generates hydraulic design data. Sets of rotational speeds and number of pumps are tabulated for each solids concentration (flow rate) considered. For example, the choice of operational variables shown in Table 1 is theoretically possible when pumping the iron ore shown in Figure 1 with a pump having the clear-water characteristics shown in Figure 2.

Table 1. Pumping of 150 tonnes/h of iron ore by centrifugal pumps a total head of 110 m. Sets of pumps and rotary speeds possible for different solids concentrations.

Concentration by weight (%)	Flow rate (m^3/s)	Rotary speed (rps)		
		5 pumps	6 pumps	7 pumps
40	0.073	17.50	16.12	15.13
45	0.061	17.60	16.18	15.18
50	0.052	17.75	16.28	15.25
55	0.045	17.92	16.42	15.35
60	0.038	18.12	16.58	15.48

The operating speed in practical applications cannot be so accurately established as shown in Table 1 because the performance is affected by the successive wear inside the pump and by changes in slurry characteristics. However, the economy of the operation is very sensitive to the choice of pump speed. Wear increases by approximately the third power of the pump speed, an assumption which is generally accepted. Therefore, even a small deviation over a long period of time from a predetermined average pump speed may be very expensive.

From Table 1 can be seen that the pump speed is influenced only slightly by increased concentrations. If the solids concentration is increased, then the pump speed must be increased in order to maintain a constant head and flow rate. For the data considered in Table 1, the rate of dry solids is constant, and therefore the flow rate of slurry decreases for increasing concentrations. The increased pump head for decreasing flow rates at a constant speed is nearly compensating for the extra speed required in order to maintain a given head at an increase in concentration (Figure 5).

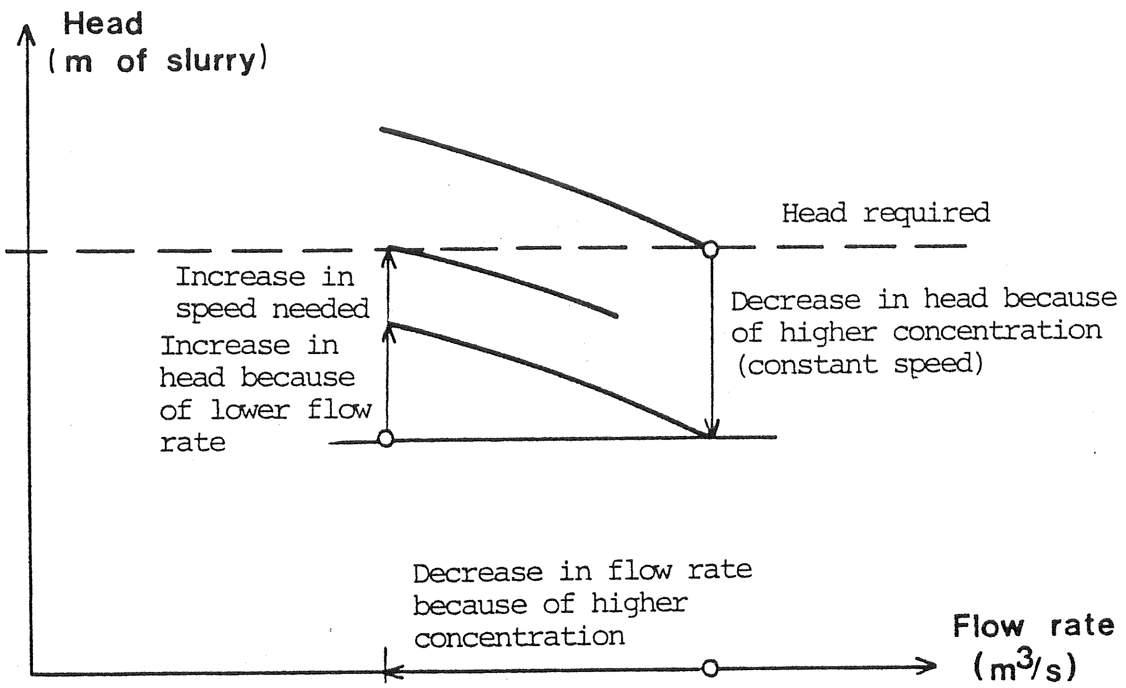

Figure 5. Schematic illustration of the influence of the solids concentration and flow rate on the pump rotary speed.

Based on actual costs of energy and investments and the simplified approach to the cost of wear, the total costs of pumping were computed in the model and tabulated together with operational data such as flow rate, concentration by weight, number of pumps, and pump speed.

For example, the total pumping costs for the set of operational data given in Table 1 are shown in Figure 6 as a function of the concentration by weight. For the iron ore and pump studied here, the cost of wear at a speed of 16.6 rps could be related to a maintenance cost of approximately 0.025 Sw.Cr./tonne and pump.

If the cost of maintenance is not included, the curves of the cost of capital and energy have configurations similar to the curves shown in Figure 6. The output pump power required to deliver a constant rate of solids decreases for increasing values of the concentration by weight. However, for increasing

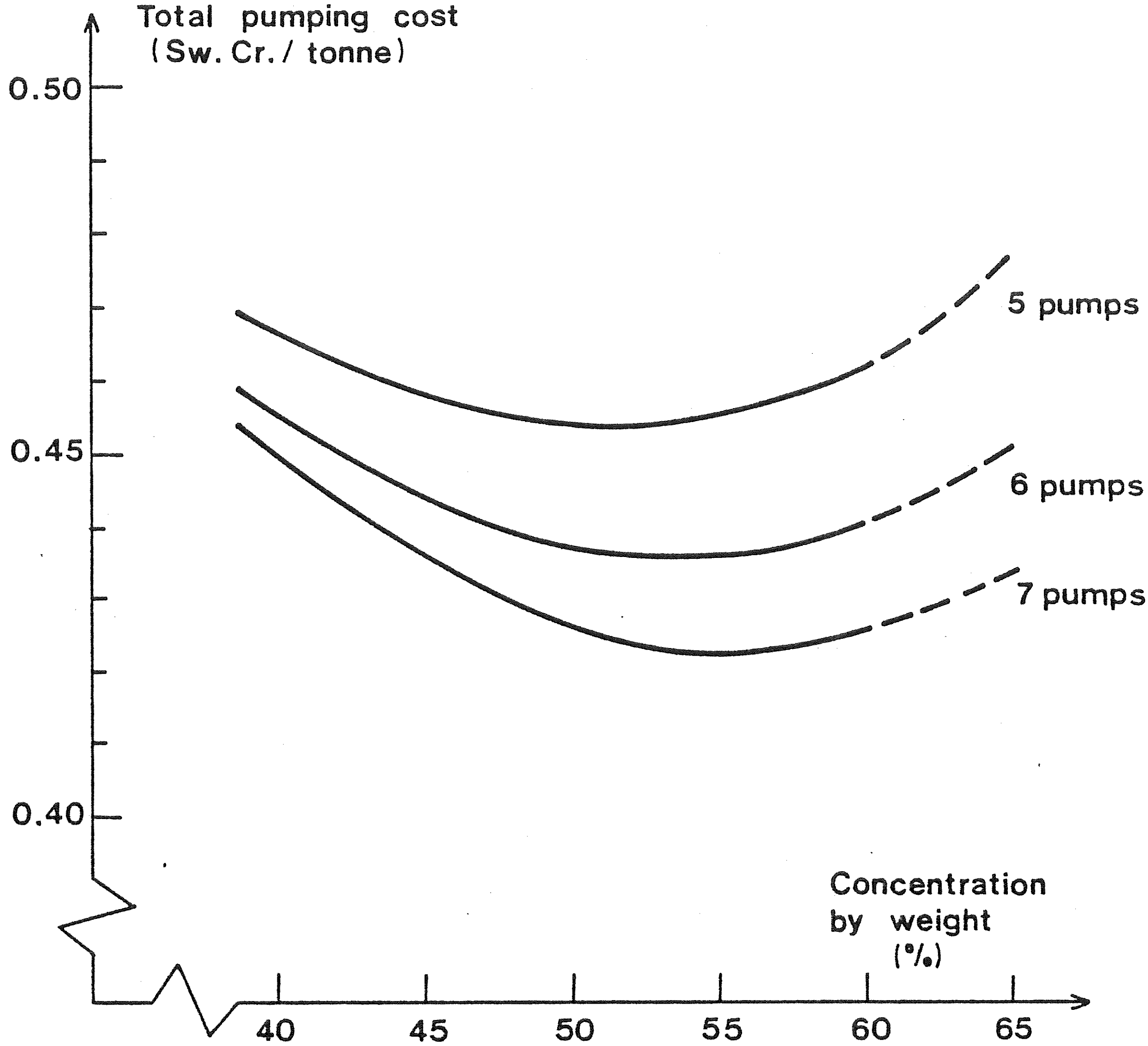

Figure 6. Total pumping cost for different numbers of pumps and concentrations by weight. Pumping of 150 tonnes/h of iron ore a total head of 110 m.

values of the concentration, the pump operates at a lower flow rate and efficiency. Therefore, for most centrifugal slurry pumps, there exists a minimum of the total cost of capital and energy with respect to the concentration by weight.

<u>Comparison between systems consisting of centrifugal pumps in series and a positive displacement pump, respectively.</u>

The two pumping systems were compared schematically in an example of hydraulic hoisting of an iron ore a vertical distance of about 200 m. Based on an operational rotary speed in the range of 16 to 17 rps in the centrifugal pumps, two series of six pumps each, situated on two levels were considered (Figure 7).

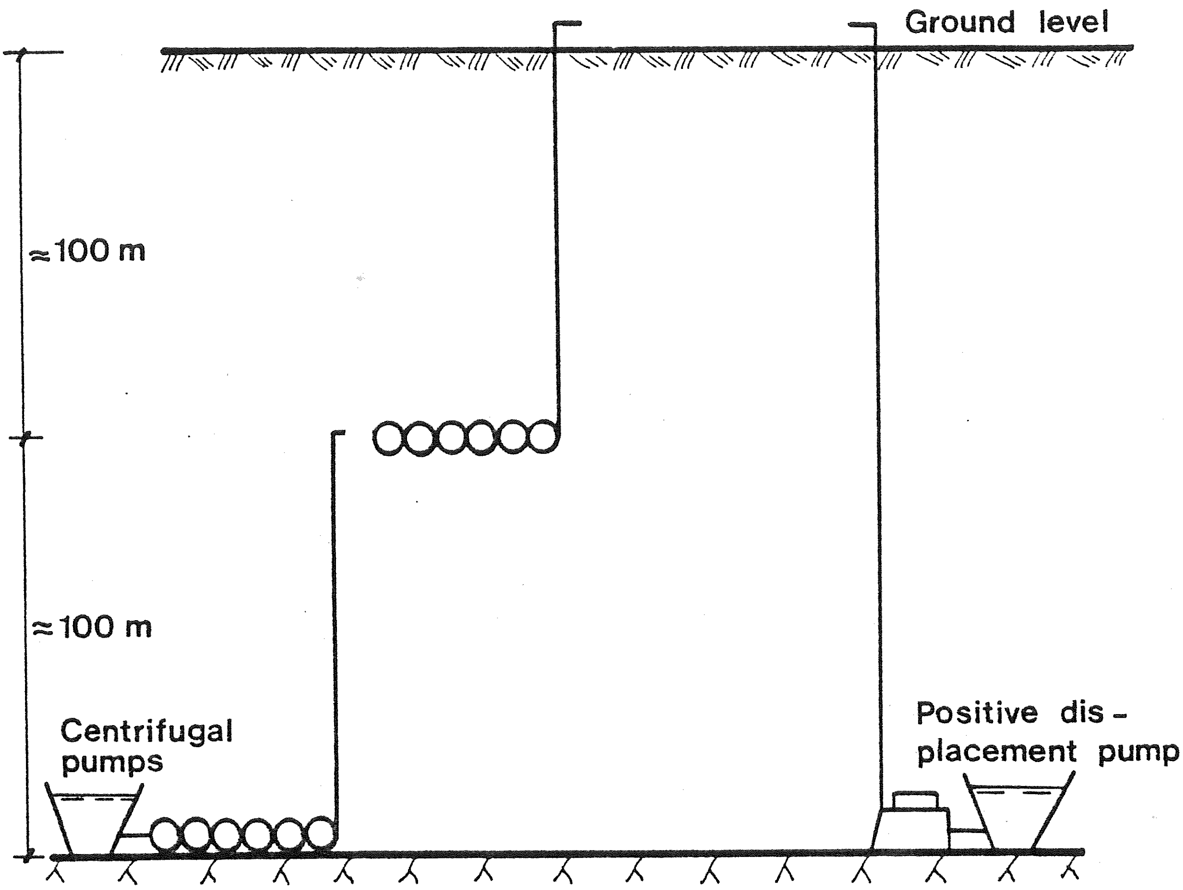

Figure 7. Vertical pumping of iron ore slurry. Comparison between centrifugal pumps in series and a positive displacement pump of the plunger type.

The positive displacement pump considered here was assumed to operate at a total efficiency of 80%.

The operational data are summarized below

Annual production = 0.9 Mtonnes
Mass flow rate = 41.7 kg/s (150 tonnes/h)
Density of iron ore = 4003 kg/m^3
Max. particle size = 1.5 mm (see Figure 1)
Concentration by weight: 50%
Mixture flow rate delivered = 0.052 m^3/s
Total pressure = 3.5 MPa

A straightforward estimation of the annual cost per tonne of pumping the ore is shown in Table 2.

Table 2. Estimation of the annual cost per tonne of pumping the ore

	Annual cost/tonne (Sw.Cr.)	
	Capital[a]	Energy[b]
System based on centrifugal pumps in series	0.12	0.46
System based on a positive displacement pump	0.28	0.23

[a]Economic lifetime = 10 years; interest = 16%

[b]0.15 Sw.Cr./kWh

The predicted costs of capital and energy reflect the fact that a positive displacement pump has a considerably higher efficiency than a centrifugal pump in the application studied here.

Based on experience of iron ore slurry handling, the total maintenance costs in the centrifugal pump system would be in the range of 0.25 to 0.35 Sw.Cr. per tonne. With positive displacement pumps, the lifetime of valve parts in abrasive long-distance iron ore service (max. particle size about 0.1 mm) will be about 500 hours according to Thompson et al. (1972). The same authors stated typical ranges of the cost of expendable parts, which today correspond to 0.30 - 0.60 Sw.Cr. per tonne and pump station. The price of pumping coarse solids is accelerated wear. In the example outlined here, ore particles of up to 1.5 mm would be pumped. At present, the cost of maintenance and supply of positive displacement pumps in such applications is not known by the author.

The number of pumps affects the reliability of the system, and additional costs of extra space needed etc. should also be considered. The costs of these factors are not included in the model, but they all work against the use of a large number of units.

A modified positive displacement pump has been developed and tested in Sweden. The pump is shortly described in the Appendix B.

Pumping of tailings from an ore dressing plant

The total pumping costs were compared for three types of commonly used centrifugal slurry pumps. The pump types were denoted A, B and C, respectively.

Assumed operational data:

Total head	=	70 m
Mass flow rate	=	60 tonnes/h
Mixture flow rate	=	0.058 m^3/s
Density of solids	=	2700 kg/m^3
Concentration by weight	=	24%
Maximum particle size	=	0.2 mm

The relation of the cost of capital, energy, and maintenance is shown in Figure 8.

The cost of maintenance is normally very small because of the small size of the pumped particles. The difference in investment costs of the pumps compared in Figure 8, had a nearly negligible effect on the total cost. It can clearly be seen from Figure 8 that the cost of energy dominates the total pumping cost. Pump type A had the best efficiency of the pumps compared and thus also the lowest total pumping cost.

Pumping of ore in an autogenous milling operation

The great influence of the maintenance cost in operations like this one is demonstrated.

Assumed operational data:

Total head	=	30 m
Mass flow rate	=	400 tonnes/h
Mixture flow rate	=	0.15 m^3/s
Density of solids	=	2800 kg/m^3
Concentration by weight	=	50%
Maximum particle size	=	ca. 5 mm

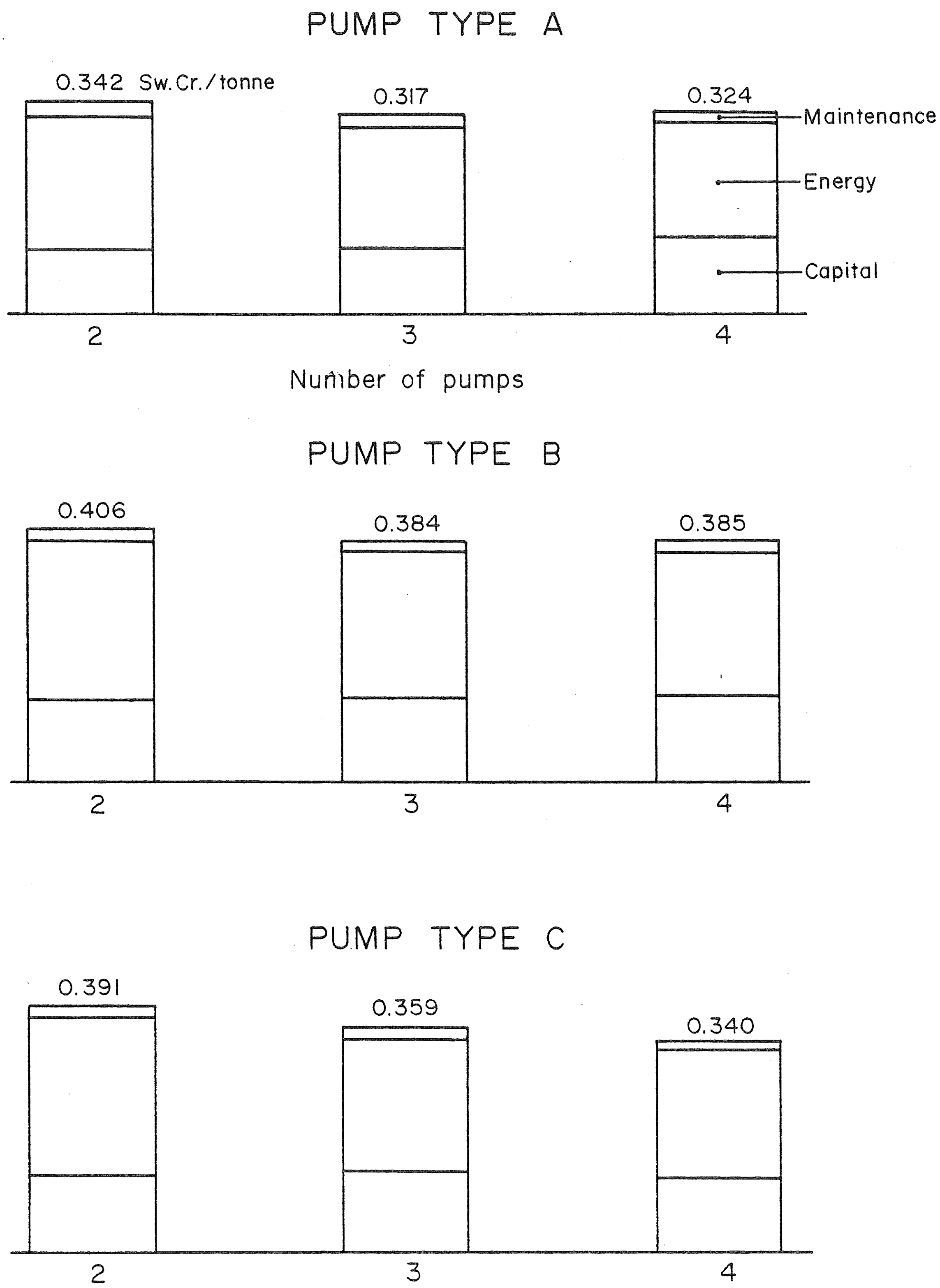

Figure 8. Comparison of pumping costs for three different centrifugal pump types in series installation. Pumping of 60 tonnes of mine tailings per hour at a total head of 70 m.

The relation of the cost of capital, energy, and maintenance is shown in Figure 9.

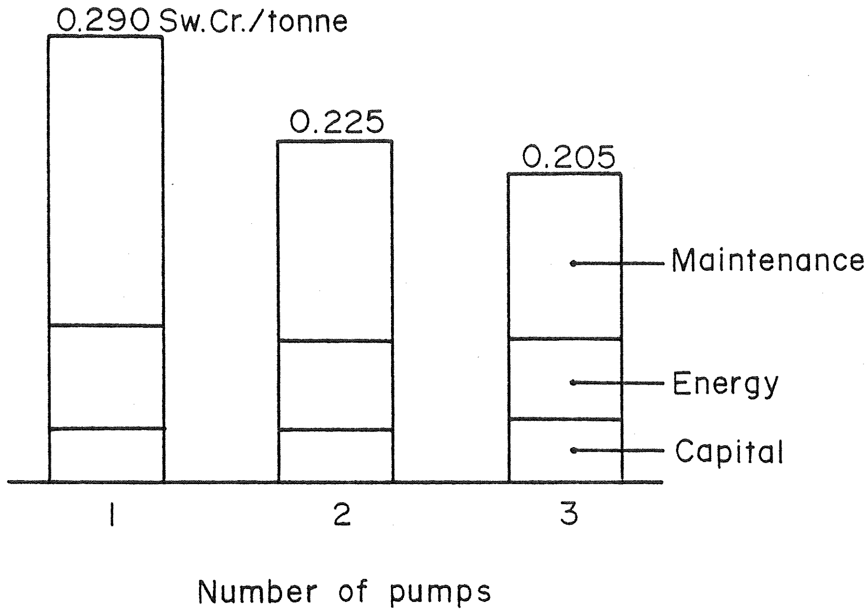

Figure 9. Demonstration of the great influence of maintenance on the total pumping costs in an autogenous ore milling application.

CONCLUSION

The model presented here makes it possible to carry out systematic studies on the effect of different operating conditions on the total pumping cost when centrifugal slurry pumps are used in series installation. The total pumping cost included the costs of capital, energy, and maintenance. The possible disadvantages of having several pumps in series were not considered in the model.

The dominant influence of the energy cost was shown in an example of the pumping of mine tailings. The importance of choosing pumps which operate with high efficiency in such applications is emphasized.

In ore pumping in autogenous milling operations, the great influence of the rotary speed on the wear and the total pumping cost was demonstrated. The use of pumps with large impellers which operate at low rotary speeds is normally of great advantage in such applications.

With a fixed number of pumps in series, there generally exists an "optimal" solids concentration for which the total pumping cost has a minimum value.

Series installation of centrifugal pumps and pumping in onestage with positive displacement pumps were compared in an example of hydraulic hoisting of iron ore (maximum particle size = 1.5 mm) a vertical distance of about 200 m. The cost of maintenance and supply of positive displacement pumps in such applications is not known in detail. However, based on the comparisons discussed in this study, it is believed that there may be an economic potential in slurry transportation by positive displacement pumps of coarser particles than those transported by such pumps today.

The model can, of course, also be applied to the pumping of arbitrary liquids, which normally is a simpler problem to analyze, as well as to the flow of air in fans. However, so far, no applications of the model have been conducted outside the field of slurry transportation.

RECOMMENDATIONS FOR FURTHER WORK

A further development of the model should include an evaluation of the extra costs of having a large number of centrifugal pumps in series and their influence on the reliability of the system.

The uncertainty of how very high solids concentrations affect the performance of a centrifugal slurry pump merits further investigation. This is an important subject because in modern ore-milling operations it is often preferable to transport the ore slurry through the operations at very high solids concentrations (up to 70-80 percent by weight).

The high cost of energy draws the attention to the use of alternative types of pumps. Operating data of positive displacement pumps in slurry service are of great interest, especially in non-traditional applications of this type of pumps.

A complete comparison of different pumping systems must also include an evaluation of different maintenance strategies. For example, the successive wear in a pump causes a drop in the efficiency; however, this drop develops in different ways in a centrifugal pump and in a positive displacement pump, respectively.

NOTATION

C	delivered volumetric concentration
C_w	delivered concentration by weight
$\bar{C}_D$	weighted drag coefficient
H	pump head developed in slurry service, meters of slurry
H_o	pump head developed in water service, meters of water
N	pump rotary speed
N_{Ref}	reference pump rotary speed
P_o	power input to pump when pumping water
Q	flow rate of slurry
Q_o	flow rate of water
Q_s	flow rate of solids
X	a function
c_1	constant
c_2	constant
g	acceleration due to gravity
ϕ	indicates "function of"
η	pump efficiency in slurry service
η_o	pump efficiency in water service
ρ_o	density of water
ρ_s	density of solids

<u>Subscripts</u> (used in the description of the model)

i	refers to an input value of flow rate
j	refers to an input value of pump rotary speed
i,j	refers to input values of pump head and power, respectively
I	refers to the maximal number of input values of flow rate
J	refers to the maximal number of input pump rotary speed
x	refers to an arbitrary flow rate, rotary speed, and head, respectively

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APPENDIX A

Hydrotransport 6

SIXTH INTERNATIONAL CONFERENCE ON THE
HYDRAULIC TRANSPORT OF SOLIDS IN PIPES
SEPTEMBER 26th-28th, 1979

PAPER G1

PERFORMANCE OF A CENTRIFUGAL PUMP WHEN PUMPING ORES AND INDUSTRIAL MINERALS.

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Summary

The present study is an experimental investigation of the performance of a rubber-lined centrifugal pump when pumping ores and industrial minerals with concentrations by weight of up to 65-70%. The upper limit of particle size was about 8 mm and the densities varied from 2,300 kg/m^3 to 4,200 kg/m^3. The solids were delivered from in-plant crushing and milling processing.

The clear water head and efficiency were lowered 30-40% by the coarsest iron ore investigated. Most experimental data were correlated to a formula expressing the influence of solid concentration and solid properties on the reduction in pump head. The drop in efficiency was equivalent to the drop in head up to volumetric concentrations of 20-25%. With higher concentrations the drop in efficiency became greater than the drop in head.

Held at University of Kent, Canterbury, U.K.
Conference Organised by BHRA Fluid Engineering, Cranfield, Bedford, England.

Nomenclature

C	delivered volumetric concentration
C_D	drag coefficient
C_w	delivered weight concentration
$\bar{C}_D$	weighted drag coefficient
D	impeller diameter
H	pump head developed in slurry service, metres of slurry
H_o	pump head developed in water service, metres of water
K	reduction factor
N	number of fractions
P	power input to pump when pumping slurry
P_o	power input to pump when pumping water
P_{in}	input power to pump motor at run
P_{out}	output pump power at run
Q	volume flow rate of mixture
R_H	reduction factor for pump head
R_η	reduction factor for pump efficiency
W_j	fractional average terminal settling velocity
Z	particle size distribution factor
c_1	constant
d	characteristic particle size
d_j	fractional particle size
d_{50}	particle median sieve size
d	weighted particle size
g	acceleration due to gravity
j	particle size fraction index
p_j	weight fraction of solid particles
s	mixture density ratio
s_s	solid density ratio
w	characteristic terminal settling velocity
$\bar{w}$	weighted terminal settling velocity
ψ	particle shape factor
ϕ	indicates, "function of"
α	exponent
β	exponent
γ	exponent
η	pump efficiency in slurry service
η_o	pump efficiency in water service
η_{Motor}	efficiency of pump motor
η_{Transm}	efficiency of pump motor transmission
η_{Tot}	total efficiency
μ_o	dynamic viscosity of water

ρ	density of mixture
ρ_o	density of water
ρ_s	density of solids
Re_w	particle Reynolds number

Introduction

In order to design a slurry transportation system based on centrifugal pumps it is necessary to know the influence of solids on the performance of the pumps. For example, even a small deviation in pump speed, over a long period of time, from a predetermined average value may be very expensive due to increased wear.

The number of published studies on the influence of solids on the performance of centrifugal pumps is relatively small. Besides, most work has generally dealt with slurries composed of particles relatively close in size.

The objective of the present work was an experimental determination of the effect of some ores and industrial minerals on the performance of a centrifugal pump. The solids were taken directly from in-plant crushing and milling processes, thus covering a wide particle size distribution, Fig. (1).

Analysis

The pump head and efficiency are generally lowered by the presence of solids. When pumping slurries, the relative reduction of the clear water head and efficiency for a constant flow rate and rotary speed may be defined by the following two ratios:

the head ratio: H/H_o

the efficiency ratio: η/η_o

where

H = head developed in slurry service, metres of slurry

H_o = head developed in water service, metres of water

η = pump efficiency in slurry service

η_o = pump efficiency in water service

Most investigational work carried out on performance tests with industrial slurries in centrifugal pumps has shown that the power consumption increases proportionally to the density ratio:

$$P = s P_o \qquad \ldots\ldots 1$$

where

P = power input to pump when pumping a mixture

P_o = power input to pump when pumping water

s = mixture density ratio, ρ/ρ_o

and that a lower head output of the pump is directly related to a reduction in efficiency:

$$\frac{H}{H_o} = \frac{\eta}{\eta_o} \qquad \ldots\ldots 2$$

see for example, Stepanoff (1965).

The slurry absorbs the same kinetic energy as a heavy fluid with density ρ. In principle, pump head can only be developed by the energy imparted to the liquid by the impeller. The diminution of the area for the liquid in the impeller channels may reduce the amount of energy transfer to the liquid in the pump. Based on this assumption, Frazier (1968) explained that the head and efficiency of a centrifugal pump were reduced in the following way:

$$\frac{H}{H_o} = \frac{\eta}{\eta_o} = \frac{1 - C}{s} \qquad \ldots\ldots 3$$

where C is the concentration by volume. Eq. (3) corresponds to the ratio of energy imparted per mass unit of mixture and water, respectively. However, experience has shown that the particle size has a marked influence on the performance, and therefore Eq. (3) alone does not present a complete representation of the flow situation of the flow situation in the pump.

The flow of solids through the pump causes additional hydraulic losses due to relative motion of coarse particles or viscous effects by a high concentration of small particles. Because of the different densities of solids and liquid, a separation of the two components will take place in the acceleration field of the pump, which also results in additional friction loss.

Observations by Herbich (1962) showed that the particle velocity leaving the impeller was 4 to 5 times as large as the velocity of the fluid (Plastic beads: diameter = 3 mm, density = 1190 kg/m^3). Similar findings from theoretical calculations were reported by Fairbanks (1941). However, at the discharge end of the pump, Wiedenroth (1970) found from experimental investigations with a quartz mineral of particle sizes from 0.1 mm to 10 mm that the velocity of the components were equal due to volute action.

Burgess et al. (1976) wrote for the flow of slurries in a particular pump

$$\frac{H}{H_o} = \phi \left(C_w, \frac{d_{50}}{D}, s_s\right) \qquad \ldots\ldots 4$$

$$\frac{\eta}{\eta_o} = \phi \left(C_w, \frac{d_{50}}{D}, s_s\right) \qquad \ldots\ldots 5$$

where C_w is the concentration by weight, d_{50} is the particle median sieve size, D is the impeller diameter and s_s is the solid density ratio. Burgess et al. assumed the head and efficiency ratio to be independent of flow rate and pump rotary speed, an approximation confirmed experimentally by Stepanoff (1965) and others.

The functional relationships given by Eqs. (4) and (5) indicate that the relative performance of a particular slurry pump of given impeller diameter is altered by concentration, particle size, and density, However, most investigations have generally dealt with slurries composed of particles relatively close in size. It is believed that the depression in head and efficiency will be less for a mixture containing solids of a broad size distribution than that of a mixture of uniform solids with comparable mean particle size. The liquid and the finer particles may form a medium in which the separation between the components is reduced, which may cause a smaller change in the internal flow characteristics. Therefore it is questionable if d_{50} can be used as a representative particle size of slurries covering a considerable range of particle sizes. More generally, the solid particles may be related to a characteristic particle size, d, and a characteristic particle distribution factor, Z.

The influence of the shape, ψ , of the particles on pump performance must also be considered. For example, Wiedenroth (1970) reported greater losses in efficiency for fresh angular particles compared to more rounded particles (particle sizes from 2 to 8 mm). The fact that centrifugal pumps cause preferential attrition in friable materials thus may be an important factor in industrial applications. Besides, attrition has to be considered when discussing results from tests where the slurry is recirculated.

The reported observations of a large difference in velocity between the components inside the pump while leaving the impeller indicates that the performance should be related to a characteristic terminal settling velocity, w. The variables discussed previously: d, Z, ψ , and w are believed to give a better expression of the solid properties than those given in Eqs. (4) and (5). The influence of d and w can be expressed by the particle Reynolds number:

$$Re_w = \frac{w\, d\, \rho_o}{\mu_o}$$

With Re_w, ψ, and Z then Eqs. (4) and (5) can be rewritten in the following form

$$\frac{H}{H_o} = \phi(Re_w, Z, \psi, s_s, \frac{d}{D}, C_w) \quad \ldots\ldots 6$$

$$\frac{\eta}{\eta_o} = \phi(Re_w, Z, \psi, s_s, \frac{d}{D}, C_w) \quad \ldots\ldots 7$$

where Re_w, ψ, Z, s_s, and d represent characteristic solid properties.

The influence of the parameters Re_w, Z, and ψ can be related to a representative drag coefficient C_D, i.e. Eqs. (6) and (7) can be expressed in the following way:

$$H/H_o = \phi(C_D, s_s, C_w) \quad \ldots\ldots 8$$

$$\eta/\eta_o = \phi(C_D, s_s, C_w) \quad \ldots\ldots 9$$

where the drag coefficient

$$C_D = \frac{4g}{3} \frac{d(s_s - 1)}{w^2}$$

A representative particle diameter may be represented by the mean value of each sieve size, d_j, i.e. although the particle size represents a square mesh opening it is considered as a diameter. If p_j is the weight fraction of solids of diameter d_j then a weighted value $\bar{d}$ may be defined as

$$\bar{d} = \sum_{j=1}^{N} p_j d_j \quad \ldots\ldots 10$$

where N is the number of fractions. An average representation of the terminal settling velocity may be defined by a weighted value, $\bar{w}$,

$$\bar{w} = \sum_{j=1}^{N} p_j W_j \quad \ldots\ldots 11$$

where W_j is an average value of randomly chosen particles from each fractional part p_j.

Experimental study

The tests were carried out in a pilot-plant test loop developed for studies on hydraulic hoisting at the Department of Hydraulics, Chalmers University of Technology. The schematic lay-out of the open loop test facility is shown in Fig. (2). The energy of the downcoming slurry was utilized to obtain uniform mixing in the mixing tank without baffles or circulating pump arrangements.

The slurry density and flow rate were measured directly by diverting the flow at the top of the vertical test loop to a sampling tank as shown in Fig. (2). The pressure rise over the pump was measured with Bourdon pressure gauges which were fitted with flushing water lines. The pump head was then determined with an accuracy of about $\pm 4\%$, from the pressure rise and flow rate readings.

The pump efficiency characteristics were obtained from the pump head determination and the measured values of slurry density, flow rate, current and voltage. The efficiency of a centrifugal pump, η , is defined as the ratio of pump energy output to the energy input applied to the pump shaft. The total efficiency, η_{Tot}, includes the efficiency of the motor, η_{Motor}, and the efficiency of the motor-pump transmission, η_{Transm}:

$$\eta_{Total} = \eta \cdot \eta_{Transm} \cdot \eta_{Motor} \qquad \ldots\ldots 12$$

$$\eta_{Total} = \frac{P_{out}}{P_{in}} \qquad \ldots\ldots 13$$

where the input power, P_{in}, was obtained from the ampere- and voltmeter readings. The output power, P_{out}, was given by:

$$P_{out} = \rho g Q H \qquad \ldots\ldots 14$$

From the ρ, Q and H measurements, P_{out} was then determined from Eq. (14) with an accuracy of about $\pm 10\%$.

The efficiency of the LD-motor, η_{Motor}, for different values of rotary speed and input current was obtained from the motor manufacturer. The losses in the belt drive transmission can generally be estimated at 4% - 5%, Nilsson (1976). Therefore, a constant value of 95.5% was used for η_{Transm}. The indirect evaluation of η from Eqs. (12) and (13) yield pump efficiency with an accuracy of about $\pm 15\%$.

The pump was a 0.152 m by 0.152 m hydroseal rubber-lined Morgårdshammar BC-pump[1)] with a 3-vane impeller of 0.43 m in diameter. Slurry dilution must be minimized for experimental reasons, therefore, a constant seal water flow of $3.3 \cdot 10^{-3}$ m^3/min was permitted which was considerably lower than the proposed 0.015 m^3/min. The motor was an ASEA LD-29 DC-motor (66 kW) with the speed regulated by a thyristor converter with tachogenerator feedback, and the pump rotary speed was regulated by a potentiometer.

The pump utilized in the study operated in the domain showed in Fig. (3). The pump performance test was carried out at two pump speeds, 12.67 rps and 19 rps.

Mineral properties

The solid density, particle size distribution, and terminal settling velocity of the pumped minerals were shown in Fig. (1). The weighted particle sizes and weighted terminal settling velocities were determined by Eqs. (10) and (11) respectively.

In the determination of $\bar{d}$, particle sizes of less than 0.074 mm were assumed to be uniformly distributed down to 0, because the particle size distribution of the fine particle part was not investigated. In the determination of $\bar{w}$, W_j was taken = 0 for particle sizes less than 0.074 mm, a reasonable approximation because the portion of small particles in general was low. The calculated values of $\bar{w}$ and the experimentally obtained distribution curves of W_j and d_j are illustrated in Fig. (4). The

[1)] manufactured under licence from the Allen-Sherman Hoff Pump Co., U.S.A.

particle size d_i corresponding to W_i was represented by the mean of each standard sieve size. For comparison, the standard C_D - Re_w relationship for spheres is represented in the form of graphs given by Weber (1973) in Fig. (4).

The relative tendency of the solids to degrade during circulation in the test loop was negligible, for the experimental data analysed in this study.

Discussion of results

The experimental results for solid concentrations by volume of about 20% for the perlite, lead ore, primary ground iron ore, and coarse iron ore agree approximately with general accepted trends, for example here illustrated by reduction curves presented by McElvain (1974), Fig. (5). McElvain expressed the dependence of the solids density, particle size, and concentration by a reduction factor K, Fig. (5). In an empirical approach, McElvain found a linear dependence between the efficiency and head ratio and the volumetric concentration. He related the factor K to C = 0.20 (20%) and obtained:

$$\frac{H}{H_o} = \frac{\eta}{\eta_o} = 1 - K \frac{C}{0.20} \qquad \ldots\ldots 15$$

The general trends were, however, not in accordance with the experimental results of the complex ore and the crushed ore as can be seen in Figs. (6) and (7), respectively.

From the graphs in Fig. (7) it can be seen that the clear water efficiency was reduced by about 70% for the angular crushed granite. On the other hand, the drop in efficiency was only a few percent below the clear water efficiency for the complex ore, Fig. (6). The graphical representations in Figs. (6) and (7) indicate that Eqs. (1) and (2) do not hold uniquely for these solids. It can be seen from Fig. (6) that:

$$\eta / \eta_o > H/H_o$$

for a given concentration of complex ore. The corresponding power requirement is less than that given by Eq. (1), i.e.

$$P < s\, P_o$$

Furthermore, for the crushed granite in Fig. (7), it can be seen that:

$$\eta / \eta_o < H/H_o \qquad \ldots\ldots 16$$

which corresponds to:

$$P > s\, P_o \qquad \ldots\ldots 17$$

Similar results as are shown in Fig. (6) were obtained with the higher pump speed = 19 rps. No experimental data were available for the high pump speed with the crushed granite, because of power limitations in the experimental facility.

The results shown in Fig. (6) and (7) demonstrate the complex influence on the pump performance of the physical and the chemical properties of the solid component. The complex ore contained a large amount of talc mineral. Talc is a friable waxy mineral, which, when mixed with water, has a pronounced lubricating effect. The almost negligible reduction in efficiency, Fig. (6), may therefore be explained by a decrease of the disc friction within the rubber-lined pump, which nearly compensated the extra hydraulic loss due to solids.

The results presented so far have not included volumetric concentrations of over about

20%. For higher concentrations, it was observed that the solid concentration has a predominant influence on the pump performance. Therefore, the dependence of the solid concentration on the relative reduction in head and efficiency was evaluated in detail.

The observed values of the efficiency ratio, η/η_o, could in some cases be correlated to Eq. (3),

$$\eta/\eta_o = \frac{1 - C}{s}$$

With

$$C_w = \frac{C\, s_s}{s}$$

and the mixture density ratio expressed by:

$$s = 1 + C(s_s - 1)$$

in Eq. (3), then it follows that:

$$\eta/\eta_o = 1 - C_w \qquad \ldots\ldots 18$$

The reduction in head and efficiency, dimensionlessly expressed by H/H_o and η/η_o were separately represented as functions of C_w and the two pump speeds for each mineral investigated. The resulting relationships were graphically represented for concentrations by volume, C, from about 20% to about 40%, which correspond to concentrations by weight, C_w, in the range of 45 to 70%. A representative graph is shown in Fig. (8).

It can be seen from Fig. (8) that the reduction in efficiency diverges from the drop in head for higher concentrations. The drop in efficiency is greater than the reduction in head which corresponds to a power requirement, P, of:

$$P > s P_o \qquad \text{(Eq. 17)}$$

Thus, the investigational data here exemplified in Fig. (8) comprising high concentrations are not in agreement with Eqs. (1) or (2). It was found that a significant reduction in efficiency occurred at concentrations by weight of approximately 60%, 40%-45% and 45%-50% for the iron ores, perlite, and lead ore, respectively. These concentrations by weight of the three solids of different densities correspond to concentrations by volume of 20%-25%. These concentrations represent threshold values over which the drop in efficiency considerably exceeds the drop in head. For the perlite and lead ore slurries, the reduction in efficiency approaches the values given by Eq. (18) for concentrations by weight larger than about 50%.

Increased power consumption due to high solid content here expressed by Eq. (17) can generally be related to increased disc friction loss. However, the mechanism of the flow inside a centrifugal pump is not completely understood, not even with clear water flow.

The drop in efficiency was not so pronounced for higher concentrations of the coarse iron ore. This behaviour may be related to that of a slurry composed of solids with a wide particle size distribution which influences the internal flow characteristic to a lesser degree than a slurry of uniformly distributed solids.

From Fig. (8) it can be seen that the drop in head is independent of the pump speed but the small number of data points shows a slightly smaller decrease in efficiency for the higher pump speed, 19 rps.

Reduction in head

It was found that the head ratios were more uniquely related to C_w than the efficiency ratios. In fact, all investigational data indicate that the head ratio may be represented by Eq. (8).

It is convenient to express the influence of solids on the clear water pump performance in the form of a reduction factor. The reduction factor of the drop in head is defined here by:

$$R_H = 1 - H/H_o \qquad \ldots\ldots 19$$

The influence of particle shape and particle distribution was simply related to the weighted particle size, $\bar{d}$, and to the weighted settling velocity, $\bar{w}$, i.e. the characteristic drag coefficient was expressed by a weighted value:

$$\bar{C}_D = \frac{4g}{3} \frac{\bar{d}(s_s - 1)}{\bar{w}^2} \qquad \ldots\ldots 20$$

Finally, with $\bar{C}_D$ from Eq. (20) in Eq. (8) and expressing this equation by the reduction factor given by Eq. (19), it follows that:

$$R_H = \phi(\bar{C}_D, s_s, C_w) \qquad \ldots\ldots 21$$

Eq. (21) was related quantitatively to all experimental data for the minerals used in this study. A preliminary comparison of the plotted experimental results of the iron ores, lead ore, and the perlite indicate that the functional relationship in Eq. (21) can be expressed in the following form:

$$R_H = c_1 C_w^{\alpha} (s_s - 1)^{\beta} \bar{C}_D^{\gamma} \qquad \ldots\ldots 22$$

where c_1, α, and β are constants ≥ 0 and γ is a constant ≤ 0. For given solid properties defined by the solid density ratio s_s and the drag coefficient $\bar{C}_D$, R_H appears to be approximately proportional to C_w with the exponent $\alpha = 0.7$. The value of the coefficients c_1, β, and γ were estimated as 0.32, 0.7 and -0.25 respectively. Thus,

$$R_H \approx 0.32\ C_w^{0.7} (s_s - 1)^{0.7}\ \bar{C}_D^{-0.25} \qquad \ldots\ldots 23$$

The validity of Eq. (22) in the form of Eq. (23) is shown in Fig. (9), where calculated values of R_H are graphically represented against observed values of $(1 - H/H_o)$. Error bands of $\pm 15\%$ have been drawn for comparison.

Reduction in efficiency

The reduction factor of the drop in efficiency was defined by:

$$R_\eta = 1 - \eta/\eta_o \qquad \ldots\ldots 24$$

Thus, the reduction factor for the complex ore and the crushed granite was in the range of 0.9 - 1 and 0.25 - 0.4, respectively, Figs. (6 - 7). Inspection of the experimental results of the other minerals and ores showed that the drop in efficiency was influenced by the concentration and individual mineral properties in a more complex way than the drop in head. It was found that the reduction in efficiency, R_η, mainly could be estimated by:

$$R_H \leq \quad R\eta \leq C_w \qquad \ldots\ldots 25$$

where the right term in Eq. (25) refers to Eq. (18). No further attempts were made to correlate individual experimentally determined efficiencies, except Eq. (25).

Conclusions

The experimental results in this study demonstrate that Eqs. (1) or (2) cannot be taken as an universal statement when pumping industrial slurries, i.e. the power consumption increases not always proportionally to the mixture density ratio.

The experimental results of the complex ore and the crushed granite, in Figs. (6) and (7), demonstrate the complex influence of not only the size, shape, and distribution of the solids but also specific mineral properties related to the basic mineral structure and the chemical composition.

For the pump used in this study, it was found that the reduction in efficiency exceeds the reduction in head if the volumetric concentration exceeds 20%-25% for the iron ores, the perlite, and the lead ore, i.e.

$$P > s\ P_o \qquad \text{(Eq. 17)}$$

The experimental results for the iron ores, lead ore, and perlite mineral showed that the expression in Eq. (23) can be used in industrial applications for determining the reduction in head within an error of about $\pm 15\%$.

A correlation in the simple form of Eq. (22) may be a promising approach for determining the reduction in head for a large variety of industrial slurries and centrifugal slurry pumps.

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Table 1. Properties of the ores and industrial minerals used in the experimental study

Solids	Average solid properties			Symbol	
	d_{50} (mm)	d_{max} (mm)	Density (kg/m^3)		
Coarse iron ore	1.8	8	4150	A	Sinter fines from the LKAB-mine at Kiruna
Primary ground iron ore	0.25	1	4003	B	Taken from the dressing process (rod mills) at the LKAB-mine at Malmberget
Complex ore	0.22	4	3395	C	Zinc-lead-copper ore from work face at the Boliden AB-mine in Garpenberg
Lead ore	0.4	2	2672	D	Taken from the dressing process (rod mills) at the Boliden-AB mine in Laisvall
Perlite	0.35	2	2341	E	From deposit in Iceland. After processing (thermal expansion), perlite has been suggested as a substitute for asbestos.
Crushed granite	3.0	4	2676	F	From crushing plant in Göteborg

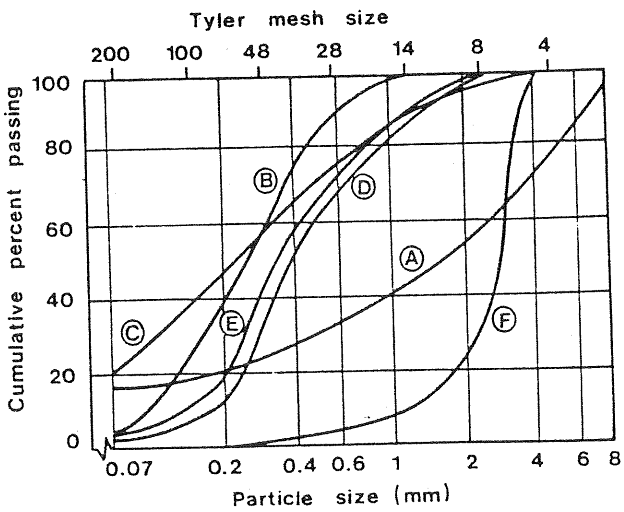

Fig. 1. Properties of the ores and industrial minerals used in the experimental study

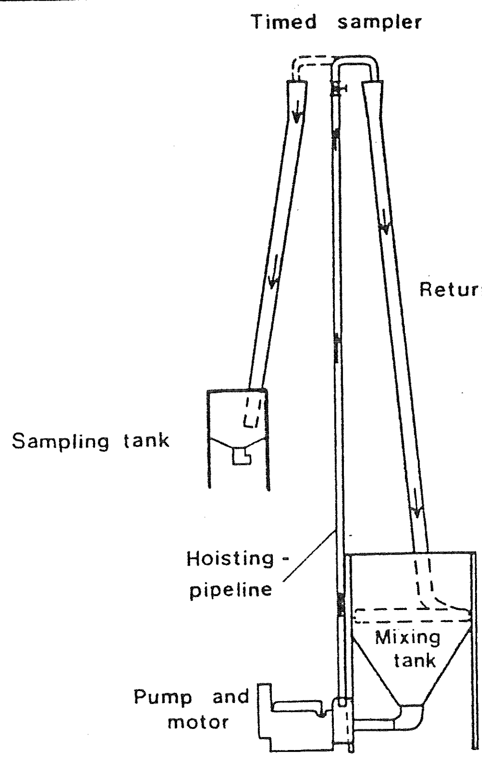

Fig. 2. Open loop testing facilities - schematic lay-out

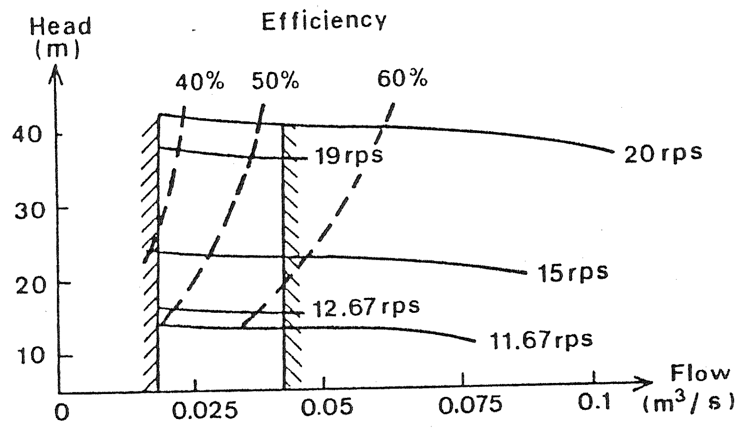

Fig. 3. Operational domain of the pump

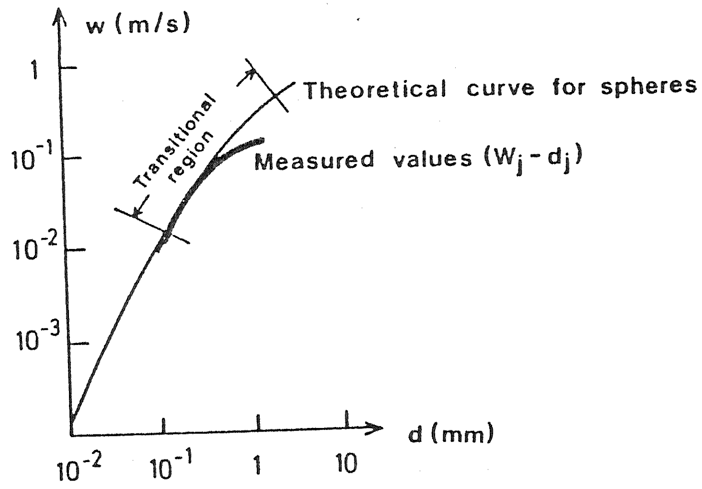

Fig. 4. Measured terminal settling velocities compared with the standard relationship for spheres.

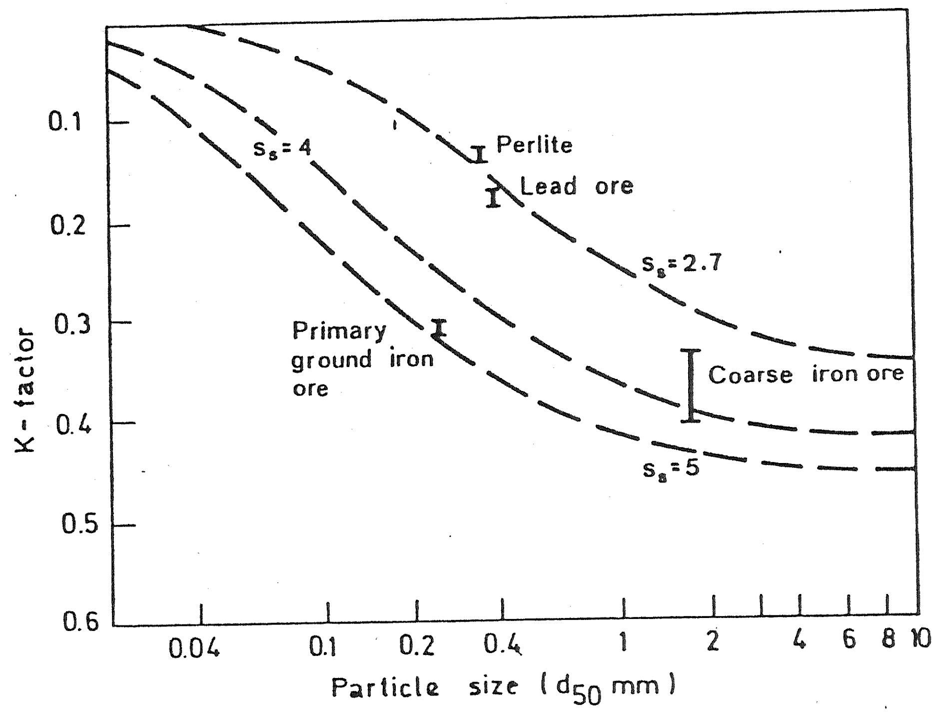

Fig. 5. Comparison of measured reduction in head and efficiency of the primary ground iron ore, lead ore, perlite mineral, and coarse iron ore to the reduction curves given by McElvain (1974). Solid concentrations in the range of 20% and pump speeds of 12.67 rps and 19 rps.

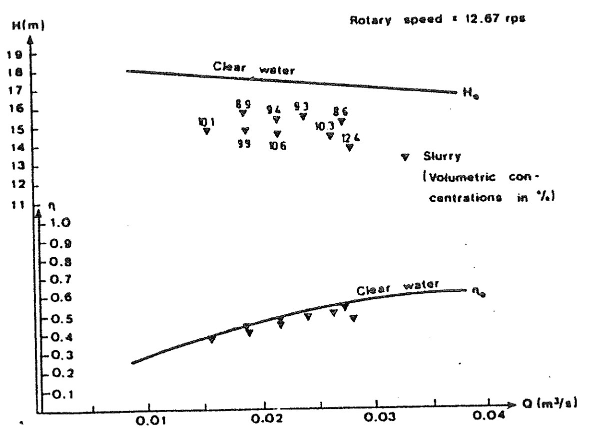

Fig. 6. Depression in head and efficiency for the complex ore.

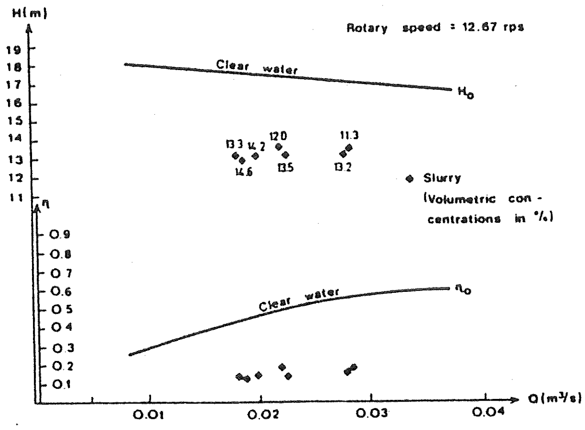

Fig. 7. Depression in head and efficiency for the crushed granite

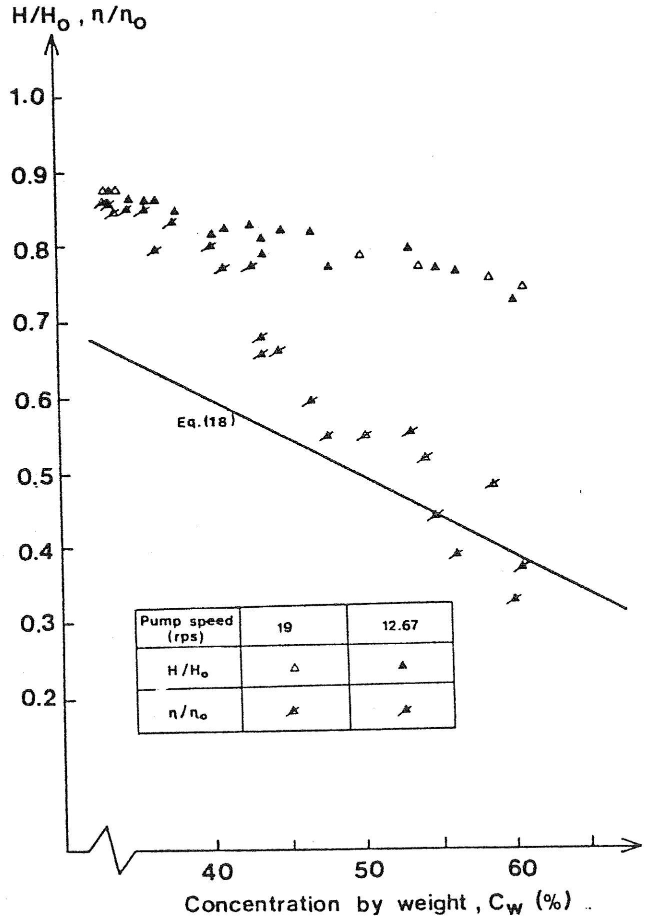

Fig. 8. Relative reduction in head and efficiency versus concentration by weight for the perlite mineral.

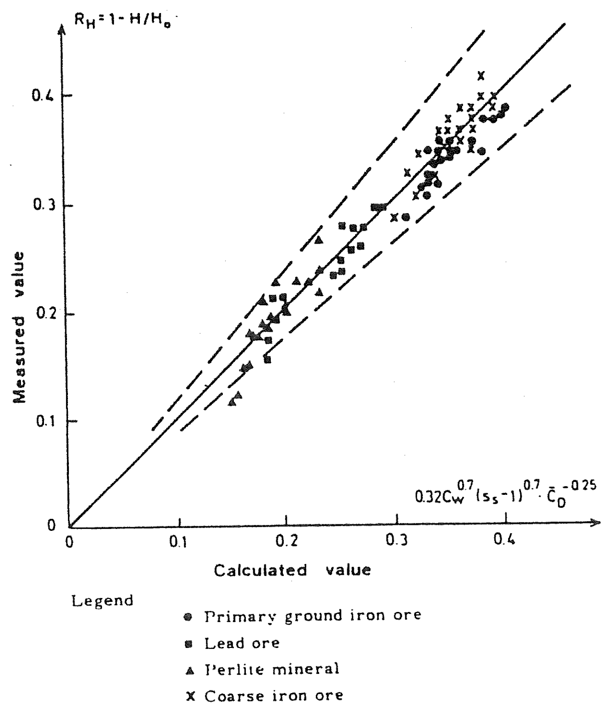

Fig. 9. Observed reduction factors in head versus calculated values with Eq. (23). The dashed lines indicate error of $\pm$ 15%.

APPENDIX B

A MODIFIED POSITIVE DISPLACEMENT PUMP

A modified positive displacement pump has been developed and patented by a Swedish inventor.[1]). The pump was first presented at the conference "Hydrotransport 5" in Hannover, Germany, in 1978. The construction is based on the piston-membrane principle, where the membrane has been replaced by a flexible tube element of rubber, (Figure 1). The tube-element is driven by a hydraulic oil transmission system from a detached drive element. The valves, however, must operate in the slurry as for conventional positive displacement pumps. The in-line construction can be built us by a number of tube elements connected in parallel.

The pump was tested in 1979 in a prototype installation at an ore dressing plant in Northern Sweden at a flow rate of about 0.004 m^3/s and a working pressure of about 1.5 MPa. The pump efficiency was about 70% for the pumping of iron ore slurry with a concentration by weight of 70%. The iron ore slurry had a solid density of 5000 kg/m^3 and a maximum particle size of about 0.15 mm.

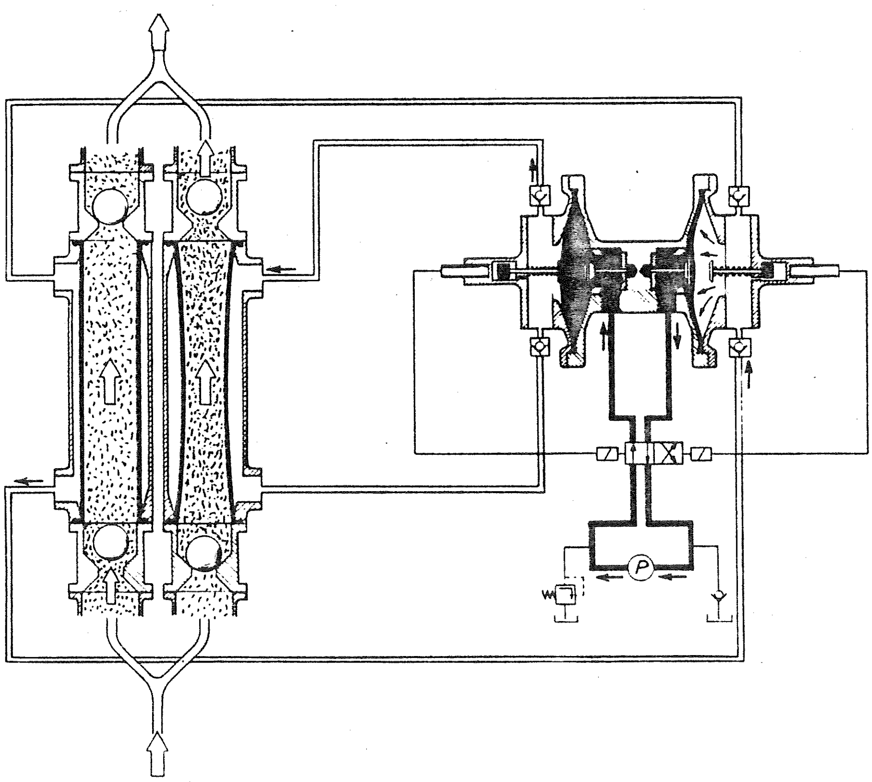

Figure 1. Schematic sketch of a new positive displacement pump.

1) Henrik Kitsnik, Eurocontrol AB, Säffle, Sweden

The first full-scale pumps are planned to be manufactured for in-plant installations and tests of maintenance and reliability at iron ore dressing plants in Sweden. The pump seems to be an interesting alternative. For example, the investment cost is estimated to be in the same range as for a centrifugal pump with a variable speed drive in ordinary in-plant iron ore handling.