A
DISSERTATION REPORT
ON

DESIGN OF 300 TPD OXYGEN DELIGNIFICATION PLANT

Submitted in partial fulfillment of the
requirement for the award of the degree
of

MASTER OF TECHNOLOGY
IN
PULP AND PAPER

submitteed By :
ANUJ MITTAL

Under the Guidance of:
DR. DHARMDUTT
AND
DR. A. K. JINDAL

DEPARTMENT OF PAPER TECHNOLOGY
INDIAN INSTITUTE OF TECHNOLOGY, ROORKEE
SAHARANPUR CAMPUS, SAHARANPUR-247 001
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CANDIDATE'S DECLARATION

I hereby declare that the work which is being presented in this dissertation report entitled **"DESIGN OF 300 TPD OXYGEN DELIGNIFICATION PLANT"** in partial fulfillment of the requirement for the award of the degree of Master of Technology in Pulp and Paper, I I T Roorkee, is an authentic work of my own efforts carried out, under the supervision of **Dr.Dharmdutt**, Associate Professor and **Dr.A.K.Jindal**, Assistant Professor, Department of Paper Technology, Indian Institute of Technology, Roorkee (Saharanpur Campus).

The work embodied in the dissertation report has not been submitted by me for the award of any other degree.

ANUJ MITTAL
M.Tech.(Final),Pulp&Paper

DATE: 28th June'2007
PLACE: SAHARANPUR

This is to certify that the above statement made by the candidate is correct to the best of my knowledge.

Dr. DHARMDUTT
(Associate Professor)
DPT, Saharanpur
GUIDE – I

Dr. A.K.JINDAL
(Assistant Professor)
DPT, Saharanpur
GUIDE – II

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Chapter -1

INTRODUCTION

Oxygen delignification can be defined as the use of oxygen and alkali to remove a substantial fraction of lignin in unbleached pulp. The process is usually conducted under pressure and delignification is normally in the range of 35 – 50 %. Oxygen is usually applied to Kraft wood pulps but can be used for sulfite, secondary fiber, nonwood, and other types of pulp.

The main benefits of oxygen bleaching are environmental. These derive from the fact that both the chemicals applied to the pulp and the materials removed from the pulp are compatible with the Kraft chemical recovery system. This enables the recycling of oxygen stage effluent to the recovery system by way of the brown stock washers, decreasing the potential environmental impact of the bleach plant. The decrease is roughly proportional to the amount of delignification achieved in the oxygen stage. This applies not only to chlorinated organic by-products, but also to other environmental parameters associated with the bleach plant effluents, including BOD, COD and color. The decrease in color, however, is larger than expected on the basis of the lignin removed in the oxygen stage.

The industrial application of oxygen bleaching has expanded very rapidly in the recent years. The first commercial system was started up until 1970 and the world capacity in 1980 was only about 10,000 tons per day. By 1992 the number of operating systems stood at 155, with a total capacity of about 85,000 tons per day as illustrated in Figure 1 (1). Furthermore, this rapid growth is likely to continue. As trends away from the use of chlorine and chlorine containing compounds intensify, both technical and economic considerations will mandate the use of oxygen. Ozone, for example, though it possesses certain advantage over oxygen, will probably see application only in conjunction with oxygen. Oxygen will be used to predelignify pulp to the point where the necessary ozone charge becomes small enough to be economical and selective.

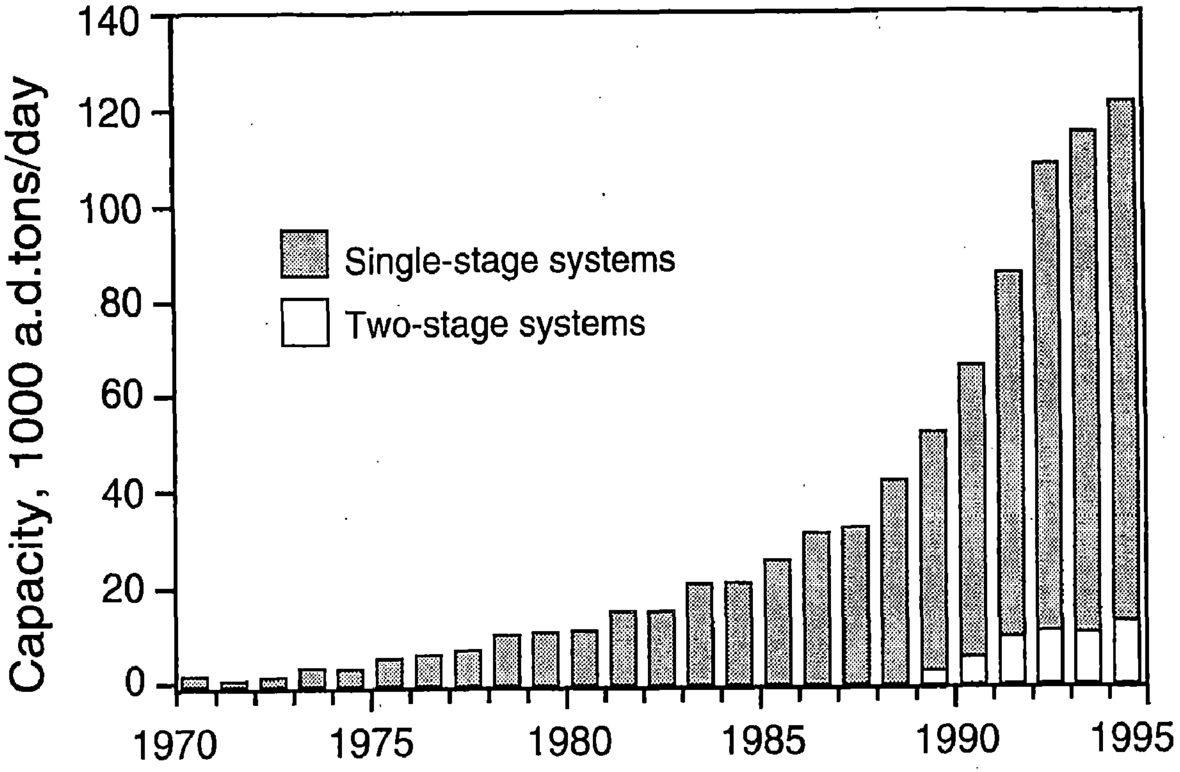


Fig. 1. Worldwide production capacity of oxygen-delignified pulp.

CHAPTER- 2

HISTORY

The gestation period of commercial oxygen bleaching was a long one. It began in 1867 with the patenting of an "Improvement in Bleaching of Paper Pulp" by Joy and Camp ell (2) that involved passing heated air through an agitated pulp suspension. In a 1915 patent, Mueller advanced the state of the art by providing for pressurized operation and addition of an "alkaline earth metal hydrate". Harris, Marshall and Sankey disclosed further improvements in 1954. During the period from 1956 to1960, Nikitin and Akim made a series of contributions stemming from their interest in combining the delignification and viscosity control steps of dissolving pulp manufacture. In 1960 and 1961, Grangaard and Saunders patented additional improvements aimed at making the process more commercially feasible.

Despite these advances, oxygen delignification remained uncommercialized. The main reason was that under conditions necessary to achieve appreciable delignification, cellulose was attacked and degraded, leading to sharp decreases in pulp viscosity and strength. This obstacle was finally removed when Robert and co-workers (3) discovered that addition of small amounts of magnesium salts sharply decreased the damaged suffered by the polysaccharide components of the pulp during oxygen bleaching. The remaining obstacle to commercialization, providing for the high mass transfer rates made necessary by the low solubility of oxygen in water, was soon overcome, and the first commercial system was started up in South Africa in 1970 (4).

PROCESS OVERVIEW:

Figure 2 shows a simplified oxygen delignification stage flow sheet. In a typical embodiment of the process, unbleached pulp discharged from the pulper of the last brown stock washer passes to a steam mixer and then to a feed tank. Alkali, usually sodium hydroxide or oxidized white liquor, is added to the pulp at the base of feed tank and mixed with the pulp by the discharge pump. The pulp then goes to one or more medium consistency

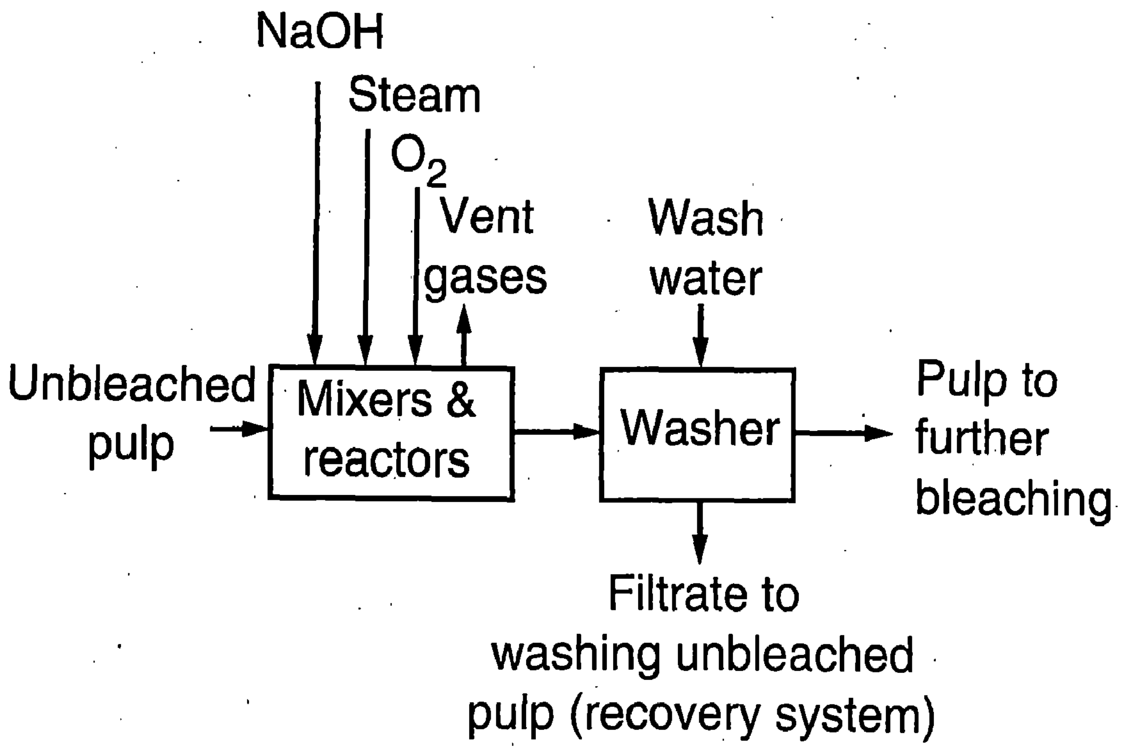


Fig. 2. Oxygen delignification stage flowsheet.

high shear mixers where oxygen gas is added together with additional steam. The mixers disperse the oxygen in the pulp, which then passes to an up flow pressurized reactor. After a retention time of about one hour, the pulp is blown to a gas separator mounted on a blow tank. Product gases and unreacted oxygen is vent out and the pulp enters the first of the two washing stages, perhaps a wash press followed by a rotary drum washer. The use of two washing stages at this point reflects the importance of complete removal of material dissolved in the oxygen stage. This is necessary to take advantage of the environmental benefits of the oxygen stage by capturing the lignin and other material dissolved here and returning it to the mill recovery system. Any dissolved material that escapes to the following bleaching stages become part of effluent of those stages and is a potential environmental liability.

In addition to the medium (10-14%) consistency type systems, high (20-28%) consistency systems are common, though few new ones are being installed. The main distinguishing characteristic of the high consistency system is the presence of a dewatering device to produce high consistency and a down flow reactor. Both types of systems are considered in more detail in a later section of this chapter.

CHAPTER- 3

ADVANTAGES AND DISADVANTAGES OF OXYGEN DELIGNIFICATION

The chief advantages of oxygen delignification are related to its beneficial effects on environmental parameters, although there are also some operating cost advantages. Its chief disadvantages are capital costs and increased demand on the mill recovery system.

The most obvious beneficial effect of installing an oxygen pre-delignification stage in a chlorine compound based bleach plant is a decreased need for oxidizing chemical in the delignifying part of the bleaching sequence. If this is chlorine or chlorine dioxide, there is a corresponding decrease in chlorinated organic byproducts in the bleach plant effluent (as measured by absorbable organic halide, AOX). When the delignification stage oxidant is chlorine dioxide, the effect is smaller because chlorine dioxide generates much less AOX, but the chemical cost saving is greater. In either case, the oxygen stage leads to a major decrease in biochemical oxygen demand (BOD), chemical oxygen demand (COD), and color.

Lower chemical cost results from the decreased requirement for delignifying oxidizing chemical (chlorine, chlorine dioxide, or ozone, for example), because oxygen is less expensive, and oxidized white liquor usually provides the necessary alkali for the oxygen stage at low cost. Further savings result from a decrease in the chlorine dioxide charge needed for final bleaching stages. Both decreases in chlorine dioxide requirement translate into the additional advantage of allowing a smaller chlorine dioxide generator to be used. Additional saving result from a decrease in the caustic required in the first extraction stage.

Of the disadvantages, capital cost is the most apparent. In 1987, in the course of one economic comparison of bleaching alternatives (5), the cost of a medium consistency oxygen delignification system was estimated to be $14.1 million. More recently, it has been estimated that the installed cost of such a system with two post oxygen washing stages is in the range $13 – 26 million, depending on equipment selection and site. A further drawback is the possibility of overloading systems in the chemical recovery area of the mill. The use of oxidized white liquor as an alkali source requires additional causticizing. Recycling of oxygen stage solids to the recovery system increases recovery boiler loading and water, added as steam and with chemicals, increases evaporator loading (6). An additional disadvantage is the tendency of oxygen bleaching to be nonselective at higher degree of delignification. This is not a problem if suitable carbohydrate protector is used and the degree of delignification is not too high, that is, not more than about 50%. Also included under the heading of disadvantages are increased steam costs, and an increase in overall process complexity.

FUNDAMENTAL ASPECTS OF OXYGEN DELIGNIFICATION:

Before considering equipment configurations and process variable effects in oxygen bleaching, it is worthwhile to review a number of its fundamental aspects. This provides a logical framework for understanding the design and behavior of the process, and for predicting the effects of changes in the hardware or mode of operation. The basic aspects of oxygen bleaching can be organized under the headings of chemistry and process fundamentals. Important chemical considerations include the chemistry of oxygen, reactions of lignin and carbohydrates, and the related concepts of selectivity and protectors. Features of the process to be considered are rates of chemical reaction and mass transfer, heat transfer, heat effects, chemical consumption, and reaction products.

CHAPTER- 4

SELECTIVITY AND PROTECTORS

Selectivity can be loosely defined as the ratio of attack on lignin to attack on carbohydrate. It is affected by the choice of process conditions and by the presence of pulp contaminants. Of the factors governing selectivity in oxygen bleaching, one of the most important is the transition metal content of the pulp, because these metals catalyze the generation of harmful radical species. Most pulp contains appreciable quantities of iron, copper and manganese, all of which have this effect. One approach for dealing with transition metals is to remove them by acid washing before the oxygen stage, another is to add compound to the pulp that inhibit carbohydrate degradation. These compounds are called carbohydrate protector. The protector of greatest commercial importance is the magnesium ion. The discovery of its effectiveness in 1963 by Robert and co-worker (7) provided a great impetus for the development of oxygen bleaching. Since then a considerable number of compounds have been found effective, but none is as economical as magnesium sulfate or its heptahydrate, Epsom salt. It is normally applied at levels as low as 0.05-0.1% Mg^{+2} on oven dry pulp. It is believed to function by precipitating as magnesium hydroxide, which adsorbs the metal ions, making them unavailable for catalysis of peroxide decomposition or by forming complexes with them. Figure 3 illustrates the effectiveness of magnesium in preserving pulp viscosity and is typical of selectivity obtained in the bleaching of softwood pulp.

It is apparent from the nonlinear nature of the curves in Figure 3 shows that selectivity does not remain constant as the kappa number of the pulp is decreased to progressively lower levels. However, selectivity usually remains nearly constant until about 50% of the lignin has been removed from the pulp after which it deteriorates. In virtually all of the existing pulping systems, the target degree of delignification is 60% or less. When delignification is no greater than 45-50%, the resulting pulp may be fully bleached to give a final product of strength equivalent to that of the corresponding conventionally bleached pulp, although its viscosity is slightly lower.

Selectivity may be different for different process types and is affected by

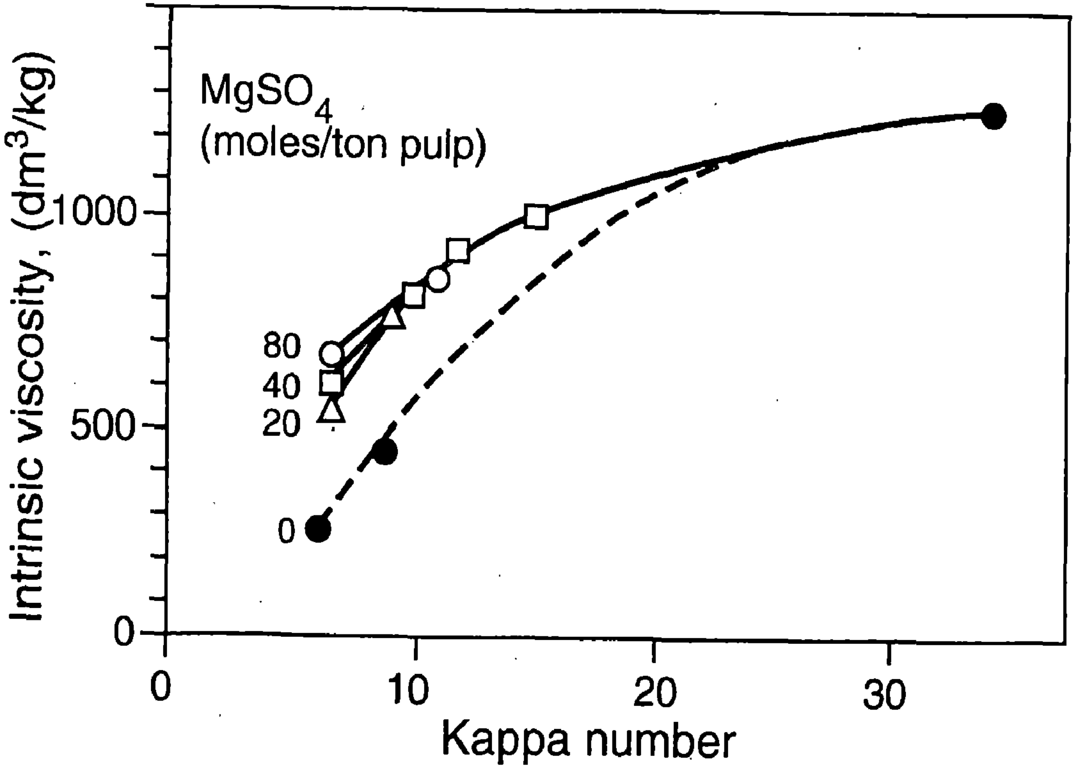


Fig-3: Effect of addition of magnesium ion on selectivity process variables, as is more fully discussed later in this chapter. The fundamental reasons for these effects are probably related to the multiplicity of reactive oxygen containing species present and their different reactivities with the variety of lignin and carbohydrate structures available to them. Hydroxyl radicals, which are among the least selective of the reactive species present, have been specifically studied from this view point by measuring their reaction rates with compounds representative of lignin and carbohydrate structures (8), the ratio of the lignin rate constant to that for carbohydrate was observed to lie between 5 and 6. Hydrogen peroxide addition can enhance selectivity presumably because it favorably affects the composition of the reactive species mixture present. Carryover of black liquor from the brown stock washers into the oxygen stage adversely affects selectivity (9).

CHAPTER- 5

REACTION AND MASS TRANSFER RATES

The rate of oxygen bleaching and its dependence on process variables are factors that determine the design and size of the equipment and the optimal choice of process conditions. In describing and predicting the rate of the overall process, it is important to distinguish between physical and chemical phenomena, either of which may determine the rate under a given set of conditions. Physical factors govern the movement of the reacting species within the pulp mass (mass transfer); while chemical factors govern the rate at which the pulp and bleaching chemicals react with one another once they are in contact (chemical kinetics).

Mass transfer is an important consideration because most oxygen bleaching stages are three phase systems. Oxygen must cross the gas-liquid interface, diffuse through the liquid film surrounding the fiber, and finally diffuse into the fiber wall before reacting. Depending on the conditions, the rate of oxygen transport can limit the rate of the overall process because of the inherent slowness of the diffusion phenomena involved.

The mass transfer problem is aggravated by the low solubility of oxygen in aqueous sodium hydroxide (10, 11). Typically, 24 times as much oxygen is consumed during bleaching at 15% consistency as can be present in the liquid phase at any one time and, at 27% consistency, 50 times as much. Obviously, a viable process must facilitate the transfer of oxygen into and through the liquid phase. This can be done by ensuring that the partial pressure of oxygen in the gas phase and the gas-liquid interfacial area are both sufficiently large.

HEAT EFFECTS, CHEMICAL CONSUMPTION AND REACTION PRODUCTS:

The requirements for application and release of heat, chemical requirements, and types of reaction products formed have significant implications for the design and operation of oxygen bleaching systems.

HEAT EFFECT:

At reasonable alkali levels, temperatures should be in the range 85-115 °C, so providing the required amount of heat in a steam mixer and recycling hot filtrate must therefore be considered, especially in medium and low consistency processes. The heat of reaction is appreciable and its removal may be a problem in high consistency, high kappa drop systems. According to one estimate, heat of reaction is 2400 kj/kg of material dissolved, another place it at 12,000 – 14,000 kJ per metric ton for each unit of kappa number reduction (12).

CHEMICAL CONSUMPTION:

The stoichiometry of oxygen bleaching is that, for each unit of kappa number decrease, about 0.13% NaOH (o.d. pulp basis) is required for softwoods and about 0.16% for hardwoods (13). In high consistency systems, alkali consumptions in this range are realized in mill scale bleaching. Medium consistency systems tend to consume slightly more alkali because the alkali concentration is lower at a given charge in these systems. The alkali is usually supplied in the form of oxidized white liquor, which can often be made available at lower cost than purchased sodium hydroxide. Unoxidized white liquor can be used (14), but there may be adverse effects on selectivity and rate. If unoxidized white liquor is used, oxygen consumption increases significantly.

Oxygen consumption is normally about 0.14% per unit of kappa number decrease for softwoods and 0.16% for hardwoods. Oxygen consumption in high consistency systems is slightly higher than medium consistency systems because of losses due to reactor venting and entrainment with the pulp leaving the reactor.

REACTION PRODUCTS:

Reaction products include organic acids and carbon dioxide, small amounts of carbon monoxide, and traces of methanol. Because the last two products are combustible, it is necessary to control their concentrations in the gas phase of high consistency reactors. This can be accomplished by continuously bleeding gas from the reactor or circulating it through a catalytic converter.

PROCESSES AND EQUIPMENT

The factors discussed above influence equipment design for oxygen delignification. In particular, the relevant reaction kinetics determine the size of the reactor, mass transfer considerations determine the configuration of the reactor and the provision for efficiently contacting the pulp with oxygen. The factor that most clearly distinguishes the available process types is pulp consistency, but most systems being installed currently are of the medium consistency type. High consistency system constituted the majority of those installed during first fifteen years of commercial oxygen bleaching, beginning in 1970. Both types have advantages and continue to be available, and a few high consistency systems have been installed recently. Typical operating data ranges for the two type are given in the table-1 (15, 16).

MEDIUM CONSISTENCY PROCESSES:

As already mentioned the low water solubility of oxygen makes it necessary to have a very large amount of water present to dissolve all the oxygen needed for delignification or use a smaller amount of water and provide the conditions that facilitate continual replacement of the dissolved oxygen by oxygen from the gas phase. This generally means providing a very large interfacial area between the liquid and gas phases. One way of doing this is to dewater the pulp to high (20-27%) consistency and fluff it to separate the fibers before the bleaching to create a dispersion of fiber in a continuous gas phase. When oxygen bleaching was first introduced, this was the only feasible method. Operating at medium consistency was not feasible because of the unavailability of equipment for intimately mixing a volume of gas with the pulp suspension (17). This problem was solved when medium consistency mixing technology became available in the late 1970s. The new high shear mixing devices make it possible to efficiently disperse oxygen as very small

TABLE-1

TYPICAL OPERATING DATAFOR OXYGEN DELIGNIFICATION OF KRAFT PULP

PARTICULARS	MEDIUM COSISTENCY	HIGH CONSISTENCY
Consistency, %	10 -14	25 – 28
Retention time, min	50 – 60	30
Initial temperature, °C	85 – 105	100 – 115
Inlet pressure, kPa	700 – 800	415 – 600
Outlet pressure, kPa	450 - 550	415 - 600
Delignification, %	40 – 45	45 – 55
Low press. steam con., kg/t	40 – 110	30 – 50
Medium press. steam con., kg/t	40 – 180	75 – 175
Power consumption, kwt/t	35 – 45	40 – 50
Alkali consumption, kg/t	18 – 28	18 – 23
Oxygen consumption, kg/t	20 - 24	15 - 24

Source: Dence and Reeve, Pulp Bleaching- Principle and Practice bubbles in 10-14 % consistency pulp. After the oxygen is dispersed, the pulp is transferred to a reactor, where the deliginification reaction allowed continuing. The dispersion is relatively stable and the buoyancy imparted by the trapped oxygen eliminates any tendency for bed compaction. An advantage of this approach is that no special dewatering equipment is needed before the oxygen stage.

Processes operating in the 10 – 14 % consistency range have been described by, among others, Kamyr, Imp co, Sunds Defibrator (18, 19) and Rauma Repola. A typical system is represented by the flow sheet in the figure 4. Partially washed brown stock is fed to a washer and the washed with the filtrate from the post oxygen washer. The pulp from the washer is charged with the caustic (NaOH) or oxidized white liquor, preheated in a low pressure steam mixer and fed to the high shear medium consistency mixer by medium consistency pump. Oxygen is added directly to the mixer from which the mixer of pulp, alkali, and oxygen gas passes to the bottom of the up flow reactor which, in some designs, is equipped with rotary distributor to prevent channeling and in others simply conical in shape. In either case, the reactor is designed with a higher aspect ratio than is typical of other types of bleaching towers to minimize the likelihood of channeling. Channeling tendency increases if the consistency decreases below 10%. Post-oxygen washer filtrate is used for the stock dilution at the top of the tower to facilitate discharge, which is mechanically assisted. The stock is brown to a blow tank through a separator, which allows entrained product gases, steam, and residual oxygen to be released.

To maximize environmental benefits and cost savings, it is normally desirable to remove as much lignin in the oxygen stage as possible, usually 40-60 %, without adversely affecting pulp quality. Achieving these levels in practice is often difficult, however, particularly in the medium consistency systems. Limitations on the degree to which all of the oxygen added can be dispersed as sufficiently small bubbles and coalescence of bubbles may contribute to a slowing of the reaction as the pulp progresses upward through the reactor. Attempts to increase delignification by increasing the oxygen charge may be counterproductive, the gas dispersing ability of the mixer decreases and the channeling becomes more likely as the volumetric fraction of gas in the pulp suspension is increased.

These considerations have led to the development of two stage medium consistency systems (20). These differ from the single stage systems only by the addition of a mixer and a second reactor between the first reactor and the blow tank, and often require no addition of oxygen or alkali between the stages. The first reactor is often smaller than the second.

HIGH-CONSISTENCY PROCESSES:

One approach to solving the three-phase mass transfer problem is to remove most of the free liquid phase. This has the two fold effect of providing a large gas liquid interfacial area and reducing the thickness of the liquid layer through which oxygen must diffuse to reach the fiber. Also, the amount of recycled, potentially oxidizable dissolved organic material in the reactor is reduced, as is the amount of water that must be heated to the reaction temperature.

For these reasons, many of the existing commercial installations operate at high consistency. Two general types are available: one Sapoxal process and other by MoDoChemetics and Imp co as the MoDo-CIL system. The MoDo-CIL process (21, 22) uses a vertical cylindrical reactor designed to contain a continuous pulp bed at a consistency of about 27 % as shown in the figure 5. Keeping the consistency at or slightly above this value prevents bed compaction, which would otherwise reduce the free volume needed to serve as an oxygen reservoir. Feeding is by a screw feeder or thick-stock pump, which maintains a seal against the reactor pressure. At the bottom of the reactor, the pulp is diluted to 5 % consistency with recycled oxygen-stage filtrate. Discharge is accomplished by a rotating conical screw and agitating arms.

Advantages claimed for operating at the lower consistencies that this arrangement permits are low bed combustibility and a lowering of the temperature rise associated with the reaction exotherm. Other advantages claimed for the tray type reactor are better control of retention time resulting from the absence of channeling and a more uniform gaseous atmosphere.

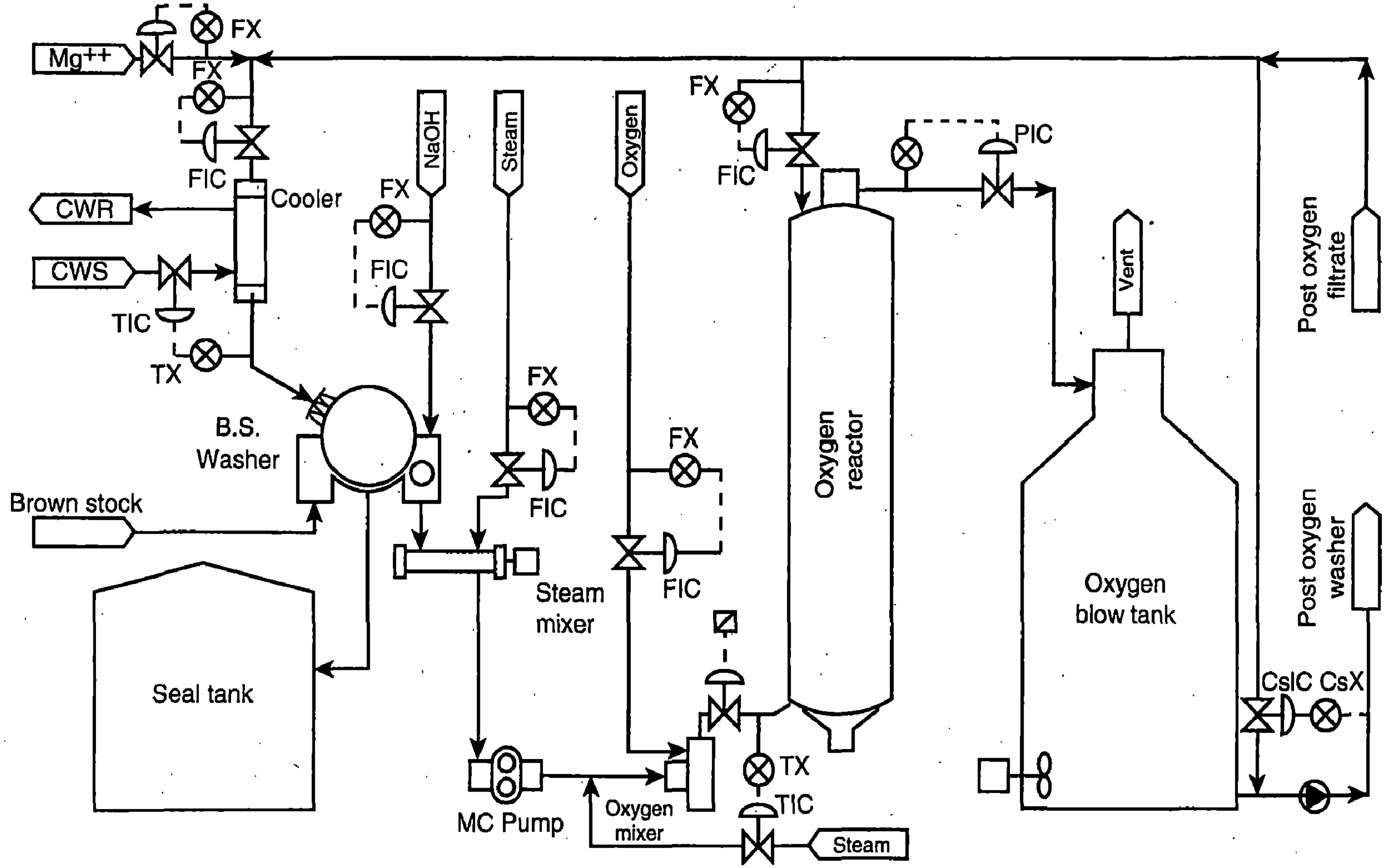


Fig-4: Flow sheet of medium consistency oxygen delignification

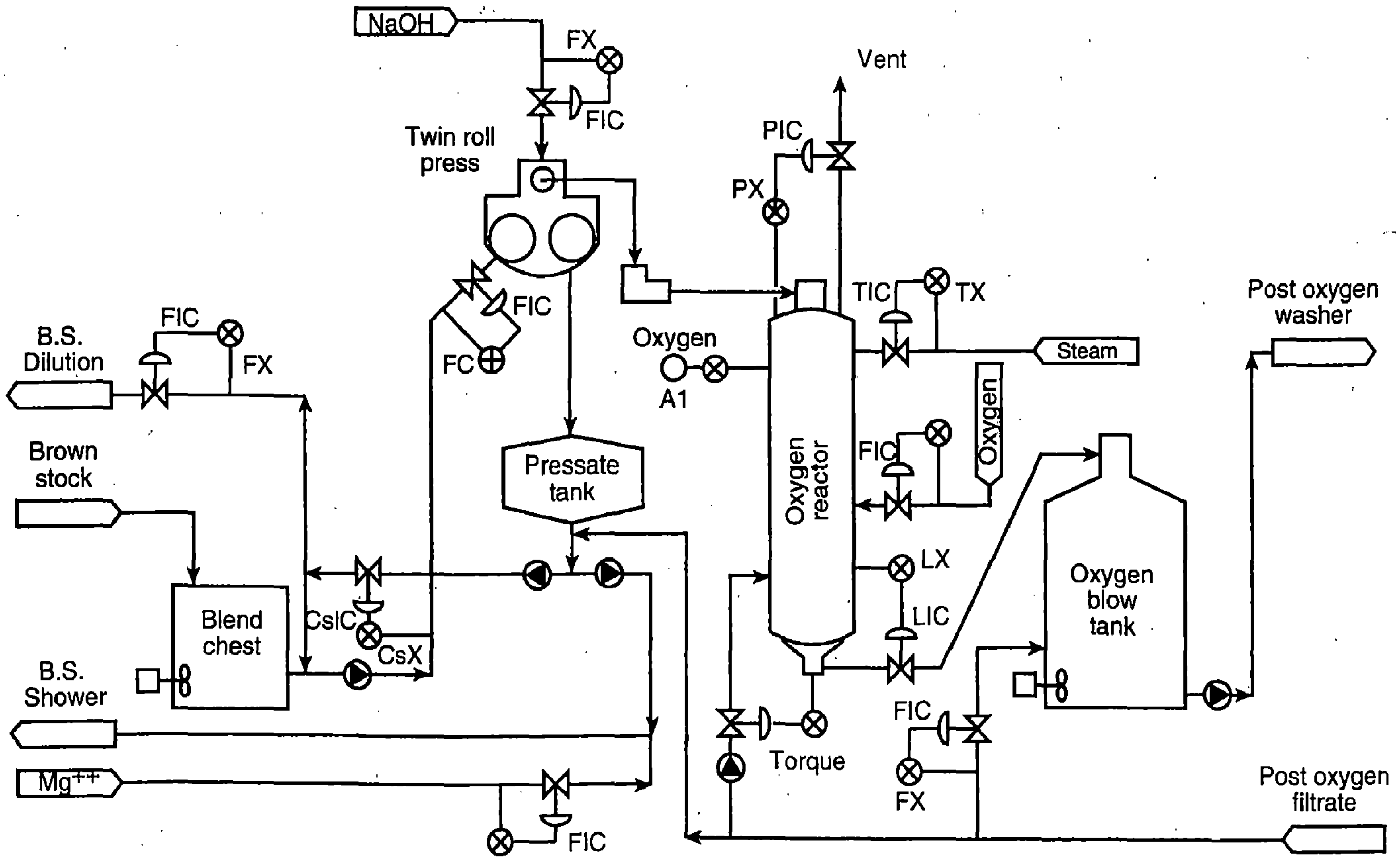


Fig-5: Flow sheet of high consistency oxygen delignification

COMPARISON OF MEDIUM AND HIGH CONSISTENCY SYSTEMS:

Both medium and high consistency oxygen delignification systems should be considered for any new installation. Their relative merits have been evaluated by several authors (23, 24).

The trends in the new installations are strongly towards the medium consistency process. Prior to 1983, virtually all installations were of high consistency type, but since 1983 virtually all have been medium consistency. According to a 1987 review, for example, of 19 new systems started up or slated for startup in the period 1985-1989, all but one were medium consistency. As of 1993, medium consistency systems accounted for 82 % of installed capacity. Among the reasons that may be cited for this trend are lower capital costs, greater ease of stock handling with medium consistency mixing and pumping technology, and improved selectivity in the presence of appreciable amounts of black liquor solids. On the other hand, chemical consumption is higher than in the high consistency systems and the extent of delignification trends to be lower.

The high consistency system still has its proponents, as evidenced by the fact that 700,000 annual tons of high consistency capacity was scheduled for startup in 1991. Those favoring the high consistency process claim that the advantages of the medium consistency approach are over started. The investment difference is mainly the result of a simpler feeding system for medium consistency and elimination of a press upstream of the reactor. However, to maintain the same level of pre-oxygen washing, the medium consistency system requires an extra washer to replace the press, lessening the advantage. Laboratory data indicate no difference in selectivity at the same level of magnesium addition (25). High consistency system often provide substantially more delignification and the chemical consumption difference may be significant. In summary. Both process types should be considered for any new installation.

PROCESS VARIABLES

The response of an oxygen bleaching systems to changes in process variables is a manifestation of the kinetics of the component chemical reactions and mass transfer processes. Kinetic studies such as those of Olm and Teder (26) therefore provide a useful framework for rationalizing and predicting these effects.

TIME AND TEMPERATURE:

At fixed alkali concentration, the decrease of kappa number with time exhibits two distinct stages, both of which are first order rate processes. There is an initial rapid kappa number drop followed by a slower one. This is interpreted as being caused by the presence of two types of lignin that differ in ease of removal. The two delignification stages are directly paralleled by two corresponding cellulose depolymerization stages.

A consequence of first order nature of the delignification process is that, given enough alkali, the kappa number continues to drop indefinitely. This is in contrast to the normal observation that the process appears to stop when a limiting kappa number is reached. The latter behavior results when the alkali charge is exhausted.The delignification is considerably accelerated by the temperature increases; the point of alkali exhaustion is reached much more rapidly at 130 °C than 85 °C.

As Figure 3 shows, the parallel between delignification and carbohydrate degradation rates leads to a good correction between viscosity and kappa number. This relationship tends to be moderately independent of the process variables although selectivity does suffer if the temperature is increased beyond about 120 °C or if all of the alkali is consumed.

ALKALI CHARGE:

Increasing the alkali concentration by increasing the alkali charge at constant consistency substantially accelerates both delignification and cellulose degradation.

OXYGEN PRESSURE:

Beyond a minimum value of about 4 atm, the effect of oxygen pressure is generally small in comparison to the effects of alkali charge and temperature. On the other hand, the kinetics studies already referred to show that oxygen pressure has an appreciable effect on the observed reaction rates at constant alkali concentration. The apparent departure of observed behavior from theoretical predictions is ascribable to the fact that, for economic reasons, alkali is normally present in limited amounts. In the absence of an excess of alkali, increase in oxygen pressure have relatively little effect.

CONSISTENCY:

Despite its far reaching equipment and process implications, the effect of consistency at fixed alkali charge is relatively small. Lowering the consistency leads to a moderate slowing down of both the delignification and carbohydrate degradation reactions as a result of the associated decrease in alkali concentration.

EFFECT ON MILL OPERATION:

The effects of installing an oxygen delignification stage extend far beyond the bleach plant. The operation and, in some cases, even the design of other equipment is affected in significant ways that must be considered before the change is made. The areas affected include screening, brown stock washing, chemical and heat recovery, downstream bleaching, stock preparation, and wastewater treatment.

PULP WASHING BEFORE AND AFTER THE OXYGEN STAGE:

If an oxygen stage is to be used for predelignification, it is most logically integrated into the brown stock washing system. Good upstream washing is essential, because black liquor solids consume oxygen and may adversely affect selectivity. Good downstream washing is necessary to fully realize the pollution abatement potential of the stage. Oxygen stage washer filtrate is used as wash liquor in the brown stock system and therefore finds its way back to the recovery system. Oxygen stage washer losses enter the chlorination stage and become unrecoverable.

The effects of pre- and post-oxygen washing efficiency have been the subject of several studies (28, 29). The dissolved solids entering the oxygen stage arise from two different sources, the digester and the oxygen stage, and the effects of each are dependent on the consistency of the oxygen stage. Oxygen stage solids find their way back into the oxygen stage because it is part of a closed countercurrent washing system. This is necessary to capitalize on the environmental benefits obtainable by preventing discharge of oxygen stage solids to the environment, directly or by way of the conventional part of the bleach plant. Filtrate from post oxygen washing, used to wash the pulp, partially displaces black liquor from the pulp before it enters the oxygen stage. Highly efficient post oxygen washing is desirable from the environmental standpoint but results in maximum recycle of oxygen stage solids. The desirability of efficient pre-oxygen washing is determined by the effect of high solids carryover into the oxygen stage, which is therefore of interest.

Greenwood (30) has reported data obtained by the laboratory bleaching of softwood Kraft pulp at 25% consistency in presence of systematically varied amounts of dissolved solids. At each level of black liquor addition, the oxygen stage was repeated four times. In each experiment, a fresh sample of pulp was diluted with filtrate from the previous experiment, simulating oxygen stage solids recirculation, and fresh black liquor was added, simulating carryover from the brown stock washers. Apparently, in the absence of recirculation of oxygen stage solids, carryover of as much as 60 kg/t black liquor solids has no effect on selectivity. On the other hand, combinations of the two types of dissolved solids, as would occur in practice, have a detrimental effect.

In medium consistency case, combinations of recirculated solids with up to 40 kg/t black liquor solids were tolerated without loss of selectivity, but higher levels had a detrimental effect. The higher tolerance of the medium consistency system for dissolved solids suggests that the important parameter is their concentration in the liquor, not the ratio of their dry weight to the dry pulp weight. To test this possibility, liquors were removed from the pulps after the oxygen stage in experiments conducted at both consistencies, with and without recirculation, at different levels of black liquor carryover.The results of determinations of their chemical oxygen demand (COD), a surrogate for total organic dissolved solids concentration, plotted against the viscosity loss in excess of that observed when fully washed pulp was bleached. A definite relationship is apparent, despite scatter in the data, which suggests that viscosity loss is incurred when the COD concentration is higher than 20g/L.

Greenwood (30) has performed calculations to demonstrate the practical implications of these phenomena, namely that highly efficient washing after the oxygen stage, essential for environment reasons, requires an increase in the efficiency of the brown stock washing system before the oxygen stage to keep the COD level in the oxygen stage below that at which viscosity is affected. In the ideal case, represented by perfectly efficient brown stock washing, the dissolved solids entering the oxygen stage consist only of recirculated oxygen stage solids. This emphasizes the importance of good washing before oxygen bleaching. Greenwood's calculations demonstrate the desirability of three washing stages after, the oxygen stage.

The current practice in high consistency system is to employ two washers (or in digester washing and one washer) before the press (31). In medium consistency system these are augmented by an additional washer to replace the press. Two washers are often employed after the oxygen stage but some systems have three.

SCREENING:

Placing the oxygen stage in the brown stock washing area makes it necessary to close the screen room water circuit to avoid the loss of oxygen stage solids. The resulting potential for foaming problems normally dictates the use of pressure screens.

The screen room may be located before or after the oxygen stage. Most mills screen the pulp before oxygen bleaching, but there are many examples of the screen room being located after the oxygen stage. Important factors to consider in locating the screen room are effects on cleanliness, heat balance, and capital costs. Leaving the knots in the pulp during medium consistency oxygen bleaching creates the risk of breaking them up to the mixer, thereby making them more difficult to remove later. Otherwise, there does not appear to be a major effect of screen room location on pulp cleanliness. Because screen room decker typically do not operate well above 80 °C, locating the screen room before the oxygen stage necessitates cooling of the oxygen stage filtrate before using it on the screen room washer, increasing overall steam consumption (30).

CHAPTER- 8

IMPLICATIONS FOR THE RECOVERY SYSTEM

Recycling of organic material dissolved in the oxygen stage to the recovery furnace typically increases the load on the furnace by 3% for softwoods and 2% for hardwoods. A further increase of about 1% results from improved recovery of black liquor solids which can be credited to the additional washing stages. The increase in the steam generation is somewhat less because of the lower heating value of oxidized solids (31).

If pure sodium hydroxide were used as alkali in the oxygen stage, the amount added with the chlorine dioxide generator spent acid, would probably exceed the requirement for sodium makeup to the liquor system. For this reason and because it is usually less expensive than purchased caustic, oxidized white liquor is usually used as a partial or sole alkali source. If the white liquor were used to satisfy the total alkali requirements of the oxygen stage, an increase of about 5% in the capacity of the caustic zing and lime reburning systems would be needed.

The additional water added with the alkali and as direct steam in the oxygen stage leads to additional evaporator steam requirements of about 4% when high consistency systems are used. The corresponding figure for medium consistency system is higher, perhaps as much as 10%, because of the higher steam usage in the medium consistency reactors. Alternatively, washing capacity can be added to allow a corresponding decrease in the dilution factor.

PULP QUALITY:

The strength equivalence of conventionally bleached pulps and pulps that have been oxygen delignified to an extent of about 50% before conventional bleaching is well-documented (31, 32, 33). Although the viscosity of the oxygen bleached pulps may be lower (as much as 4 mPa.s on the TAPPI T230 scale), their tear factor at a given tensile strength is generally the same as that of conventionally bleached pulps. The effects of the oxygen bleaching on pulp cleanliness, pitch, brightness stability, and refining energy requirement are the least neutral and in some cases slightly beneficial. The effect on cleanliness may be indirect in as much as oxygen bleaching is usually followed by short bleaching sequences consisting of high concentration chlorine dioxide stages, which are known to improve pulp cleanliness. Refining energy requirements have been observed to decease upon installation of an oxygen stage (34).

PART: II

DESIGN METHODOLOGY OF OXYGEN DELIGNIFICATION PLANT

1. SELECTION OF LOCATION AND EQUIPMENTS
2. BLOCK DIAGRAM OF OXYGEN DELIGNIFICATION PLANT WITH MATERIAL BALANCE
3. CALCULATIONS OF PUMP CAPACITY
4. CALCULATIONS OF PUMP HEAD, POWER AND EFFICIENCY
5. CALCULATIONS OF TANK DIMENSIONS
6. CALCULATIONS OF OXYGEN REACTORS AND POST OXYGEN WASHERS DIMENSIONS
7. CALCULATIONS OF PIPE LINE WEIGHT FOR COST DETERMENTION
8. CALCULATIONS OF TANK MATERIAL WEIGHT FOR COST DETERMINATION
9. CALCULATIONS OF HEAT EXCHANGER SPECIFICATION
10. CALCULATIONS OF POWER FOR AGITATORS
11. ESTIMATION OF PROJECT COST

DESIGN METHODOLOGY

SELECTION OF LOCATION:

Usually cooking and washing is followed by oxygen delignification and then bleaching. Effluents from oxygen delignification are used as wash media in brown stock washing and effluents from bleaching are generally sewered via effluent treatment.

SELECTION OF SYSTEM AND EQUIPMENTS:

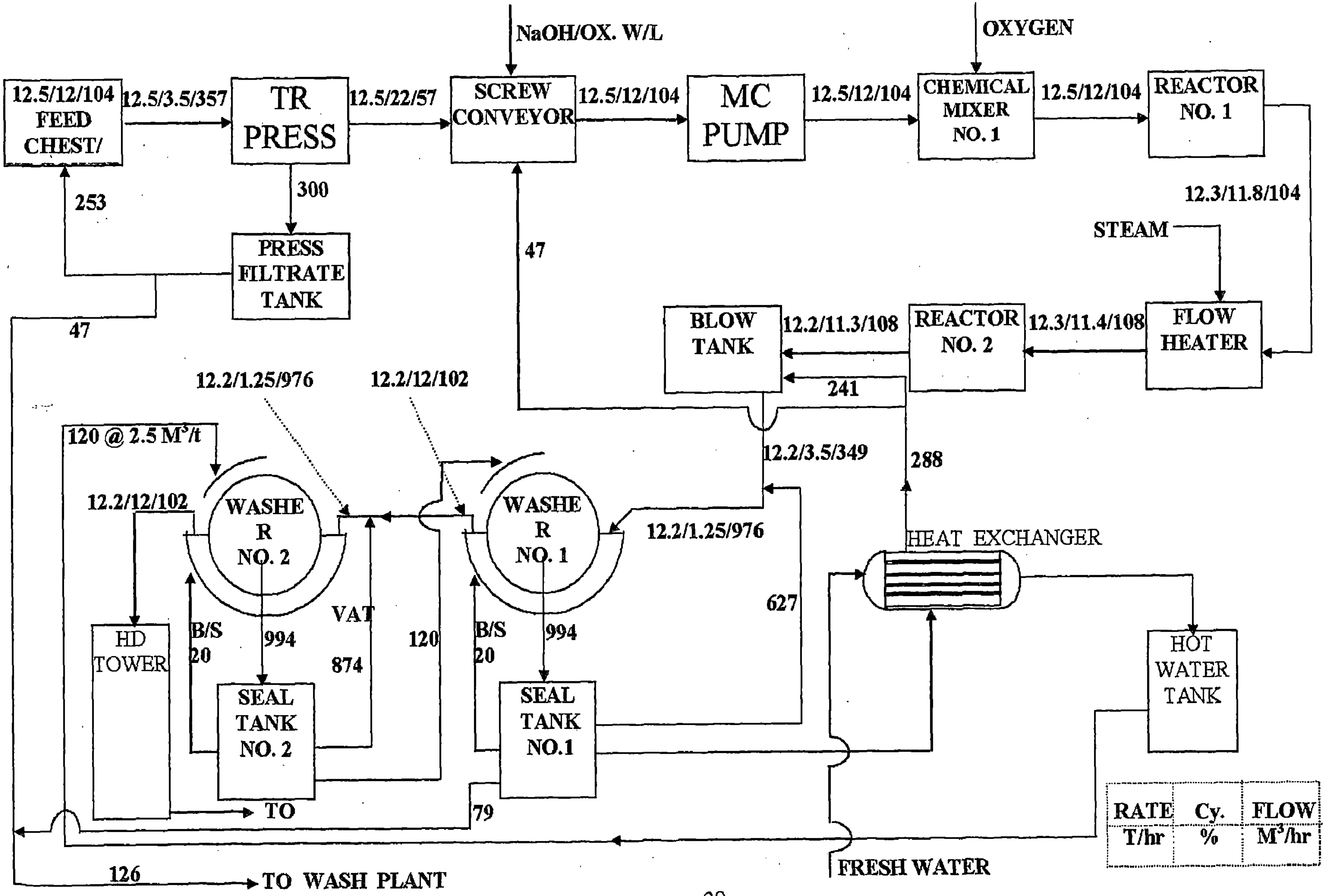
Medium consistency stages are in use today because of lower capital cost, ease of pulp transport and improved selectivity. The most beneficial oxygen delignification process uses two stages medium consistency system. The two stage system is used to preserve pulp strength. The first stage of a two stage system is a high pressure, high alkali, high temperature, short residence time so that large reduction in kappa number is there with modest drop in viscosity. The second stage using high temperature, longer residence time and lower chemical concentration to extend the delignification without drastically reducing the pulp strength.

The typical two stage system requires a medium consistency pump before the first stage and high shear mixers before each stage. To achieve acceptable washing results, we need two stages counter – current washing which is usually done on vacuum washers.

After washing and screening, we need a twin roll press so that unbleached pulp can be squeezed to remove max. organic and inorganic solids which would result in efficient oxygen delignification.

The block diagram of oxygen delignification plant with material balance is given figure 1.

FIGURE-1 **BLOCK DIAGRAM OF OXYGEN DELIGNIFICATION PLANT**



CALULATIONS OF PUMP CAPACITY:

On the basis of material balance shown in figure 1, we can calculate the pump capacity as done below:

1. **Washer No. 1 vat dilution pump:**

$$\{(98.75/1.25) - (96.5/3.5)\}12.2 = 627 \text{ m}^3/\text{ hr}$$

Adding 20 % more for design purpose

$$\text{Capacity} = 1.2 \times 627 = 752.9 \text{ m}^3/\text{ hr}$$

$$\approx 750 \text{ m}^3/\text{ hr}$$

2. **Washer No. 2 vat dilution pump :**

$$\{(98.75/1.25) - (88/12)\}\ 12.2 = 874.3 \text{ m}^3/\text{ hr}$$

Adding 20 % more for design purpose

$$\text{Capacity} = 1.2 \times 874.3 = 1049 \text{ m}^3/\text{ hr}$$

$$\approx 1100 \text{ m}^3/\text{ hr}$$

3. **Shower pump for both Washers:**

Capacity = 120 $\text{m}^3/$ hr as per material balance on the basis of 2.5 m^3/t dilution factor.

Adding 20 % more for design purpose

$$\text{Capacity} = 1.2 \times 120 = 144 \text{ m}^3/\text{ hr}$$

$$\approx 140 \text{ m}^3/\text{ hr}$$

4. **Back Shower pump for both Washers :**

Capacity = 20 $\text{m}^3/$ hr based on washer length.

5. **Blow Tank dilution pump :**

{(96.5/3.5) – (88/12)} 12.2 = 241 m^3/ hr

Adding 20 % more for design purpose

Capacity = 1.2 x 241 = 289 m^3/ hr

≈ 300 m^3/ hr

6. **Blow Tank transport pump :**

Capacity = 349 m^3/ hr as shown in material balance.

Adding 20 % more for design purpose

Capacity = 1.2 x 349 = 418.8 m^3/ hr

≈ 420 m^3/ hr

7. **Press Feed pump :**

Inlet consistency to press = 3.5 %

Rate of feed pulp = 12.5 T/hr

Total volume to be lifted = 12.5 x 0.035 = 357.1 m^3/ hr

Adding 20 % more for design purpose

Capacity = 1.2 x 357.1 = 428.6 m^3/ hr

≈ 430 m^3/ hr

8. **Feed Chest dilution pump :**

Feed consistency of pulp to chest = 12 %

Required inlet consistency to press = 3.5 %

Dilution required = { (96.5/3.5) – (88/12) }12.5 = 253 m^3/ hr

Adding 20 % more for design purpose

Capacity = 1.2 x 253 = 303 m^3/ hr

≈ 300 m^3/ hr

9. **Wash and Dilution pump :**

The total balance filtrate to be lifted to wash plant as per calculations shown in material balance = 47 + 79 = 126 m^3/ hr

Adding 20 % more for design purpose

Capacity = 1.2 x 126 = 151.2 m^3/ hr

≈ 150 m^3/ hr

10. **Medium Consistency pump :**

Capacity = 104 m^3/ hr as per material balance

Adding 20 % more for design purpose

Capacity = 1.2 x 104 = 124.8 m^3/ hr

≈ 125 m^3/ hr

CALCULATIONS OF PUMP HEAD, POWER AND EFFICIENCY:

Knowing the pump capacity, we can calculate the head, power and efficiency of the pump. Following are the pumps required as per material balance shown in figure 1.

Head calculations:

The basic equation used for computing the frictional head loss in the pipe lines is given by:

$$h_L = f.L.v^2.\rho / d.2g \quad \text{..............................} (1)$$

Where:

h_L = frictional head loss

f = friction factor to be found from the Reynolds number.

L = equivalent length of pipe in mt

d = internal diameter of pipe in mt

v = mean velocity of liquid in mt/sec

ρ = density of liquid in kg/ m^3

Also when flow is laminar and Re < 2000

Then f = 64/Re (for pipes)

and when flow is turbulent, f depends upon the Reynolds number and the relative roughness, k/d of the surface of the pipe wall, where k is the absolute roughness of the pipe surface.

In the figure -1, curves of friction factor, f versus Re are plotted to relative roughness k/d for ready reference. The value of k is taken from Table-1.

TABLE – 1: ROUGHNESS COEFFICIENT OF PIPE, k in mm

S.NO	PIPE MATERIAL	PIPE SURFACE CONDITION	k
1.	C.I. Pipes	i. New	0.2-0.3
		ii. For water supply-average	1-3
2.	i. Steel cast	i. New	0.03-0.05
	ii. Seamless	ii. Old/used	0.1-0.3
3.	Welded steel pipes	i. Water lines	0.4-1.2
		ii. Blast Fac. Gas	1-2
		iii. Coke oven gas	0.5
		iv. Compressed air and steam	0.2-0.4
4.	Cement lined	--------------------	0.03-0.1
5.	PE and PVC pipes	i. New	0.007
		ii. Old	0.003

Head loss in valves and fittings:

$$h_f = k_f . v^2/2g \text{ x numbers of valves/fittings.} \quad ......(2)$$

Where k_f = coeff. Of friction for valves/ fittings.

The value of k_f are taken from Table-2 and Table-3.

Frictional head loss in pipe lines:

Kinematic viscosity of the liquid $\gamma = \mu / \rho$

Reynolds number $Re = v.d/\gamma$

Also, k/d is calculated by taking value of k from the table.

From the graph, we can read the value of friction factor f

Therefore, frictional head loss in pipe lines is given by:

$$h = f. L. v^2. \rho / 2dg \quad ....(3)$$

Loss of head using equivalent length of pipe line L:

The equivalent length represents a length of straight pipe across which loss of head will be the same as the loss of head across the fittings. All the equivalent lengths of various fittings are added to the length of the straight pipe to obtain the total equivalent length of the piping systems. Standard equivalent lengths of pipe fittings and valves are given in Table-4 and Table-5.

Total head loss = head loss in valves and fittings + frictional head loss + equivalent length.

The total head will be sum of vertical height from ground floor and head losses in the pipe lines.

For design purpose, we take 30 % more of total head calculated as above.

TABLE-2: COEFFICIENT OF FRICTION FOR VALVES, K_f

S.NO	TYPE OF VALVE	K_f
1.	Gate valve	0.2
2.	Globe valve	3.8-4.0
3.	Diaphragm valve	1.5-2.0
4.	Angle valve	2.8-3.0
5.	Non return valve(vertical stem)	5.0-6.0
6.	Non return valve(inclined stem)	2.5-3.0
7.	Suction strainer	2.2-2.5

TABLE-3: COEFFICIENT K_f FOR PIPE FITTINGS

(K_f for 90° bends)

S.NO	PIPE LINE SIZE (mm)	K_f
1.	50	0.26
2.	100	0.23
3.	200	0.21
4.	300	0.19
5.	500	0.18

TABLE-5: EQUIVALENT LENGTH OF PIPE FITTINGS

NB mm	EQUIVALENT STRAIGHT LENGTH OF PIPE IN M FOR FLANGE							
	90° big bend	90° small bend	45° bend	180° bend	Equal Tee			Short piece
					→	↕	→	
25	0.20	0.25	0.10	0.25	0.20	0.50	0.70	0.50
32	0.25	0.35	0.15	0.35	0.25	0.70	0.90	0.75
40	0.30	0.45	0.20	0.45	0.30	0.85	1.2	0.90
50	0.45	0.60	0.25	0.60	0.45	1.2	1.6	1.2
65	0.55	0.75	0.30	0.75	0.55	1.4	2.0	1.5
80	0.70	0.95	0.40	0.95	0.70	1.9	2.6	1.9
100	0.95	1.3	0.55	1.3	0.95	2.6	3.7	2.7
125	1.3	1.8	0.70	1.7	1.3	3.5	4.9	3.6
150	1.6	2.2	0.90	2.2	1.6	4.3	6.1	4.4
200	2.2	3.1	1.2	3.0	2.2	6.0	8.3	6.2
250	2.8	3.8	1.5	3.8	2.8	7.6	10.6	7.7
300	3.6	5.0	2.0	5.0	3.6	9.8	13.6	10.1
350	3.8	5.5	2.2	5.4	3.9	10.6	14.8	11.1
400	4.7	6.6	2.7	6.5	4.7	12.8	17.8	13.3
450	5.5	7.6	3.0	7.7	5.5	15.0	21.0	15.0
500	6.1	8.4	3.4	8.3	6.1	16.6	23.0	17.0

TABLE -6: EQUIVALENT LENGTH OF VALVES

NB mm	EQUIVALENT STRAIGHT LENGTH OF PIPE IN M FOR FLANGE					
	Gate valve	Globe valve	Swing type Check valve	Lift type Check valve	Angle valve	Foot valve
25	0.10	6.8	1.4	9.9	2.6	15.0
32	0.15	9.6	2.0	14.0	3.6	21.0
40	0.20	12.0	2.5	17.0	4.4	26.0
50	0.30	16.0	3.4	23.0	6.0	35.0
65	0.35	20.0	4.2	29.0	7.5	43.0
80	0.45	26.0	5.5	38.0	9.8	56.0
100	0.60	36.0	7.7	53.0	14.0	79.0
125	0.80	47.0	10.0	69.0	18.0	104.0
150	1.0	60.0	12.0	86.0	22.0	129.0
200	1.4	83.0	17.0	121.0	31.0	181.0
250	1.6	103.0	22.0	150.0	39.0	225.0
300	2.2	135.0	28.0	197.0	51.0	296.0
350	2.6	147.0	31.0	215.0	56.0	323.0
400	3.1	176.0	37.0	257.0	67.0	286.0
450	3.5	200.0	44.0	290.0	77.0	450.0
500	3.8	225.0	50.0	330.0	87.0	510.0

Power calculations:

The required power $P = Q.\ H\ \ \rho / K_2 . \eta$ ....(4)

Where:

Q = capacity of pump, m^3/sec

H = head of pump, mt

ρ = density of liquid, gm/cc

K_2 = constant i.e. 367

η = efficiency of centrifugal pump , %

Add 30 % extra to power required for design purpose.

Efficiency of pump:

The efficiency of pump is taken from the table 6.

TABLE-6: EFFICIENCY OF PUMPS

S.NO	STOCK CONSISTENCY,%	PUMP EFFICIENCY,%
1.	Up to 2 %	75
2.	Up to 6 %	50 – 60
3.	Above 6 %	40 – 50
4.	Centrifugal pumps	75 – 90
5.	Piston pumps	60 - 80

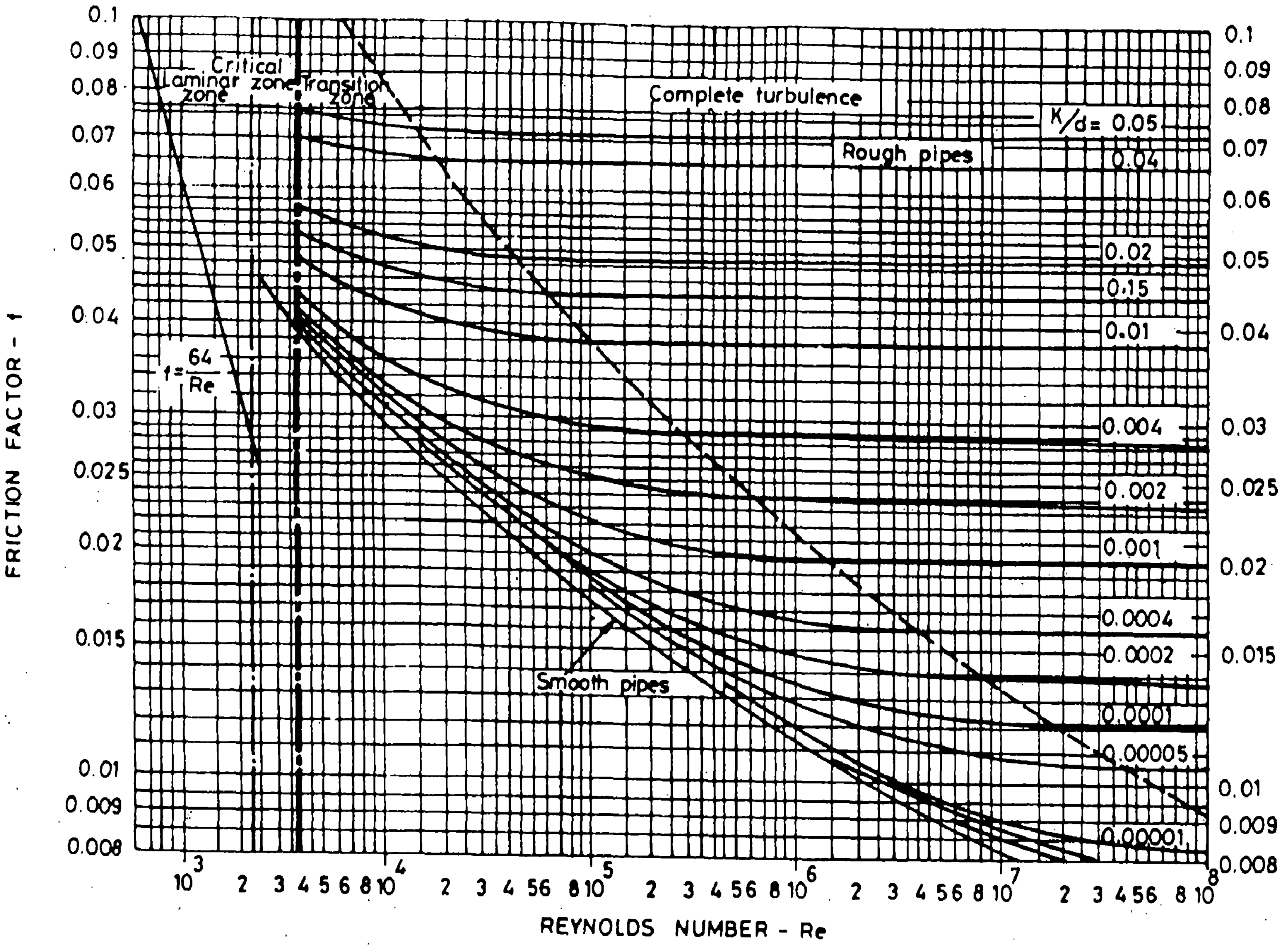


Fig-2: Friction factor for commercial pipes

1. **Washer No.1 vat dilution pump :**

Capacity = 750 m^3/hr as per material balance.

Line size = 350 mm

Velocity of liquid V = Q/A

$= 750/ 3600 \times \pi/4(0.35)^2$

= 2.17 mt/sec.

Total numbers of valves in washer no.1 vat dilution line are:

1. no. of gate valves = 2	$k_f = 0.2$
2. no. of non return valves = 1	$k_f = 2.5$
3. no. of control valve = 1	$k_f = 1.5$

(the values of k_f are taken from table)

Head loss in valves:

$h_1 = \{0.2\,(2.17)^2/2x9.8\}2 = 0.10$ mwc

$h_2 = \{2.5\,(2.17)^2/2x9.8\}1 = 0.60$ mwc

$h_3 = \{1.5\,(2.17)^2/2x9.8\}\ 1 = 0.36$ mwc

Total head loss in valves = 0.10 + 0.60 + 0.36 = 1.06 mwc

Head loss in Fittings:

No. of bends = 6 $k_f = 0.19$ from the table

$h = \{0.19 \times (2.17)^2 / 2x\ 9.8\}\ 6 = 0.27$ mwc

Frictional head loss in pipe lines:

Here, $\rho = 1050\ kg/m^3$

Dynamic viscosity (μ) of the liquid = 20 cp.

= 0.20 poise (gm/cm-sec)

Therefore, Kinematic viscosity of liquid $\gamma = \mu / \rho$

$= 0.20$ gm/cm-sec / 1.05 gm/cm^3

$= 1.905 \times 10^{-5}\ m^2/sec$

Reynolds number Re = v. d / γ

$Re = 2.17 \times 0.35 / 1.905 \times 10^{-5}$

$$Re = 3.99 \times 10^4$$

Also, k/d = 0.04 / 350 = 0.0001

(The value of k is taken from table)

From the graph, f = 0.028

Equivalent length of pipe line L = 35m + 6x3.8m + 2x2.6m + 1x215m + 1x147m

L = 425 mt.

Frictional head loss in pipe lines $h = 0.028\ (425/0.35)\ \{(2.17)^2 \times 1050 / 2x9.8\}$

h = 8576.92 mmwc

h = 8.58 mwc.

Total head loss = 1.06 + 0.27 + 8.58 = 9.91 mwc

The vertical height from the ground level = 15 mt

Therefore, the pump head required = 15 + 9.91 = 24.91 mwc

Taking 30 % extra for design purpose = 1.3 x 24.91 = 32.4 mwc

≈ 35 mwc.

Assuming efficiency of centrifugal pump = 80 %

The Power required P = 750 x 35 x 1.05 / 367 x 0.80

= 93.87 kW

Taking 30 % extra for design purpose = 1.3 x 93.87

= 122.04 kW

P ≈ 122 Kw

2. Washer No.2 vat dilution pump :

Capacity = 1100 m^3/hr as per material balance.

Line size = 450 mm

Velocity of liquid V = Q/A

$$= 1100 / 3600 \times \pi/4(0.45)^2$$

= 1.92 mt/sec.

Total number of valves in washer no.2 vat dilution line are:

1. no. of gate valves = 2	$k_f = 0.2$
2. no. of non return valves = 1	$k_f = 2.5$
3. no. of control valve = 1	$k_f = 1.5$

(the values of k_f are taken from table)

Head loss in valves:

$h_1 = \{0.2\,(1.92)^2/2x9.8\}2 = 0.07$ mwc

$h_2 = \{2.5\,(1.92)^2/2x9.8\}1 = 0.47$ mwc

$h_3 = \{1.5\,(1.92)^2/2x9.8\}\;1 = 0.28$ mwc

Total head loss in valves = 0.07 + 0.47 + 0.28 = 0.82 mwc

Head loss in Fittings:

No. of bends = 5 $k_f = 0.18$ from the table

$h = \{0.18 \times (1.92)^2 / 2 \times 9.8\}\; 5 = 0.17$ mwc

Frictional head loss in pipe lines:

Here, $\rho = 1050\ kg/m^3$

Dynamic viscosity (μ) of the liquid = 20 cp.

= 0.20 poise (gm/cm-sec)

Therefore, Kinematic viscosity of liquid $\gamma = \mu / \rho$

$= 0.20\ gm/cm\text{-}sec / 1.05\ gm/cm^3$

$= 1.905 \times 10^{-5}\ m^2/sec$

Reynolds number $Re = v.\ d / \gamma$

$Re = 1.92 \times 0.45 / 1.905 \times 10^{-5}$

$Re = 4.53 \times 10^4$

Also, k/d = 0.04 / 450 = 0.00009

(the value of k is taken from table)

From the graph, f = 0.03

Equivalent length of pipe line L = 35m + 5x5.5m + 2x3.5m + 1x290m + 1x200m

L = 559.5 mt.

Frictional head loss in pipe lines h = 0.03 (559.5/0.45) $\{(1.92)^2$ x 1050 / 2x9.8$\}$

h = 7366.2 mmwc

h = 7.37 mwc.

Total head loss = 0.82 + 0.17 + 7.37 = 8.36 mwc

The vertical height from the ground level = 15 mt

Therefore, the pump head required = 15 + 8.36 = 23.36 mwc

Taking 30 % extra for design purpose = 1.3 x 23.36 = 30.4 mwc

≈ 30 mwc.

Assuming efficiency of centrifugal pump = 80 %

The Power required P = 1100 x 30 x 1.05 / 367 x 0.80

= 118.02 kW

Taking 30 % extra for design purpose = 1.3 x 118.02

= 153.4 kW

P ≈ 153 kW

3. **Shower pump for both washers :**

Capacity = 140 m^3/hr as per material balance.

Line size = 200 mm

Velocity of liquid V = Q/A

= 140/ 3600 x $\pi/4(0.20)^2$

= 1.24 mt/sec.

Total numbers of valves in shower pump line are:

1. no. of gate valves = 2 (200 mm) $k_f = 0.2$
2. no. of gate valves = 4 (50 mm) $k_f = 0.2$
3. no. of control valve = 1 (200 mm) $k_f = 1.5$

(the values of k_f are taken from table)

Head loss in valves:

$h_1 = \{0.2\,(1.24)^2/2x9.8\}6 = 0.09$ mwc

$h_2 = \{1.5\,(1.24)^2/2x9.8\}\ 1 = 0.12$ mwc

Total head loss in valves = 0.09 + 0.12 = 0.21 mwc

Head loss in Fittings:

No. of bends = 10 (200 mm) $k_f = 0.21$ from the table

No. of bends = 4 (50 mm) $k_f = 0.26$ from the table

Flow meter = 1 $k_f = 10$ from the table

$h_1 = \{0.21 \times (1.24)^2 / 2x\,9.8\}\ 10 = 0.16$ mwc

$h_2 = \{0.26 \times (1.24)^2 / 2x\,9.8\}\ 4 = 0.08$ mwc

$h_3 = \{10 \times (1.24)^2 / 2x\,9.8\}\ 1 = 0.78$ mwc

$h = 0.16 + 0.08 + 0.78 = 1.02$ mwc

Frictional head loss in pipe lines:

Here, $\rho = 1050$ kg/m^3

Dynamic viscosity (μ) of the liquid = 20 cp.

= 0.20 poise (gm/cm-sec)

Therefore, Kinematic viscosity of liquid $\gamma = \mu / \rho$

= 0.20 gm/cm-sec / 1.05 gm/cm^3

= 1.905×10^{-5} m^2/sec

Reynolds number Re = v. d / γ

$Re = 1.24 \times 0.20 / 1.905 \times 10^{-5}$

$Re = 1.302 \times 10^4$

Also, k/d = 0.04 / 200 = 0.0002

(the value of k is taken from table)

From the graph, f = 0.04

Equivalent length of pipe line L = 70m + 2x1.4m + 4x0.3m + 1x121m + 10x3.1m + 4x0.60 + 1x10

L = 238.4 mt.

Frictional head loss in pipe lines h = 0.04 (238.4/0.20) {$(1.24)^2$ x 1050 / 2x9.8}

h = 3927.47 mmwc

h = 3.93 mwc.

Total head loss = 0.21 + 1.02 + 3.93 = 5.16 mwc

The vertical height from the ground level = 15 mt

Therefore, the pump head required = 15 + 5.16 = 20.16 mwc

Taking 30 % extra for design purpose = 1.3 x 20.16 = 26.22 mwc

≈ 30 mwc.

Assuming efficiency of centrifugal pump = 80 %

The Power required P = 140 x 30 x 1.05 / 367 x 0.80

= 15.02 kW

Taking 30 % extra for design purpose = 1.3 x 15.02

= 19.53 kW

P ≈ 20 kW

4. **Back shower pump for both washers :**

Capacity = 20 m^3/hr as per length of washer.

Line size = 75 mm

Velocity of liquid V = Q/A

= 20/ 3600 x π/4$(0.075)^2$

= 1.26 mt/sec.

Total numbers of valves in back shower pump line are:

1. no. of gate valves = 2 (75 mm) k_f = 0.2

(the values of k_f is taken from table)

Head loss in valves:

h_1 = {0.2 $(1.26)^2$/2x9.8}2 = 0.032 mwc

Head loss in Fittings:

No. of bends = 6 (75 mm) k_f = 0.25 from the table

$h_1 = \{0.25 \times (1.26)^2 / 2 \times 9.8\}\ 6 = 0.121$ mwc

Frictional head loss in pipe lines:

Here, $\rho = 1050\ kg/m^3$

Dynamic viscosity (μ) of the liquid = 20 cp.

= 0.20 poise (gm/cm-sec)

Therefore, Kinematic viscosity of liquid $\gamma = \mu / \rho$

$= 0.20\ gm/cm\text{-}sec / 1.05\ gm/cm^3$

$= 1.905 \times 10^{-5}\ m^2/sec$

Reynolds number $Re = v \,.\, d / \gamma$

$Re = 1.26 \times 0.075 / 1.905 \times 10^{-5}$

$Re = 4.96 \times 10^3$

Also, k/d = 0.04 / 75 = 0.0005

(the value of k is taken from table)

From the graph, f = 0.042

Equivalent length of pipe line L = 70m + 2x0.45m + 6x0.95m

L = 76.6 mt.

Frictional head loss in pipe lines $h = 0.042\ (76.6/0.075)\ \{(1.26)^2 \times 1050 / 2 \times 9.8\}$

h = 3648.3 mmwc

h = 3.65 mwc.

The vertical height from the ground level = 15 mt

Therefore, the pump head required = 15 + 0.12 + 3.65 = 18.77 mwc

Taking 30 % extra for design purpose = 1.3 x 18.77 = 24.24 mwc

≈ 25 mwc.

Since we require 7 kg/m^3 pressure at back shower of washer

So, we need a pump of 70 mwc

Assuming efficiency of centrifugal pump = 60 %

The Power required P = 20 x 70 x 1.05 / 367 x 0.60

= 6.68 kW

Taking 50 % extra for design purpose = 1.5 x 6.68

= 10.02 kW

$P \approx 10$ kW

5. **Blow tank dilution pump :**

Capacity = 300 m^3/hr as per material balance.

Line size = 250 mm

Velocity of liquid V = Q/A

$= 300/\ 3600 \times \pi/4(0.25)^2$

= 1.7 mt/sec.

Total numbers of valves in blow tank dilution line are:

1. no. of gate valves = 2 $k_f = 0.2$
2. no. of non return valves = 1 $k_f = 2.5$
3. no. of control valve = 1 $k_f = 1.5$

(the values of k_f are taken from table)

Head loss in valves:

$h_1 = \{0.2\ (1.7)^2/2x9.8\}2 = 0.06$ mwc

$h_2 = \{2.5\ (1.7)^2/2x9.8\}1 = 0.37$ mwc

$h_3 = \{1.5\ (1.7)^2/2x9.8\}\ 1 = 0.22$ mwc

Total head loss in valves = 0.06 + 0.37 + 0.22 = 0.65 mwc

Head loss in Fittings:

No. of bends = 6 k_f = 0.20 from the table

No. of Tee = 1 k_f = 3.0 from the table

$h_1 = \{0.2 \times (1.7)^2\ /\ 2x\ 9.8\}\ 6 = 0.18$ mwc

$h_2 = \{3.0 \times (1.7)^2 / 2 \times 9.8\}\ 1 = 0.44$ mwc

$h = 0.18 + 0.44 = 0.62$ mwc

Frictional head loss in pipe lines:

Here, $\rho = 1050$ kg/m^3

Dynamic viscosity (μ) of the liquid = 20 cp.

= 0.20 poise (gm/cm-sec)

Therefore, Kinematic viscosity of liquid $\gamma = \mu / \rho$

$= 0.20$ gm/cm-sec / 1.05 gm/cm^3

$= 1.905 \times 10^{-5}$ m^2/sec

Reynolds number $Re = v.\ d / \gamma$

$Re = 1.7 \times 0.25 / 1.905 \times 10^{-5}$

$Re = 2.23 \times 10^4$

Also, $k/d = 0.04 / 250 = 0.00016$

(the value of k is taken from table)

From the graph, $f = 0.03$

Equivalent length of pipe line L = 50m + 6x2.8m + 1x7.6m + 2x1.6m + 1x150m + 1x103m

L = 329 mt.

Frictional head loss in pipe lines $h = 0.03\ (329/0.25)\ \{(1.7)^2 \times 1050 / 2 \times 9.8\}$

$h = 6112.35$ mmwc

$h = 6.11$ mwc.

Total head loss = 0.65 + 0.62 + 6.11 = 7.38 mwc

The vertical height from the ground level = 15 mt

Therefore, the pump head required = 15 + 7.38 = 22.38 mwc

Taking 30 % extra for design purpose = 1.3 x 22.38 = 29.09 mwc

≈ 30 mwc.

Assuming efficiency of centrifugal pump = 80 %

The Power required P = 300 x 30 x 1.05 / 367 x 0.80

= 32.19 kW

Taking 30 % extra for design purpose = 1.3 x 32.19

= 41.8 kW

$P \approx 42$ kW

6. **Blow tank transport pump :**

Capacity = 420 m^3/hr as per material balance.

Line size = 250 mm

Velocity of liquid V = Q/A

$= 420/\ 3600 \times \pi/4(0.25)^2$

= 2.38 mt/sec.

Total numbers of valves in blow tank transport line are:

1. no. of gate valves = 2 $k_f = 0.2$
2. no. of control valve = 1 $k_f = 1.5$

(the values of k_f are taken from table)

Head loss in valves:

$h_1 = \{0.2\ (2.38)^2/2\text{x}9.8\}2 = 0.12$ mwc

$h_2 = \{1.5\ (2.38)^2/2\text{x}9.8\}\ 1 = 0.43$ mwc

Total head loss in valves = 0.12 + 0.43 = 0.55 mwc

Head loss in Fittings:

No. of bends = 6 $k_f = 0.20$ from the table

$h = \{0.2 \times (2.38)^2 / 2\text{x}\ 9.8\}\ 6 = 0.35$ mwc

Frictional head loss in pipe lines :

Here, $\rho = 1050$ kg/m^3

Dynamic viscosity (μ) of the liquid = 20 cp.

= 0.20 poise (gm/cm-sec)

Therefore, Kinematic viscosity of liquid $\gamma = \mu / \rho$

$= 0.20$ gm/cm-sec / 1.05 gm/cm^3

$= 1.905 \times 10^{-5}$ m^2/sec

Reynolds number $Re = v. d / \gamma$

$Re = 2.38 \times 0.25 / 1.905 \times 10^{-5}$

$Re = 3.12 \times 10^4$

Also, $k/d = 0.04 / 250 = 0.00016$

(the value of k is taken from table)

From the graph, $f = 0.03$

Equivalent length of pipe line L = 35m + 6x5.5m + 2x3.5m +1x200m

L = 271.5 mt.

Frictional head loss in pipe lines $h = 0.03\ (271.5/0.25)\ \{(2.38)^2 \times 1050 / 2\times9.8\}$

h = 9886.4 mmwc

h = 9.89 mwc.

Total head loss = 0.55 + 0.35 + 9.89 = 10.79 mwc

The vertical height from the ground level = 15 mt

Therefore, the pump head required = 15 + 10.79 = 25.79 mwc

Taking 30 % extra for design purpose = 1.3 x 25.79 = 33.53 mwc

≈ 35 mwc.

Assuming efficiency of centrifugal pump = 60 %

The Power required P = 300 x 30 x 1.05 / 367 x 0.60

= 70.1 kW

Taking 30 % extra for design purpose = 1.3 x 70.1

= 91.2 kW

P ≈ 90 kW

7. **Press Feed pump :**

Capacity = 430 m^3/hr as per material balance.

Line size = 300 mm

Velocity of liquid V = Q/A

$= 430/ 3600 \times \pi/4(0.30)^2$

= 1.69 mt/sec.

Total number of valves in press feed pump line are:

1. no. of gate valves = 2 $k_f = 0.2$
2. no. of control valve = 1 $k_f = 1.5$

(the values of k_f are taken from table)

Head loss in valves:

$h_1 = \{0.2\,(1.69)^2/2x9.8\}2 = 0.06$ mwc

$h_2 = \{1.5\,(1.69)^2/2x9.8\}\ 1 = 0.23$ mwc

Total head loss in valves = 0.06 + 0.23 = 0.29 mwc

Head loss in Fittings:

No. of bends = 12	$k_f = 0.19$ from the table
No. flow meter = 1	$k_f = 10$ from the table

$h_1 = \{0.19 \times (1.69)^2 / 2 \times 9.8\}\ 12 = 0.33$ mwc

$h_2 = \{10 \times (1.69)^2 / 2 \times 9.8\}\ 1 = 1.46$ mwc

h = 0.33 + 1.46 = 1.79 mwc

Frictional head loss in pipe lines:

Here, $\rho = 1050$ kg/m^3

Dynamic viscosity (μ) of the liquid = 20 cp.

= 0.20 poise (gm/cm-sec)

Therefore, Kinematic viscosity of liquid $\gamma = \mu / \rho$

= 0.20 gm/cm-sec / 1.05 gm/cm^3

$= 1.905 \times 10^{-5}$ m^2/sec

Reynolds number Re = v. d / γ

$$Re = 1.69 \times 0.3 / 1.905 \times 10^{-5}$$

$$Re = 2.67 \times 10^{4}$$

Also, k/d = 0.04 / 300 = 0.0001

(the value of k is taken from table)

From the graph, f = 0.025

Equivalent length of pipe line L = 150m + 12x3.6m + 2x2.2m + 1x51m + 1x10m

L = 258.6 mt.

Frictional head loss in pipe lines h = 0.025 (258.6/0.30) $\{(1.69)^2 \times 1050 / 2x9.8\}$

h = 3297.26 mmwc

h = 3.29 mwc.

Total head loss = 0.29 + 1.79 + 3.29 = 5.57 mwc

The vertical height from the ground level = 15 mt

Therefore, the pump head required = 15 + 5.57 = 20.37 mwc

Since this pump has to work against a press vat pressure of 1 kg/m^2 i.e. 10.33 mwc

So, net head becomes = 20.37 + 10.33 = 30.70 mwc

Taking 30 % extra for design purpose = 1.3 x 30.70 = 39.91 mwc

≈ 40 mwc.

Assuming efficiency of centrifugal pump = 60 %

The Power required P = 300 x 30 x 1.05 / 367 x 0.60

= 82.02 kW

Taking 30 % extra for design purpose = 1.3 x 82.02

= 106.6 kW

P ≈ 107 kW

8. **Feed chest dilution pump:**

Capacity = 300 m^3/hr as per material balance.

Line size = 250 mm

Velocity of liquid V = Q/A

$= 300/\ 3600 \times \pi/4(0.25)^2$

$= 1.7$ mt/sec.

Total numbers of valves in blow tank dilution line are:

1. no. of gate valves = 2	$k_f = 0.2$
2. no. of non return valves = 1	$k_f = 2.5$
3. no. of control valve = 1	$k_f = 1.5$

(the values of k_f are taken from table)

Head loss in valves:

$h_1 = \{0.2\,(1.7)^2/2\text{x}9.8\}2 = 0.06$ mwc

$h_2 = \{2.5\,(1.7)^2/2\text{x}9.8\}1 = 0.37$ mwc

$h_3 = \{1.5\,(1.7)^2/2\text{x}9.8\}\ 1 = 0.22$ mwc

Total head loss in valves = 0.06 + 0.37 + 0.22 = 0.65 mwc

Head loss in Fittings:

No. of bends = 12	$k_f = 0.20$ from the table
No. of flow meter = 1	$k_f = 10$ from the table

$h_1 = \{0.2 \times (1.7)^2 / 2\text{x}\ 9.8\}\ 12 = 0.35$ mwc

$h_2 = \{10 \times (1.7)^2 / 2\text{x}\ 9.8\}\ 1 = 1.47$ mwc

h = 0.35 + 1.47 = 1.82 mwc

Frictional head loss in pipe lines:

Here, $\rho = 1050$ kg/m^3

Dynamic viscosity (μ) of the liquid = 20 cp.

= 0.20 poise (gm/cm-sec)

Therefore, Kinematics viscosity of liquid $\gamma = \mu / \rho$

= 0.20 gm/cm-sec / 1.05 gm/cm^3

$= 1.905 \times 10^{-5}$ m^2/sec

Reynolds number Re = v. d / γ

Re = 1.7 x 0.25 / 1.905 x 10^{-5}

Re = 2.23 x 10^4

Also, k/d = 0.04 / 250 = 0.0001

(the value of k is taken from table)

From the graph, f = 0.03

Equivalent length of pipe line L = 150m + 12x2.8m + 2x1.6m + 1x150m + 1x103m

L = 439.8 mt.

Frictional head loss in pipe lines h = 0.03 (439.8/0.25) {$(1.7)^2$ x 1050 / 2x9.8}

h = 8170.85 mmwc

h = 8.17 mwc.

Total head loss = 0.65 + 1.82 + 8.17 = 10.64 mwc

The vertical height from the ground level = 15 mt

Therefore, the pump head required = 15 + 10.64 = 25.64 mwc

Taking 30 % extra for design purpose = 1.3 x 25.64 = 33.33 mwc

≈ 35 mwc.

Assuming efficiency of centrifugal pump = 80 %

The Power required P = 300 x 35 x 1.05 / 367 x 0.80

= 37.55 kW

Taking 30 % extra for design purpose = 1.3 x 37.55

= 48.82 kW

P ≈ 50 kW

9. Wash and Dilution pump :

Capacity = 150 m^3/hr as per material balance.

Line size = 200 mm

Velocity of liquid $V = Q/A$

$= 150/ 3600 \times \pi/4(0.20)^2$

$= 1.33$ mt/sec.

Total number of valves in press feed pump line are:

1. no. of gate valves = 2 $k_f = 0.2$
2. no. of control valve = 1 $k_f = 1.5$

(the values of k_f are taken from table)

Head loss in valves:

$h_1 = \{0.2\,(1.33)^2/2x9.8\}2 = 0.04$ mwc

$h_2 = \{1.5\,(1.33)^2/2x9.8\}\ 1 = 0.14$ mwc

Total head loss in valves = 0.04 + 0.14 = 0.18 mwc

Head loss in Fittings:

No. of bends = 15	$k_f = 0.21$ from the table
No. flow meter = 1	$k_f = 10$ from the table

$h_1 = \{0.21 \times (1.33)^2 / 2 \times 9.8\}\ 15 = 0.28$ mwc

$h_2 = \{10 \times (1.33)^2 / 2 \times 9.8\}\ 1 = 0.90$ mwc

$h = 0.28 + 0.90 = 1.18$ mwc

Frictional head loss in pipe lines:

Here, $\rho = 1050\ kg/m^3$

Dynamic viscosity (μ) of the liquid = 20 cp.

= 0.20 poise (gm/cm-sec)

Therefore, Kinematic viscosity of liquid $\gamma = \mu / \rho$

$= 0.20$ gm/cm-sec / 1.05 gm/cm^3

$= 1.905 \times 10^{-5}\ m^2/sec$

Reynolds number $Re = v.\ d / \gamma$

$Re = 1.33 \times 0.2 / 1.905 \times 10^{-5}$

$Re = 1.39 \times 10^4$

Also, k/d = 0.04 / 200 = 0.0002

(the value of k is taken from table)

From the graph, f = 0.04

Equivalent length of pipe line L = 200m + 10x3.1m + 2x1.4m + 1x121m + 1x10m

L = 364.8 mt.

Frictional head loss in pipe lines

$h = 0.04\ (364.8/0.20)\ \{(1.33)^2 \times 1050\ /\ 2\text{x}9.8\}$

h = 6913.87 mmwc

h = 6.91 mwc.

Total head loss = 0.18 + 1.18 + 6.91 = 8.27 mwc

The vertical height from the ground level = 15 mt

Therefore, the pump head required = 15 + 8.27 = 23.27 mwc

Taking 30 % extra for design purpose = 1.3 x 23.27 = 30.25 mwc

≈ 30 mwc.

Assuming efficiency of centrifugal pump = 80 %

The Power required P = 150 x 30 x 1.05 / 367 x 0.80

= 16.09 kW

Taking 30 % extra for design purpose = 1.3 x 16.09

= 20.92 kW

P ≈ 21 kW

10. Medium consistency (MC) pump :

Capacity = 125 m^3/hr as per material balance.

Line size = 300 mm

Velocity of liquid V = Q/A

$= 125/\ 3600 \times \pi/4(0.30)^2$

= 0.49 mt/sec.

The head loss in valves and fittings are negligible

And, the max. height of the reactor = 23.7 mt

The max. reactor top pressure = 9 kg/m^2 = 9 x 10.33 mwc

= 92.97 mwc

Total head = 23.7 + 92.97 = 116.67 mwc

≈ 117 mwc

Assuming pump efficiency = 40 %

The Power required P = 125 x 117 x 1.05 / 367 x 0.40

= 104.61 kW

Taking 30 % extra for design purpose = 1.3 x 104.61

= 135.99 kW

P ≈ 136 kW

TABLE 7: SPECIFICATION OF PUMPS

S.No	NAME OF PUMP	CAPACITY (m^3/hr)	LINE DIA. (mm)	HEAD mwc	PUMP EFF. (%)	MOTOR POWER (kw)
1.	Washer No.1 Vat dil. pump	750	350	35	80	122
2.	Washer No.2 Vat dil. pump	1100	450	30	80	153
3.	Shower pump for both washers	140	200	30	80	20
4.	Back shower pump for both washers	20	75	70	60	10
5.	Blow Tank dil. Pump	300	250	30	80	42
6.	Blow Tank transport pump	420	250	30	80	90
7.	Press Feed pump	430	300	40	60	107
8.	Feed Chest dil. Pump	300	250	35	80	50
9.	Wash and Dilution pump	150	200	30	80	21
10.	Medium Cy. Pump	125	300	117	40	136

CALCULATION OF TANK DIMENTIONS:

1. PRESS FILTRATE TANK :

Filtrate coming in tank from T.R. Press as shown in material balance = 300 m^3/hr.

Keeping 30 min. filtrate in a tank,

The volume of tank required = 150 m^3

Also, D = 0.8 H

Where D = diameter of tank, mt

H = height of tank, mt

Therefore,

Volume of tank = 150 m^3/hr

π . (D2/4) H = 150

π .[(0.8 H)^2/4] H = 150

H = 7.2 mt

Then, D = 0.8 x 7.2 = 5.8 mt.

2. WASHER'S SEAL TANK :

Filtrate coming in seal tank from washers as shown in material bal. = 994 m^3/hr.

Keeping 30 min. filtrate in a tank,

The volume of tank required = 994/2 = 497 m^3

Also, D = 0.8 H

Where D = diameter of tank, mt

H = height of tank, mt

Therefore,

Volume of tank = 497 m^3/hr

π. (D2/4) H = 497

π .[(0.8 H)^2/4] H = 497

H = 10.7 mt

Then, D = 0.8 x 10.7 = 8.6 mt.

3. HOT WATER TANK :

Total volume of water to be used on washer shower @ 2.5 m^3/t (D.F = 2.5) as shown in material balance = 120 m^3/hr.

Keeping 30 min. water volume in a tank,

The volume of tank required = 120/2 = 60 m^3

Also, D = 0.8 H

Where D = diameter of tank, mt

H = height of tank, mt

Therefore,

Volume of tank = 60 m^3/hr

$\pi . (D^2/4) H = 60$

$\pi .[(0.8 H)^2/4] H = 60$

H = 5.3 mt

Then, D = 0.8 x 5.3 = 4.2 mt

4. HIGH DENSITY (H.D) TOWER : (CIVIL)

Washed pulp coming in H.D.Tower from washer no.2 as shown in material balance = 102 m^3/hr.

Keeping 5 hours pulp in the tower,

The volume of tower required = 102 x 5 = 510 m^3/hr

Also, D = 0.8 H

Where D = diameter of tower, mt

H = height of tower, mt

Therefore,

Volume of tower = 510 m^3/hr

π. (D2/4) H = 510

π .[(0.8 H)^2/4] H = 510

H = 10.8 mt

Then, D = 0.8 x 10.8 = 8.6 mt.

CALCULATION OF OXYGEN REACTORS AND POST OXYGEN WASHERS DIMENTIONS:

1. OXYGEN REACTOR NO.1 :

The feed volume of pulp to oxygen reactor no.1 = 104 m^3/hr.

To provide, 30 min. pulp retention inside the reactor, the volume of reactor should be

= 104/2 m^3.

= 52 m^3

As per reactor design,

D = 0.1 H

Where,

D = diameter of reactor, mt

H = height of reactor, mt

Therefore,

Volume of reactor = 52 m^3/hr

$\pi. (D2/4) H = 52$

$\pi .[(0.1 H)^2/4] H = 52$

H = 18.8 mt

Then, D = 0.1 x 18.8 = 1.9 mt.

2. OXYGEN REACTOR NO.2 :

The feed volume of pulp to oxygen reactor no.2 = 108 m^3/hr.

To provide, 60 min. pulp retention inside the reactor, the volume of reactor remains the same = 108 m^3

As per reactor design,

$D = 0.1\ H$

Where,

D = diameter of reactor, mt

H = height of reactor, mt

Therefore,

Volume of reactor = 108 m^3

$\pi.\ (D2/4)\ H = 108$

$\pi\ .[(0.1\ H)^2/4]\ H = 108$

H = 24.0 mt

Then, D = 0.1 x 24.0 = 2.4 mt.

3. OXYGEN BLOW TANK :

The feed volume of pulp to oxygen blow tank = 108 m^3/hr.

To provide, 30 min. pulp residence inside the oxygen blow tank, the volume of blow tank should be

= 108/2 m^3

= 54 m^3

As per reactor design,

$D = 0.25\ H$

Where,

D = diameter of reactor, mt

H = height of reactor, mt

Therefore,

Volume of reactor = 54 m^3/h

$\pi\ .\ (D^2/4)\ H = 54$

$\pi\ .[(0.25\ H)^2/4]\ H = 54$

H = 10.3 mt

Then, D = 0.25 x 10.3 = 2.6 mt.

4. POST OXYGEN WASHERS :

We know from the literature that washer loading factor for Hard Wood

$= 0.6$ ODT/day/ft^2

In other words,

0.6 ODT/day pulp will wash on $=1$ ft^2 surface area

300 ODT/day pulp will wash on $= (1/0.6)300$ ft^2 surface area

$= 500$ ft^2

Since, 40 % of washer's surface area remains submersed

Therefore, adding 40 % area to the area calculated above

We get,

$\{(0.4 \times 500) + 500\} = 700$ ft^2

As per design of washer,

$$D = 0.6\,L$$

Surface area of the washer $= \pi\,.D.L$

Where,

D = diameter of washer, ft^2

L = face length of washer, ft^2

$\pi\,.D.L = 700$ ft^2

$\pi.\,(0.6\,L).L = 700$ ft^2

$L = 19.3$ ft

And $D = 0.6 \times 19.3 = 11.5$ ft.

TABLE 8: DIMENTIONS OF TANKS, REACTORS AND WASHERS

S.No.	Name of Tank	Quantity (Nos.)	Height of Tank (Mts)	Diameter of Tank (Mts)	Wall Thickness (MM)
1.	Press filtrate	1	7.2	5.8	10
2.	Seal tank	2	10.7	8.6	10
3.	Hot water	1	5.3	4.2	10
4.	Reactor no.1	1	18.8	1.9	-
5.	Reactor no.2	1	24.0	2.4	-
6.	Blow tank	1	10.3	2.6	10
7.	Washer	2	5.9	3.5	-
8.	H.D.Tower	1	10.8	8.6	-

CALCULATIONS OF PIPE LINE WEIGHT PER METRE FOR COST DETERMINATION:

The Surface area of a pipe is given by $S = \pi \times D \times L$ ....(1)

Where:

D = diameter of pipe line, mt

L = length of pipe line, mt

Knowing the surface area of the pipe, we can determine the volume of material by multiplying surface area with wall thickness of material i.e.

Volume of the material = surface area x wall thickness of material

or, $V = S \times t$ ....(2)

Mass of the material can further be calculated as given below:

Mass of material = volume of material x density of material

or, $M = V \times d$ .... (3)

Finally, we can determine the cost of material by multiplying weight of material with the cost per kg.

LINE SIZE: 450 mm

For 1 mt length of pipe –

$S = 3.14 \times 0.45 \times 1$ [from eq.(1)]

$S = 1.41\ m^2$

Since, pipe wall thickness $t = 6$ mm = 0.006 mt

Therefore, volume of material $V = 1.41 \times 0.006$ [from eq.(2)]

$V = 0.0085\ m^3$

Density of stainless steel (SS) $d = 7850\ kg/m^3$

Therefore, weight of material per unit length = 0.0085 x 7850

= 66.7 kg/ mt

Similarly we can calculate weight per unit for other pipe lines.

LINE SIZE: 350 mm

Weight per unit length = S x t x d

= π D L x t x d

= 3.14 x 0.35 x 1 x 0.006 x 7850

= 51.8 kg/mt

LINE SIZE: 300 mm

Weight per unit length = S x t x d

= π D L x t x d

= 3.14 x 0.30 x 1 x 0.006 x 7850

= 44.4 kg/mt

LINE SIZE: 250 mm

Weight per unit length = S x t x d

= π D L x t x d

= 3.14 x 0.25 x 1 x 0.006 x 7850

= 37.0 kg/mt

LINE SIZE: 200 mm

Weight per unit length = S x t x d

= π D L x t x d

= 3.14 x 0.20 x 1 x 0.006 x 7850

= 29.6 kg/mt

LINE SIZE: 75 mm

Weight per unit length = S x t x d

= π D L x t x d

= 3.14 x 0.075 x 1 x 0.006 x 7850

= 11.1 kg/mt

LINE SIZE: 50 mm

Weight per unit length = S x t x d

= π D L x t x d

= 3.14 x 0.050 x 1 x 0.006 x 7850

= 7.4 kg/mt

TABLE 9: PIPING COST

S.NO.	LINE DIAMETER (mm)	REQD. LINE LENGTH (mt) A	LINE WEIGHT PER METRE (kg/mt) B	COST OF MATERIAL (Rs per kg) C	COST OF PIPE (Rs) AxBxC
1.	450	35	66.7	260	606970
2.	350	35	51.8	260	471380
3.	300	220	44.4	260	2539680
4.	250	185	37.0	260	1779700
5.	200	340	29.6	260	2616640
6.	75	140	11.1	260	404040
7.	50	20	7.4	260	38480
TOTAL					84,56,890/-

TABLE 10: BEND COST

S.NO.	LINE DIAMETER (mm)	NO. OF BENDS A	LINE WEIGHT (kg/mt) B	COST OF MATERIAL (Rs per kg) C	COST OF BENDS (Rs) AxBxC
1.	450	5	66.7	520	173420
2.	350	6	51.8	520	161616
3.	300	12	44.4	520	277056
4.	250	24	37.0	520	461760
5.	200	35	26.6	520	538720
6.	75	12	11.1	520	69264
7.	50	8	7.4	520	30784
TOTAL					17,12,620/-

Hence, the total cost of piping with bends = Rs.84, 56,890 + Rs.17, 12,620
= Rs. 1, 01, 69,510 /-
≈ Rs. 102 lakhs.

CALCULATION OF TANK MATERIAL WEIGHT FOR COST DETERMINATION:

In the same fashion as we calculate the pipe line material weight/mt, we can determine the tank material weight with the help of same relations as given above.

1. Press Filtrate Tank:

Material of construction = Mild steel
Density of material (d) = 7860 kg/m^3
Rate of material = Rs. 40 per kg.
Wall thickness of tank (t) = 10 mm
= 0.01 mt

Weight of material = surface area of tank x wall thickness of tank x density of material.
= π D L x t x d
= 3.14 x 5.8 x 7.2 x 0.01 x 7860
= 10306.5 kg

Cost of material = weight of material x rate of material
= 10306.5 x 40
= Rs.412261.4
≈ Rs. 4, 12,262 /-

2. Washers Seal Tank No.1:

Material of construction = Stainless steel

Density of material (d) = 7850 kg/m^3

Rate of material = Rs. 260 per kg.

Wall thickness of tank (t) = 10 mm

= 0.01 mt

Weight of material = surface area of tank x wall thickness of tank x density of material.

= π D L x t x d

= 3.14 x 8.6 x 10.7 x 0.01 x 7850

= 22682.0 kg

Cost of material = weight of material x rate of material

= 22682.0 x 260

= Rs.5897322.5

≈ Rs. 58, 97,323 /-

3. **Washers Seal Tank No.2**:

Material of construction = Mild steel

Density of material (d) = 7860 kg/m^3

Rate of material = Rs. 40 per kg.

Wall thickness of tank (t) = 10 mm

= 0.01 mt

Weight of material = surface area of tank x wall thickness of tank x density of material.

= π D L x t x d

= 3.14 x 8.6 x 10.7 x 0.01 x 7860

= 22710.9 kg

Cost of material = weight of material x rate of material

= 22710.9 x 40

= Rs.908436.2

≈ Rs. 9, 08,437 /-

4. Hot Water Tank:

Material of construction = Mild steel

Density of material (d) = 7860 kg/m^3

Rate of material = Rs. 40 per kg.

Wall thickness of tank (t) = 10 mm

= 0.01 mt

Weight of material = surface area of tank x wall thickness of tank x density of material.

$= \pi D L \times t \times d$

$= 3.14 \times 4.2 \times 5.3 \times 0.01 \times 7860$

= 5493.8 kg

Cost of material = weight of material x rate of material

= 5493.8 x 40

= Rs.219754.3

≈ Rs. 2, 19,755 /-

5. Oxygen Blow Tank:

Material of construction = Stainless steel

Density of material (d) = 7850 kg/m^3
Rate of material = Rs. 260 per kg.
Wall thickness of tank (t) = 10 mm
= 0.01 mt

Weight of material = surface area of tank x wall thickness of tank x density of material.
= π D L x t x d
= 3.14 x 2.6 x 10.3 x 0.01 x 7850
= 6601.0 kg

Cost of material = weight of material x rate of material
= 6601.0 x 260
= Rs.1716260.6
≈ Rs. 17, 16,261 /-

TABLE 11: TANK MATERIAL COST

S.No.	NAME OF TANK	WEIGHT OF MATERIAL (kgs) A	RATE OF MATERIAL (Rs/kg) B	COST OF MATERIAL (Rs) AxB
1.	Press Filtrate Tank	10306.5	40.00	412262
2.	Seal Tank No.1	22682.0	260.00	5897323
3.	Seal Tank No.2	22710.9	40.00	908437
4.	Hot Water	5493.8	40.00	219755
5.	Blow Tank	6601.0	260.00	1716261
TOTAL				91,54,038/-

Total Rs.≈ 91.6 lakhs.

CALCULATION OF HEAT EXCHANGER SPECIFICATIONS:

The mill water temperature (T2) = 27°C

The wash water temperature required (T1) = 60°C

Specific heat of weak black liquor (Cp) = 0.95 kcal/kg°C

Therefore, the heat required (Q) to raise the temperature of mill water from 27°C to 60°C will be :

$$Q = m\,C_p\,(T_2 - T_1)$$

Where m = mass of water, kg/hr

Here m = 120 m^3/hr as per material balance calculations.

$$Q = 120 \times 1000 \times 0.95\,(60 - 27)$$

$$Q = 3762000 \text{ kcal.}$$

Now, the heat transfer rate is given by relation:

$$Q/A = (T_2 - T_1) / (1/U)$$

or, $$Q = A\,U\,(T_2 - T_1) \quad \ldots\ldots\ldots\ldots\ldots\ldots\ldots\ldots (1)$$

Where, U = heat transfer coeff. of S.S material.

= 4000 kcal/m^2.hr.°C

A = surface area of heat exchanger, m^2

From equation (1)-

$$3762000 = A \times 4000\,(60 - 27)$$

or, $$A = 28.5 \text{ m}^2$$

Taking number of tubes in heat exchanger = 100

and, diameter of each tube = 25.4 mm

Surface area of tube = $\pi D L$

Where, D = diameter of tube, m

L = length of tube, m

POWER CALCULATIONS FOR AGITATORS:

BLOW TANK AGITATOR:

The agitator recommended for the pulp agitation in the blow tank is of paddle type. The power required to drive this agitator can be calculated by formula given below :

$$Ne = i \cdot k \cdot \gamma \cdot F \cdot v^3 / g \cdot 102 \cdot d \qquad kw$$

Where:

i = no. of paddles in the agitator.

k = constant and value may be known from the chart given below.

l/s	1	2	4	10	8
k	1.1	1.15	1.19	1.20	1.40

l = length of paddle, mt

s = width of paddle, mt

F = surface area of paddle, mt^2

γ = density of stock, kg/mt^3

v = circumferential speed of agitation which is ≤ 0.75 mt/sec

g = acceleration due to gravity, mt/sec^2

d = diameter of shaft, mt

Now,

Length of paddle of agitator = 0.3 mt

Width of paddle of agitator = 0.0375 mt

Therefore,

l/s = 0.3/0.0375 = 8

the value of k from the chart is equal to 1.40

since, the paddle of the agitator is in rectangular in shape

therefore,

surface area of paddle $F = l \times s$

$= 0.3 \times 0.0375$

$= 0.01125 \text{ mt}^2$

Density of stock (γ) = 1050 kg/mt^3

Number of paddles in the agitator (i) = 8

Diameter of agitator shaft (d) = 0.032 mt

Hence,

Power required $Ne = 8 \,.\, 1.40 \,.\, 1050 \,.\, 8(0.01125) \,.\, (0.75)^3 / 9.81 \,.\, 102 \,.\, 0.032$

$= 13.9$ kw

Adding 30 % extra power for design purpose –

$Ne = 13.9 \times 1.3 = 18.07$ kw

≈ 18 kw

HIGH DENSITY (H.D) TOWER :

Type of agitator recommended for H.D Tower is of propeller type. The power required to drive propeller type agitator is given by –

$$Ne = k_2 \,.\, \gamma / 75 \,.\, g \{ (n/60)^5 \, d^5 \} \qquad \text{kw}$$

Where ;

k_2 = constant and selected from the table given below.

S.No.	Propeller Ratio	Value of k_2
1.	s : d = 1.0	0.30
2.	s : d = 1.5	0.62
3.	s : d = 2.0	1.00
4.	s : d = 2.5	1.35

s & d are surface area and diameter of propeller of agitator respectively.

Also, diameter of propeller (d) = 1.07 mt

Length of propeller (l) = 0.8 mt

Therefore,

Surface area of propeller (s) = π d l

= 3.14 x 1.07 x 0.8

= 2.69 mt^2

Hence,

s/d = 2.69/1.07 = 2.5

or, s : d = 2.5

the value of k_2 from the table given above is equal to 1.35

γ = density of stock (1050 kg/mt^3)

g = acceleration due to gravity (9.81 mt/sec^2)

n = no. of revolution of propeller (150 rev./min)

Using the formula given above,

Ne = 1.35. 1050 / 75. 9.81 { $(150/60)^3$ $(1.07)^5$ }

Ne = 42.1 kw

Adding 30 % extra power for design purpose –

Ne = 42.1 x 1.3 = 54.8 kw

≈ 55 kw

TABLE-12

ESTIMATION OF OXYGEN DELIGNIFICATION PLANT COST:

1. **Equipment cost –**

S.NO	COST HEAD	QUANTITY	COST,Rs lacs
1.	Twin roll press including hydraulic unit	1 no.	350
2.	Shredder / conveyor	1 no.	20
3.	Chemical mixers	2 no.	30
4.	O_2 reactor no.1	1 no.	200
5.	O_2 reactor no.2	1 no.	250
6.	Flow heater	1 no.	05
7.	Washer drum including conveyor repulper and other accessories	2 no.	300
8.	Heat exchanger	1 no.	20
TOTAL			1175

2. **Civil cost –**

S.NO	COST HEAD	QUANTITY	COST,Rs lacs
1.	H.D.Tower construction	1 no.	50
2.	Foundations, pathways, staircases, flooring, roofing etc.	-	130
TOTAL			180

3. **Mechanical cost –**

S.NO	COST HEAD	QUANTITY	COST,Rs lacs
1.	Pumps	20 no.	230
2.	Agitators	2 no.	20
3.	Piping including bends	-	101.7
4.	Tank material	-	91.6
5.	Hand operated valves	50 no.	10
6.	Labour cost	-	100
TOTAL			553.3

4. **Electrical cost –**

S.NO	COST HEAD	QUANTITY	COST,Rs lacs
1.	Motors	30 no.	30
2.	Electrical cables	-	50
3.	Transformer / MCC panels	-	60
4.	Labour cost	-	50
TOTAL			190

5. Instrument cost –

S.NO	COST HEAD	QUANTITY	COST,Rs lacs
1.	Control valves / Flow meters	-	100
2.	DCS panels / programming	-	200
3.	Cables and air piping	-	50
4.	Labour cost	-	50
TOTAL			400

GRAND TOTAL OF S.No. 1, 2, 3, 4 and 5 = 2498.3 lacs

= 24.98 crores

$\boxed{\approx 25 \text{ crores}}$

PART: III

EFFECTS OF OXYGEN DELIGNIFICATION PLANT ON MILL

1. IMPACT OF OXYGEN DELIGNIFICATION ON PULP QUALITY
2. IMPACT OF OXYGEN DELIGNIFICATION ON ENVIRONMENT
3. COMPERATIVE STUDY- BEFORE AND AFTER INSTALLATION OF OXYGEN DELIGNIFICATION PLANT
4. RESULTS AND DISCUSSION
5. CONCLUTIONS
6. GRAPHICAL REPRESENTATION OF RESULTS

IMPACT OF OXYGEN DELIGNIFICATION OF CHEMICAL PULP ON PULP QUALITY:

The oxygen delignification stage is used to continue the removal of lignin, which is initiated in the cooking process, in a more selective manner i.e. with less cellulose degradation, that what is possible with a prolonged cooking stage. The pulp becomes cleaner and brighter as more and more lignin is removed.

The operating conditions in an oxygen delignification stage are chosen such that as much as possible of the lignin is removed while maintaining pulp yield, minimizing pulp viscosity loss, and without damaging pulp strength. If the oxygen delignification stage inlet kappa number is properly controlled and the deviation of the outlet kappa number is low, then the target kappa number can be lowered without having a negative effect on pulp viscosity or strength.

The strength properties of conventionally bleached pulps and pulps that have been oxygen delignified to an extent of about 50% are quite comparable. The effect of oxygen bleaching on pulp cleanliness, pitch, brightness stability is slightly beneficial. The refining energy requirements are found to be slightly less than normal unbleached pulp.

Oxygen Delignification details -

Mill unbleached pulp comprising of 60 % eucalyptus and 40 % poplar was oxygen delignified using the operating conditions described in Table – 1.

TABLE – 1: OXYGEN DELIGNIFICATION PROCESS CONDITIONS

S.No	CONDITIONS	UNIT	VALUES
1.	Inlet Pulp consistency	%	11
2.	Chemical applied :		
	Oxygen	kg/T	18
	Caustic	kg/T	20
3.	Temperature		
	Ist stage	°C	65-70
	IInd stage	°C	90-100
4.	Residence Time		
	Ist stage	Min.	30
	IInd stage	Min.	60
5.	Reactor Top Pressure		
	Ist stage	kg/cm^2	7.5
	IInd stage	kg/cm^2	3.5
6.	End pH	-	10.4-10.8

The mill unbleached pulp was evaluated as per TAPPI standard test methods and the results are arranged in Table – 2.

TABLE – 2: EVALUATION OF MILL UNBLEACHED PULP

S.No.	PARTICULARS	UNIT	RESULTS
1.	Permanganate No.	No.	13.4
2.	Brightness	°PV	30.1
3.	Viscosity	cps	28

Further, the mill unbleached pulp was oxygen delignified under process conditions shown in Table – 1 and results are tabulated in the table given below:

TABLE – 3: EVALUATION OF OXYGEN DELIGNIFIED PULP

S.No.	PARTICULARS	UNIT	RESULTS
1.	Permanganate No.	No.	8.0
2.	Brightness	°PV	45.2
3.	Viscosity	cps	22

Now, the mill unbleached pulp and oxygen delignified pulp were compared for pulp evaluation results and difference in various pulp properties are shown in Table – 4.

TABLE – 4: COMPARISION OF MILL UNBLEACHED PULP AND OXYGEN DELIGNIFIED PULP

S.No.	PARTICULARS	UNIT	% REDUCTION/ GAIN
1.	Permanganate no. reduction	%	40.3
2.	Brightness gain	%	33.4
3.	Viscosity loss	%	21.4
4.	Pulp shrinkage	%	2.4

After pulp comparison, mill unbleached pulp and oxygen delignified pulp were beaten to 30 °SR freeness and evaluated for physical strength properties. The results are shown in Table – 5.

TABLE – 5: COMPARISON OF PHYSICAL STRENGTH PROPERTIES OF MILL UNBLEACHED PULP AND OXYGEN DELIGNIFIED PULP

S.No.	PARTICULARS	UNIT	MILL UB PULP	MILL OD PULP	% DIFFERENCE
1.	Initial Freeness	°SR	16	17	-
2.	Final Freeness	°SR	30	30	-
3.	Beating time	Min.	25	22	-
4.	Burst Factor	-	36	37	+ 2.7
5.	Tear Factor	-	81	78	- 3.7
6.	Double Fold	Nos.	56	42	-25.0
7.	Breaking Length	Mts	5342	5168	-3.2
8.	Bulk density	c.c/gm	1.44	1.42	-1.4

IMPACT OF OXYGEN DELIGNIFICATION OF CHEMICAL PULP ON ENVIRONMENT:

Strict legislation of central pollution control board in the country and increasing cost of energy, chemicals, utilities and increasing demand of high brightness paper has forced the paper industry for modification of present bleaching practices. Therefore, incorporation of oxygen delignification at pre-bleaching stage will reduce the use of chlorine and its compounds.

The oxygen delignified pulp filtrate was analyzed and data's are reported in Table – 6.

TABLE -6: EVALUATION OF OXYGEN DELIGNIFICATION FILTRATE

S.No.	PARTICULARS	UNIT	VALUES
1.	pH	No.	10.5
2.	COD	mg/lt	3176
3.	Suspended Solids	mg/lt	52
4.	Dissolved Solids	mg/lt	5782
5.	Total Solids	mg/lt	5834
6.	Chlorides	mg/lt	100

Chemical oxygen demands at different intermediate stages of oxygen delignification were calculated in the form of kgs per oven dry pulp and are given as under:

- **COD in pulp after brown stock washing = 100 kg/ODT pulp**
- **COD in pulp after Twin Roll Press = 40 kg/ODT pulp**
- **COD in pulp after Post Oxygen Washer = 10 kg/ODT pulp**

The above figures show that almost 90 % of COD in pulp was reduced before pulp enters at chlorination stage of bleaching.

The combined effluent generated by CEpHH and OCEopHH bleaching sequences were analyzed for effluent characteristics and results are tabulated in Table-7

TABLE- 7: ANALYSIS OF COMBINED EFFLUENT GENERATED BY TWO DIFFERENT BLEACHING SEQUENCES

S.No.	PARTICULARS	UNIT	BLEACHING SEQUENCE		% REDUCTION
			CEpHH	OCEopHH	
1.	COD	mg/lt	4539	2697	40.6
2.	SS	mg/lt	1265	952	24.7
3.	DS	mg/lt	24002	15235	36.5
4.	TS	mg/lt	25267	16187	35.9
5.	Chlorides	mg/lt	9800	6400	34.7

COMPERATIVE STUDY BEFORE AND AFTER INSTALLATION OF OXYGEN DELIGNIFICATION PLANT:

1. PHYSICAL STRENGTH PROPERTIES OF BLEACHED PULP-

The bleached pulp made with bleaching sequences CEpHH and OCEopHH were compared for physical strength properties. The comparison is given in Table – 8.

TABLE-8: COMPARISON OF PHYSICAL STRENGTH PROPERTIES USING CEpHH and OCEopHH BLEACHING SEQUENCES

S.NO.	PARTICULARS	UNIT	BLEACHING SEQUENCES		% DIFFERENCE
			CEpHH	OCEopHH	
1.	Initial Freeness	°SR	18	18	_
2.	Final Freeness	°SR	40	40	_
3.	Burst Factor	_	32	32	_
4.	Tear Factor	_	50	52	+3.8
5.	Breaking Length	Mts	4945	5092	+2.9
6.	Double Fold	Nos.	35	31	-11.4
7.	Brightness	°PV	86	89	+3.4
8.	Viscosity	cps	7	9	+22.2

2. TOTAL SOLIDS CONTENT IN WEAK BLACK LIQUOR-

Wash water flow rate on post oxygen washer no.2 = 9.8 m^3/t pulp
@ DF=2.5

Total solids in washer filtrate = 5.834 kg/m^3

Therefore, total solids = 9.8 m^3/t pulp x 5.834 kg/m^3
= 57.17 kg/t pulp
= 0.0572 t/t pulp

Total solids in black liquor from wash plant before ODL plant = 1.60 t/t pulp

Total solids in black liquor from wash plant after ODL plant = (1.60 + 0.0572) t/t pulp
= 1.66 t/t pulp

% gain in total solids with ODL plant = {(1.66 – 1.60) / 1.66}100 = 3.6 %

Hence, increase in black liquor solids with oxygen delignification plant = 3.6 %

3. COST COMPARISON OF BLEACHED PULP-

The cost comparison of two bleaching sequences i.e. CEpHH and OCEopHH was worked out and found that the cost of bleached was increased by Rs.747/- per MT. The cost comparison is shown in Table-9.

TABLE-9: COMPARISON OF BLEACHED PULP COST USING CEpHH AND OCEopHH BLEACHING SEQUENCES

S.No	PARTICULARS	BLEACHED PULP COST					
		CEpHH Sequence			OCEopHH Sequence		
		Consumption kg/t	Rate Rs/kg	Bleaching cost Rs/t	Consumption kg/t	Rate Rs/t	Bleaching cost Rs/t
1.	Oxygen	-	-	-	20^{*}	8	160
2.	Chlorine	56	10	560	42	10	420
3.	Alkali	35	20	700	$55^{\#}$	20	1100
4.	Hydrogen Peroxide	12	45	540	12	45	540
5.	Chlorine as hypo	56	12	672	42	12	504
6.	Power	$163^{\blacksquare}$	4	652	$223^{\blacksquare}$	4	892
7.	L.P.Steam	500	0.4	200	500	0.4	200
8.	M.P.Steam	-	-	-	400	0.6	240
9.	Water	$35^{\Diamond}$	1.5	52.5	$45^{\Diamond}$	1.5	67.5
TOTAL				Rs. 3376.5			Rs. 4123.5

* 16 kg/t in ODL and 4 kg/t in Eop stage making total 20 kg/t
# 20 kg/t in ODL and 35 kg/t in Eop stage making total 55 kg/t
■ figures in kwh/t
◊ figures in m^3/t

Net increase in bleach pulp cost by OCEopHH sequence = Rs 747/- per MT.

4. MILL EFFLUENT GENERATION-

The combined mill effluent at the inlet of the effluent treatment plant after the installation of oxygen delignification plant was compared with the data's available before this plant. The comparison is shown in Table – 10.

TABLE-10: MILL EFFLUENT AT INLET OF ETP BEFORE AND AFTER OXYGEN DELIGNIFICATION PLANT (ODL)

S.NO	PARTICULARS	UNIT	BEFORE ODL	AFTER ODL	% REDUCTION
1.	COD	mg/lt	1200	800	33.3
2	Suspended Solids	mg/lt	800	600	25.0
3.	AOX	mg/lt	1.5	1.0	33.3
4.	Water consumption	m^3/t paper	120	120	-

RESULTS AND DISCUSSION:

1. The permanganate number reduction of mill unbleached pulp by oxygen delignification plant was observed 40% (approx.). The loss in viscosity at this reduction was found to be 21% (approx.).

2. The pulp shrinkage after oxygen delignification was 2.4% without using any cellulose protector.

3. The comparison of mill unbleached pulp with oxygen delignified pulp for physical strength properties showed that there was marginal increase in burst factor with decrease in double fold by 25%. Other strength properties were within range.

4. Higher amount of COD, suspended solids, dissolved solids and color in the effluent was observed. However this filtrate can be mixed with weak black liquor and can be recycled to soda recovery for reducing pollution load.

5. It was seen that Twin roll press plays an important role in oxygen delignification plant. It reduces 60% COD of pulp by squeezing it. Further COD reduction up to 90% was achieved by oxygen delignification.

6. Significant reduction in values of COD, SS, DS, TS and Chlorides of combined effluent generated by OCEopHH and CEpHH bleaching sequences was there.

7. In continuation in reduction of mill effluent load, the analysis of combined effluent at inlet of ETP showed that there was 25 to 33 % reduction in COD,TS etc. by installation of oxygen delignification plant. AOX generation was reduced by 33 %.

8. There was no change in water consumption per MT of paper by the introduction of oxygen delignification plant.

9. It has been observed that there was 25% reduction in chlorine consumption per MT of pulp.

10. There was increase in oxygen consumption, alkali consumption, power consumption and steam consumption per MT of pulp resulting in increase in bleached pulp cost by Rs.747/- per MT.

11. The final bleached pulp brightness was increased by 3°PV and final bleached pulp viscosity showed an increased by 2 cps.

12. Other strength properties of oxygen delignified bleached pulp such as BF, TF and Breaking length were marginally high as compared to without oxygen delignified bleached pulp. However DF was found to be slightly less.

CONCLUSIONS:

- **Oxygen delignification results in reduction of unbleached pulp permanganate number by 40.3%, unbleached pulp brightness gain by 33.4% and unbleached pulp viscosity loss by 21.4%.**

- **With the introduction of oxygen delignification at pre bleaching stage, the final bleached pulp brightness and viscosity was increased by 3.4 % and 22.2 % respectively.**

- **The mill effluent analysis showed a remarkable reduction in COD, SS and AOX values by 33.3%, 25.0% and 33.3% respectively.**

- **The black liquor solid content going to soda recovery was increased by 3.4%.**

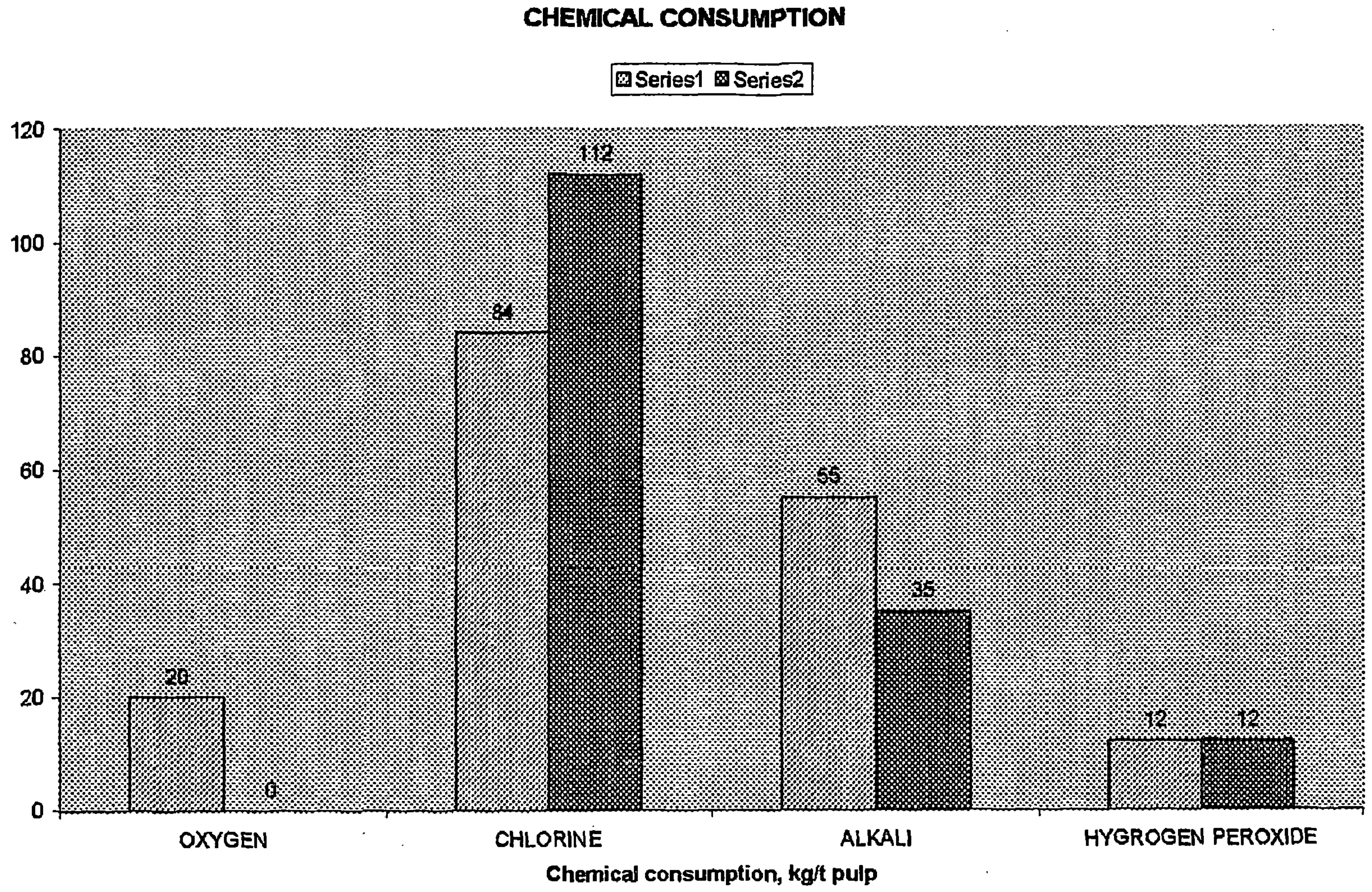
- **The increase in bleached pulp cost by Rs.747/- per MT was there. However, reduction in chlorine consumption by 25% was achieved.**

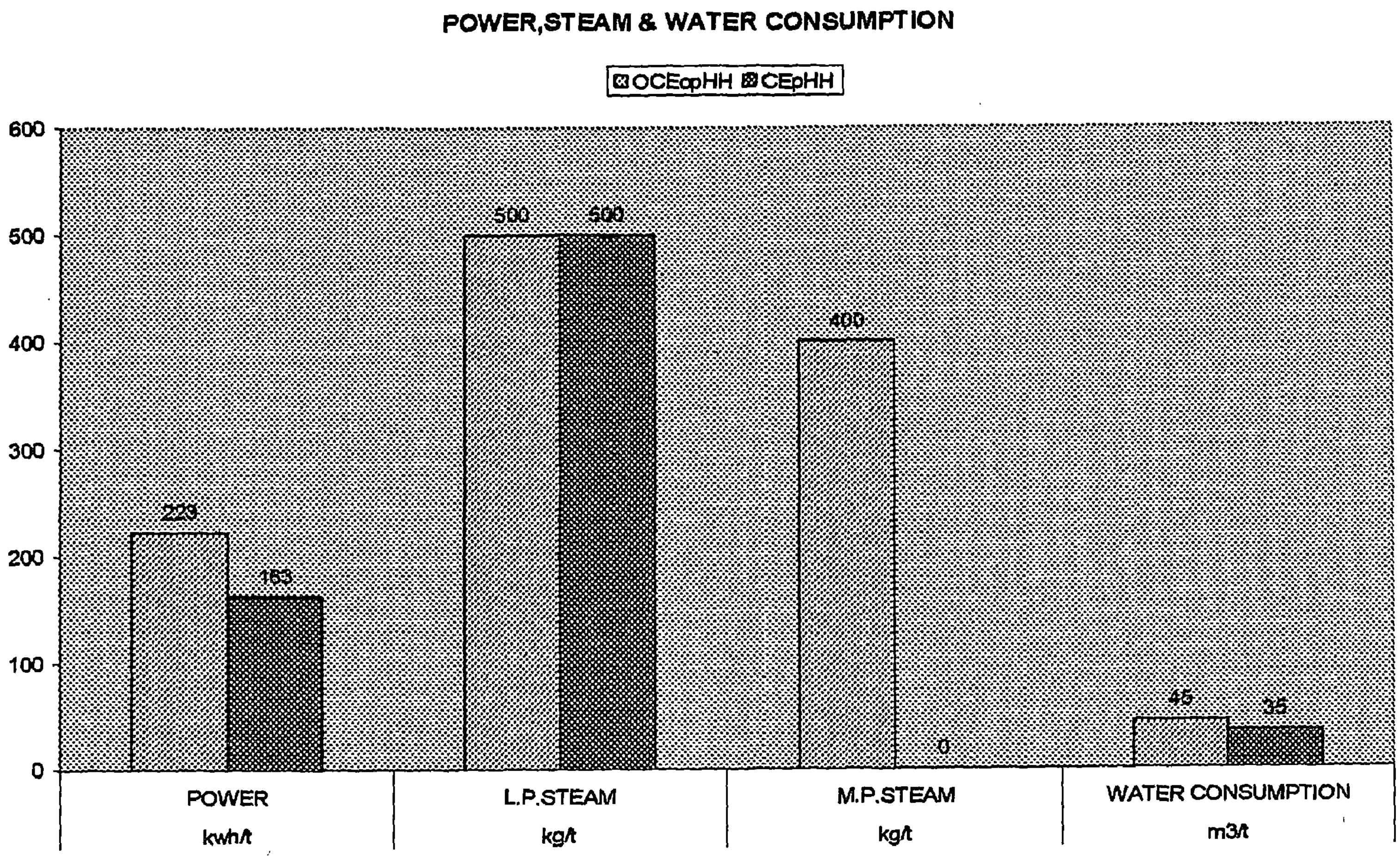
- **Other benefits include increase in pulp cleanliness, and increase in filler loading in paper due to better strength properties was seen.**

GRAPHICAL REPRESENTATION

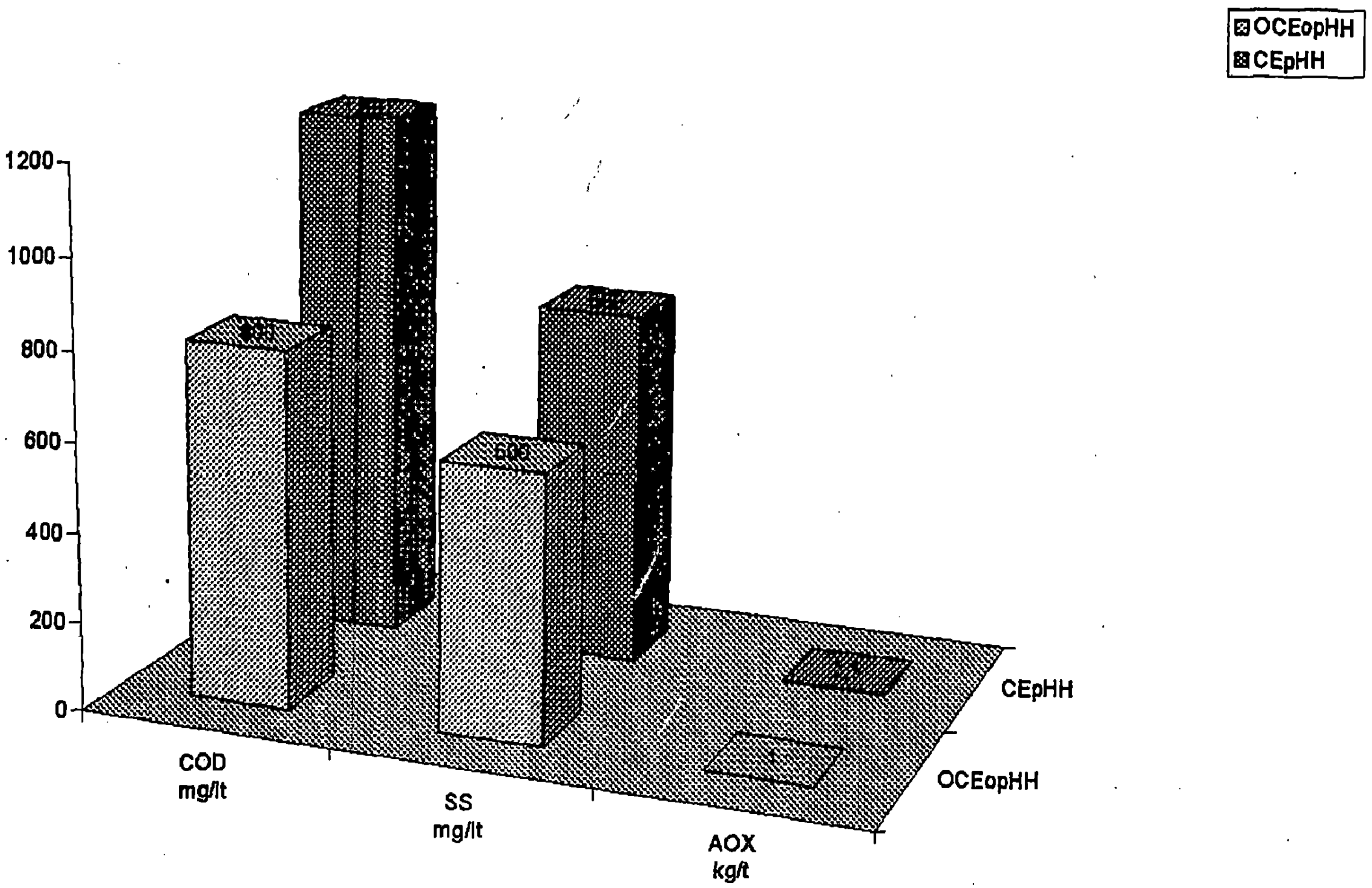
OF

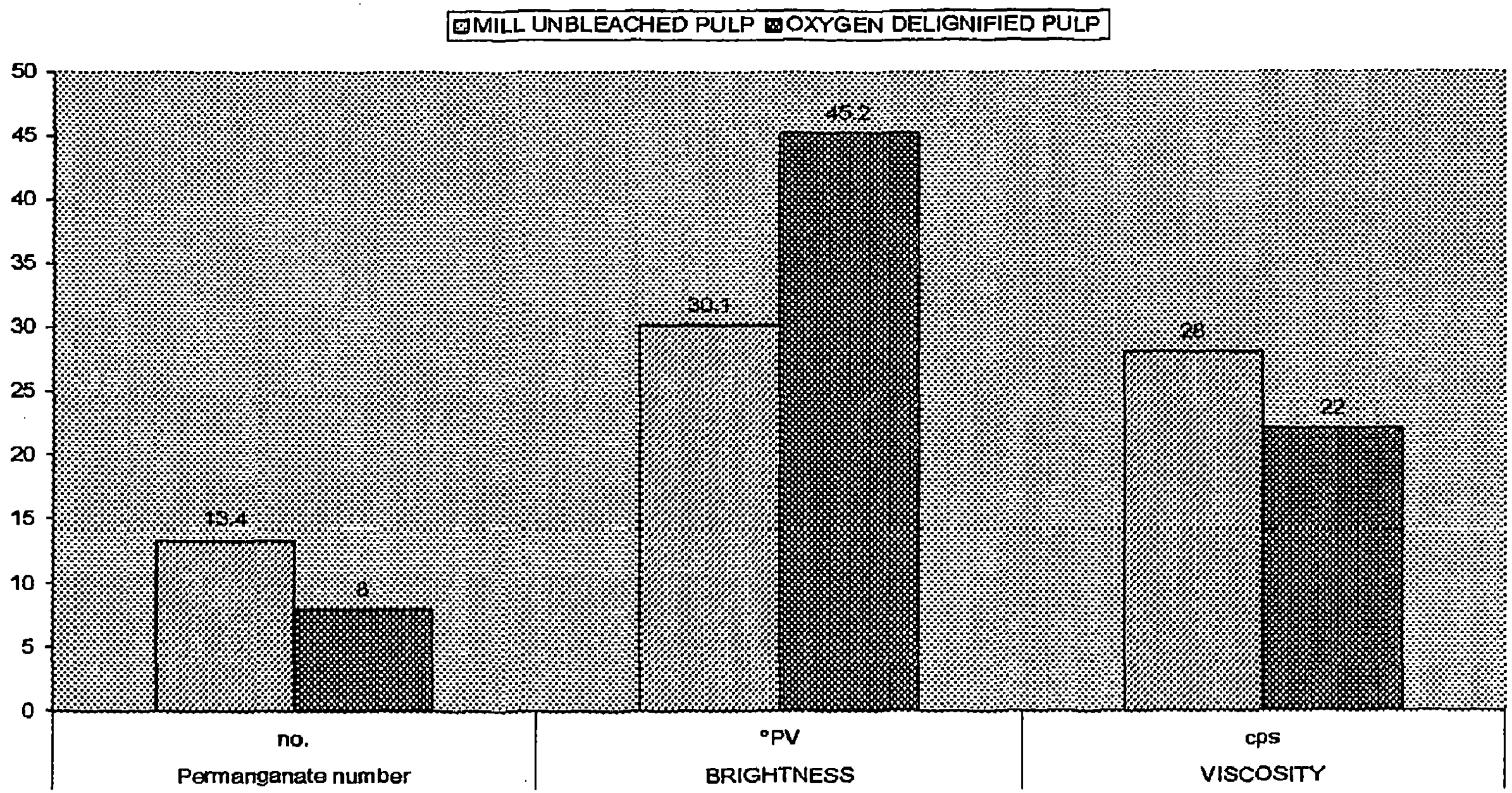
RESULTS

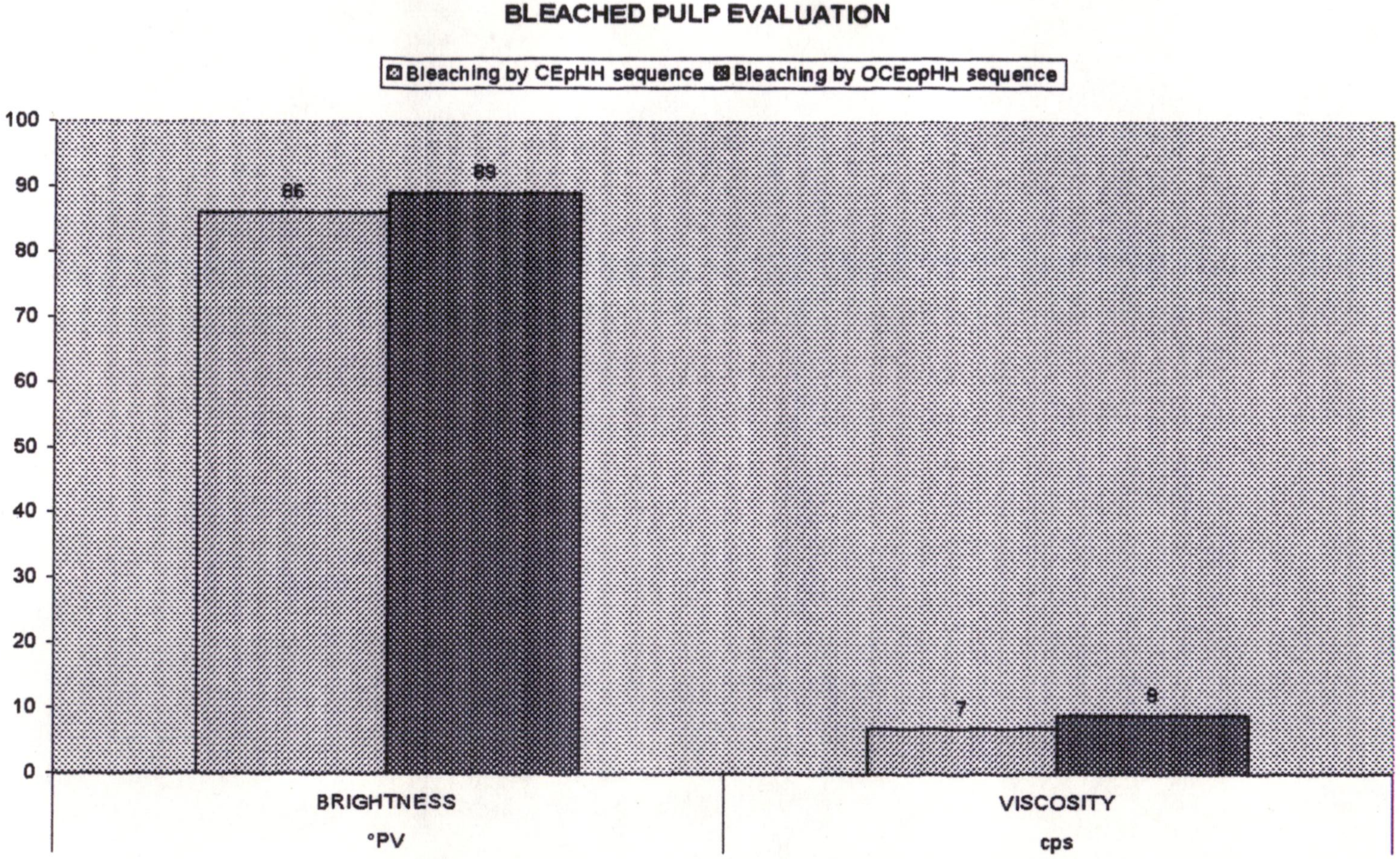

CHEMICAL CONSUMPTION
Series1
Series2
120
100
80
60
40
20
0
20
0
84
112
55
35
12
12
OXYGEN
CHLORINE
ALKALI
HYGROGEN PEROXIDE
Chemical consumption, kg/t pulp

POWER,STEAM & WATER CONSUMPTION
▨ OCEopHH ▨ CEpHH
600
500
400
300
200
100
0
223
163
500
500
400
0
45
35
POWER
kwh/t
L.P.STEAM
kg/t
M.P.STEAM
kg/t
WATER CONSUMPTION
m3/t

EFFLUENT DISCHARGE




UNBLEACHED PULP EVALUATION


BLEACHED PULP EVALUATION
▨ Bleaching by CEpHH sequence
▩ Bleaching by OCEopHH sequence
100
90
80
70
60
50
40
30
20
10
0
86
89
7
9
BRIGHTNESS
°PV
VISCOSITY
cps

PART: IV

MISCELLANEOUS

A. GREEK ALPHAPETS USED
B. STANDARD CODE USED
C. BIBLIOGRAPHY
D. REFERENCES

GREEK ALPHABETS USED

S.NO	NAME	SYMBOL
1.	Gamma	γ
2.	Eta	ή
3.	Mu	μ
4.	Pi	π
5.	Rho	ρ

STANDARD CODE USED

IS : 2825 – 1969	Code for unfired pressure vessel.
IS : 803 – 1976	Code of practice for design, fabrication and erection of vertical mild steel cylindrical welded oil storage tank.

BIBLIOGRAPHY

The various standard test methods used in the experimental investigations are as under:

S.NO	TESTING	TEST METHOD
1.	Consistency of pulp suspension	T 240 om-02
2.	Laboratory beating of pulp (PFI mill method)	T 248 sp-00
3.	Gram mage of paper and paperboard	T 410 om-98
4.	Physical testing of pulp hand sheets	T 220 sp-01
5.	Permanganate number of pulp	T 214 su-71
6.	Cuprammoniam disperse viscosity of pulp	T 206 os-63
7.	Pulp brightness	T 452 om-98
8.	Freeness of pulp	T 227 om-99
9.	Burst factor	T 403 om-97
10.	Tear Factor	T 414 om-98
11.	Double Fold	T 511 om-02
12.	Breaking Length	T 404 cm-92
13.	Bulk Density	T 426 wd-70

REFERENCES

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