CHAPTER 13

Beneficiation – Materials Handling

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- Alpha Coal Project, Galilee Basin, Queensland
- WICET Project, Gladstone, Queensland
- MMX Serra Azul Iron Ore EPCM Project, EBX Group, Brazil
- Tsagaan Suvarga Copper-Molybdenum concentrator, MAK, Mongolia
- Chagres smelter development, Anglo American, Chile.

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CHAPTER CONTENTS

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Beneficiation – Materials Handling

This chapter considers costing of materials handling associated with beneficiation, including conveying, slurry transfer systems, thickeners, clarifiers and filters.

CONVEYING

Cost considerations for conveying systems are considered in this section.

Conveyor capacity

The width of a troughed belt conveyor is determined by the required capacity, or the maximum size of the lumps to be conveyed.

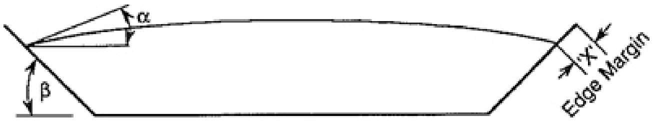
The capacity is directly proportional to the belt speed, the cross-sectional area of the material on the belt and the material density. For maximum efficiency, conveyors should operate fully loaded at maximum recommended speed.

Conveyor capacity tables

Table 13.1 shows conveyor capacity in tonnes/hour (t/h) for a material of density 1000 kg/m^3 at a belt speed of 1.0 m/s for various belt widths, trough angles and surcharge angles. Table 13.1 is derived from in-house data.

Angle of surcharge

The natural repose angle of the material being conveyed is decreased by the undulations of the belt passing over each idler. This decreased angle is known as the surcharge angle, and is one of the most important characteristics in determining carrying capacity, as it directly governs the cross-sectional area of the material on the belt, and hence the volume carried (Table 13.2). The surcharge angle is shown by the following schematic:



where:

β troughing angle 20°, 25°, 30°, 35° 40° or 45°

α surcharge angle 0°, 5°, 10°, 15°, 20° or 25°

X edge margin given by Table 13.2

As a general guide, values of surcharge angle α for various materials are:

TABLE 13.1A

Conveyor capacity.

Angle of surcharge (degrees)	Belt width (mm)	Idler angle (degrees)					
		20	25	30	35	40	45
		(t/h)					
0	400	22	27	31	35	39	42
	500	37	45	53	59	65	70
	650	67	82	95	107	118	126
	800	106	129	150	169	186	199
	1000	172	209	244	274	300	322
	1200	253	308	359	404	443	475
	1400	350	427	496	559	612	656
	1600	463	564	656	738	809	867
	1800	591	721	838	943	1033	1107
	2000	735	896	1043	1173	1284	1377
5	400	27	31	35	39	42	45
	500	45	53	60	66	72	76
	650	81	95	108	119	129	137
	800	128	150	170	188	204	216
	1000	206	243	276	304	329	349
	1200	304	358	406	448	484	513
	1400	420	494	561	619	669	709
	1600	556	653	741	819	884	936
	1800	710	834	947	1045	1128	1195
	2000	882	1038	1177	1300	1403	1486
10	400	31	36	41	43	46	49
	500	53	60	69	73	78	82
	650	95	109	125	132	141	148
	800	150	171	197	207	221	232
	1000	242	277	318	335	357	375
	1200	355	407	469	493	526	551
	1400	491	563	648	681	726	762
	1600	649	743	856	899	959	1006
	1800	829	949	1093	1148	1225	1284
	2000	1030	1180	1359	1427	1522	1596

TABLE 13.1B
Conveyor capacity.

Angle of surcharge (degrees)	Belt width (mm)	Idler angle (degrees)					
		20	25	30	35	40	45
		(t/h)					
15	400	36	42	44	48	50	53
	500	61	70	74	80	85	88
	650	109	127	134	144	152	158
	800	172	200	211	227	239	249
	1000	277	322	340	366	386	402
	1200	408	474	501	538	568	591
	1400	563	655	692	743	784	816
	1600	744	865	914	981	1036	1077
	1800	949	1105	1166	1252	1322	1374
	2000	1180	1373	1450	1557	1643	1708
20	400	41	45	49	52	55	56
	500	69	76	82	87	91	94
	650	124	136	147	157	164	169
	800	194	214	232	246	258	266
	1000	313	346	374	397	416	429
	1200	461	508	549	584	611	630
	1400	636	702	759	806	843	870
	1600	840	927	1002	1064	1113	1149
	1800	1072	1183	1279	1358	1421	1466
	2000	1333	1470	1589	1688	1766	1821
25	400	46	50	54	56	59	60
	500	77	84	90	94	98	101
	650	138	151	161	170	176	181
	800	217	237	253	266	277	284
	1000	350	381	408	429	446	457
	1200	515	560	599	631	655	671
	1400	711	774	827	871	904	926
	1600	939	1022	1092	1150	1193	1222
	1800	1198	1304	1394	1467	1522	1560
	2000	1489	1621	1732	1823	1892	1938

- fluid materials and cereal grains, 0° - 5°
- fine dry free-flowing materials, 5° - 10°
- average flowing materials, 15°
- non-flowing materials, 20°
- mixed lump and fines, 25°.

Typical conveyor surcharge angles for various materials are average flowing materials 15°, as shown in Table 13.3 (Rivis *et al*, 1993).

TABLE 13.2
Surcharge classification – values of edge margin 'X'.

Belt width (mm)	Dimension 'X' (mm)
400	45
500	50
650	58
800	65
1000	75
1200	85
1400	95
1600	105
1800	115
2000	125

TABLE 13.3
Characteristics of materials.

Material	Density (kg/m^3)	Angle of repose (degrees)	Angle of surcharge (degrees)
Alumina	800 - 960	22	5
Ashes, boiler	560 - 690	38 - 45	25
Bauxite, crushed	1200 - 1360	30 - 35	5 - 15
Clay, dry, loose	1010 - 1440	24 - 45	15 - 25
Coal, run-of-mine	720 - 880	35	25
Concrete, wet	1760 - 2400	Various	5
Copper ores, crushed	2080 - 2400	Various	25
Granite (40 - 50 mm)	1360 - 1440	25	Various
Gravel, dry sharp	1440 - 1600	30 - 40	25
Gypsum (50 - 75 mm)	1200 - 1280	30	20
Iron ores	1600 - 3200	35	5 - 25
Iron pyrite (50 - 75 mm)	2160 - 2320	Various	20
Lead ores	3200 - 4320	30	15
Limestone (50 - 75 mm)	1440 - 1520	35 - 40	25
Manganese ore	2000 - 2240	39	Various
Phosphate rock	1360	Various	Various
Quartz (50 - 75 mm)	1440 - 1520	35	Various
Quartz dust	1120 - 1280	40	Various
Sand, wet	1600 - 2080	15 - 30	5 - 15
Sand, dry	1440 - 1600	34 - 45	15

Maximum lump size

Lump size is a factor that determines the minimum belt width, particularly with low-capacity conveyors.

Table 13.4 (Rivis *et al*, 1993) gives maximum lump sizes recommended for various belt widths.

TABLE 13.4
Recommended maximum lump sizes.

Belt width (mm)	Uniform lumps (mm)	If mixed with approx 80% fines (mm)
400	75	125
500	100	175
650	125	250
800	150	300
1000	200	375
1200	300	450
1400	300	600
1600	375	600
1800	450	600
2000	450	600

Troughing angle

In Australia, the most commonly used troughing angles are 30° and 35°. The use of 20° troughing angles is generally limited to narrow low-capacity conveyors. The troughability (flexibility of belts to form a trough) of narrow belts influences trough idler angle selection.

Belt widths

The belt widths in Tables 13.4 and 13.5 are the preferred International Organization for Standardisation (ISO) standard belt widths, which are adopted as Australian Standards. Other belt widths can be manufactured, but the Standard recommends that for all new installations, the preferred widths are used.

TABLE 13.5
Typical belt speeds (m/s).

Belt width (mm)	Coal and earth	Hard ores and stone
400	1.5	-
500	2.2	1.7
650	2.7	2.5
800	3.0 - 3.5	2.7
1000	3.0 - 3.5	3.0
1200	3.5 - 4.0	3.0
1400	3.5 - 4.0	3.0
1600	4.0	3.0
1800	4.0	3.0
2000	4.0	3.0

Belt speeds

Typical belt speeds for general plant use are shown in Table 13.5 (adapted from BTR-Apex Handbook). Higher belt speeds are used on long overland and drift conveyors, where speeds of 5.0 m/s are not uncommon. In this case, detailed attention has to be given to the design and maintenance of loading, transfer and discharge points.

Angles of inclination

The maximum conveyor slope angles for various materials are shown in Table 13.6 (Rivis *et al*, 1993). Unless otherwise stated, the table refers to dry conditions, usually based on broken material in sizes most commonly found in conveyor systems.

TABLE 13.6
Conveyor maximum slope angles.

Material	Maximum slope angle (degrees)
Bituminous coal – ROM	18
Bituminous coal – sized	15 - 16
Bituminous coal – slack	20
Brown coal – ROM	18
Cement, Portland – loose	20
Clay – fine and dry	22
Clay – wet lump	18
Concrete – normal	15
Concrete – wet	10 - 12
Earth, loose and dry	18 - 20
Gravel, washed	12 -15
Gravel and sand – dry	18 - 20
Gravel and sand – wet	20 - 22
Lime, powdered	22
Ores – fines only	20
Ores – mixed lumps and fines	18
Ores – sized	16
Rock – fines only	20
Rock – mixed lumps and fines	18
Rock – sized	16
Sand – damp	18 - 20
Sand – dry	15
Sulfur – powdered	22

Where the material to be conveyed has unusual slumping characteristics, or where sufficient water is present to provide lubrication between the material and the belt cover, the slope angle is determined by test and/or experience.

Note that for drift conveyors out of coalmines handling run-of-mine (ROM) bituminous coal, slope angles of 15° to 16° are more commonly chosen than the permissible 18°.

Belt power calculation

The power to drive a simple conveyor in kW is calculated using Tables 13.7, 13.8 and 13.9 (Rivis *et al*, 1993), which are based on the arbitrary division of the total power requirements into three components:

1. P_e = power to drive the belt empty
2. P_h = power to carry the burden horizontally
3. P_l = power to raise or lower the burden.

The total power at the belt line $P_T = P_e + P_h + P_l$. The minimum motor power $Mp = P_T/f$, where f is the drive efficiency (say 95 per cent). The actual motor is then selected as the next nearest size, shown in Table 13.10.

TABLE 13.7

P_e – power (kW) to operate and unloaded conveyor per 1 m/s.

Width (mm)	Conveyor length (m)														
	10	16	25	40	63	80	100	125	160	200	250	315	400	500	630
400	0.4	0.4	0.4	0.5	0.7	0.8	0.9	1.1	1.2	1.5	1.8	2.1	2.6	3.2	4.0
500	0.5	0.5	0.6	0.7	0.9	1.0	1.1	1.4	1.7	20.	2.4	3.0	3.6	4.3	5.3
650	0.7	0.7	0.8	1.0	1.2	1.4	1.6	1.9	2.3	2.7	3.2	3.9	4.9	5.9	7.3
800	0.8	0.9	1.0	1.3	1.6	1.8	2.1	2.5	2.9	3.5	4.2	5.1	6.3	7.7	9.5
1000	1.1	1.2	1.4	1.6	2.0	2.4	2.7	3.2	3.8	4.5	5.4	6.6	8.1	10.0	12.3
1200	1.3	1.5	1.7	2.0	2.5	2.9	3.3	3.9	4.6	5.5	6.6	8.1	9.9	12.2	15.6
1400	1.6	1.7	2.0	2.3	2.9	3.4	3.9	4.6	5.5	6.5	6.8	9.5	11.7	14.3	17.7
1600	1.9	2.0	2.3	2.8	3.5	4.0	4.7	5.4	6.5	7.8	9.3	11.3	14.0	17.1	21.1
1800	2.1	2.3	2.6	3.1	3.9	4.5	5.1	6.0	7.2	8.6	10.3	12.6	15.4	18.9	23.3
2000	2.4	2.5	2.9	3.4	4.3	5.0	5.7	6.7	8.0	9.6	11.5	14.0	17.2	21.0	26.0

TABLE 13.8

P_h – power (kW) to convey the burden horizontally.

Capacity (t/h)	Conveyor length (m)														
	10	16	25	40	63	80	100	125	160	200	250	315	400	500	630
40	0.2	0.2	0.2	0.2	0.3	0.4	0.4	0.5	0.6	0.7	0.8	1.0	1.2	1.5	1.9
60	0.2	0.3	0.3	0.4	0.5	0.6	0.6	0.8	0.9	1.1	1.3	1.6	1.9	2.4	2.9
100	0.4	0.5	0.5	0.6	0.8	0.9	1.0	1.2	1.4	1.7	2.0	2.5	3.1	3.7	4.6
160	0.7	0.7	0.8	1.0	1.2	1.4	1.6	1.9	2.2	2.7	3.2	3.9	4.8	5.9	7.3
200	0.8	0.9	1.0	1.2	1.5	1.8	2.0	2.4	2.9	3.3	4.1	5.0	6.1	7.5	9.3
250	1.0	1.1	1.3	1.5	1.9	2.2	2.6	3.0	3.6	4.1	5.1	6.2	7.7	9.7	11.6
320	1.3	1.4	1.6	1.9	2.4	2.7	3.2	3.7	4.5	5.4	6.4	7.8	9.6	11.8	14.6
400	1.6	1.8	2.0	2.4	3.1	3.5	4.1	4.8	5.7	6.8	8.2	9.9	12.2	15.0	18.5
500	2.0	2.2	2.6	3.1	3.8	4.4	5.1	6.0	7.1	8.3	10.2	12.4	15.3	18.7	23.1
630	2.6	2.8	3.2	3.9	4.8	5.6	6.4	7.5	9.0	10.7	12.8	15.6	19.3	23.5	29.1
800	3.3	3.6	4.1	4.9	6.1	7.1	8.2	9.5	11.4	13.6	16.3	19.8	24.5	29.9	37.0
1000	4.1	4.5	5.1	6.1	7.7	8.8	10.2	11.9	14.3	16.5	20.4	24.8	30.6	37.4	46.3
1250	5.1	5.6	6.4	7.7	9.7	11.1	12.8	15.0	18.0	21.4	25.7	31.2	38.5	47.1	58.2
1600	6.6	7.2	8.2	9.8	12.3	14.1	16.3	19.0	22.9	27.2	32.6	39.7	49.0	59.8	74.0
2000	8.2	9.0	10.2	12.2	15.4	17.7	20.4	23.8	28.6	33.0	40.8	49.6	61.2	74.8	92.5
2500	10.2	11.2	12.8	15.3	19.2	22.0	25.5	29.8	35.7	42.5	51.0	52.0	76.5	93.5	115.7
3200	13.1	14.1	16.0	19.3	24.2	27.3	32.1	37.4	45.0	53.5	64.2	78.0	96.3	117.7	145.5

TABLE 13.9

P_l – power (kW) to raise or lower the burden.

Capacity (t/h)	Elevation (m)										
	5	6.3	8	10	12.5	16	20	25	31.5	40	50
40	0.5	0.7	0.9	1.1	1.4	1.7	2.2	2.7	3.4	4.4	5.5
63	0.8	1.0	1.3	1.7	2.1	2.7	3.4	4.3	5.4	6.8	8.5
100	1.3	1.6	2.1	2.7	3.2	4.3	5.4	6.3	8.6	10.9	13.6
160	2.1	2.7	3.4	4.4	5.5	6.9	8.7	10.9	13.7	17.4	21.8
200	2.7	3.4	4.3	5.4	6.8	8.6	10.9	13.6	17.2	21.8	27.2
250	3.4	4.2	5.4	6.8	8.5	10.7	13.6	17.0	21.4	27.2	34.0
320	4.4	5.4	6.9	8.7	10.9	13.7	17.4	21.8	27.5	34.9	43.6
400	5.4	6.8	8.6	10.9	13.6	17.1	21.8	27.2	34.3	43.5	54.4
500	6.9	8.5	10.7	13.6	17.0	21.4	27.2	34.0	42.9	54.4	68.1
630	8.5	10.6	13.4	17.0	21.3	26.8	34.0	42.5	53.6	68.1	85.0
800	10.9	13.6	17.1	21.8	27.2	34.3	43.5	54.5	68.5	87.0	108.8
1000	13.6	17.0	21.4	27.2	34.0	42.9	54.4	68.1	85.7	108.9	136.1
1250	17.0	21.3	26.8	34.0	42.6	53.6	64.1	85.1	107.1	128.1	170.2
1600	21.8	27.2	34.3	43.6	54.5	68.6	87.1	108.9	137.2	174.2	217.8
2000	27.2	34.0	42.6	54.4	68.1	85.2	108.9	136.1	171.5	217.8	272.2
2500	34.0	42.5	53.6	68.1	85.1	107.1	136.1	170.1	214.2	272.2	340.3
3200	43.6	54.5	68.7	87.2	109.0	137.3	174.4	218.0	274.6	348.8	436.0

TABLE 13.10

Standard motor ratings (kW).

kW	kW	kW	kW
0.37	5.5	45	200
0.55	7.5	55	220
0.75	11	75	250
1.1	15	90	280
1.5	18.5	110	315
2.2	22	132	355
3.0	30	150	400
4.0	37	185	

Worked example

A conveyor is required to handle 1000 t/h of primary crushed iron ore over 350 m, and raise the material by 10 m. The maximum lump size is 200 mm and the density of the ore is 2400 kg/m^3.

The conveyor is to be designed for 1200 t/h, to allow for load variations (see Figure 13.1).

Step 1 – select belt width and speed for required design capacity

- Refer Table 13.4 – minimum belt width is 650 mm when maximum lump size is mixed with 80 per cent fines.
- Refer Table 13.5 – for 650 mm belt width, typical belt speed is 2.5 m/s for iron ore.
- Refer Table 13.3 – for iron ore, angle of surcharge is indicated as 5° to 25°. For primary crushed iron ore, 25° is selected.
- Refer Table 13.1 – for 25° surcharge angle, 650 mm belt width, 35° idler angle, conveyor capacity is 170 t/h at 1 m/s for material of density of 1000 kg/m^3.
- Adjust conveyor capacity for belt speed of 2.5 m/s and material density of 2400 kg/m^3:

$$170 \times (2.5/1) \times (2400/1000) = 1020 \text{ t/h}$$

  This does not meet the 1200 t/h requirement.
- Try next belt width, ie 800 mm.
- Refer Table 13.1 – for 25° surcharge angle, 800 mm belt width and 35° idler angle, conveyor capacity is 266 t/h at 1 m/s for material density of 1000 kg/m^3.
- Adjust conveyor capacity for belt speed of 2.5 m/s and material density of 2400 kg/m^3:

$$266 \times (2.5/1) \times (2400/1000) = 1596 \text{ t/h}$$

  This exceeds the 1200 t/h requirement.
- Adjust belt speed to convey 1200 t/h so that 2.5 × (1200/1596) = 1.87 m/s
- Adopt belt speed as 2.0 m/s.

Step 2 – calculate belt power

- Refer to Table 13.7 for belt width of 800 mm and conveyor length of 350 m:

Pe = 5.1 kW for belt speed of 1 m/s for 315 m length

Increase for 350 m length, say 5.1 × 350/315 = 5.7 kW. Adjust for belt speed of 2.0 m/s:

$$P_e = 5.7 \times 2.0 = 11.4 \text{ kW}$$

- Refer to Table 13.8 for conveyor length of 350 m and capacity of 1250 t/h:

$$P_h = 31.2 \text{ kW for 315 m}$$

Adjust for 1200 t/h and 350 m length:

$$P_h = 31.2 \times (1200/1250) \times (350/315) = 33.3 \text{ kW}$$

- Refer to Table 13.9 for conveyor elevation of 10 m and capacity of 1250 t/h:

$$P_l = 34.0 \text{ kW}$$

Adjust for 1200 t/h:

$$P_l = 34 \times (1200/1250) = 32.6 \text{ kW}$$

- Belt power $P_T = P_e + P_h + P_l$:

$$P_T = 11.4 + 33.3 + 32.6 = 72.8 \text{ kW}$$

- Minimum motor power:

$$Mp = PT/f = 77.3/0.95 = 81.4 \text{ kW}$$

- Refer to Table 13.10 to select motor size from standard motor ratings.

  Select 90 kW motor

- Using a slope of 16° (from Table 13.6), the length of the elevated conveyor is approximately 10/tan 16° = 35 m.

See Figure 13.1 for diagrammatic layout.

Conveyor data summary

Design capacity = 1200 t/h

Belt width = 800 mm

Belt speed = 2.00 m/s

Motor power = 90 kW

Capital cost estimates

Although many details of the final conveyor system are yet to be determined, the preliminary information developed above is used to prepare an order-of-magnitude estimate of the installed cost of the system. Typical costs for the design, supply and installation of plant conveyors are set out in Tables 13.11 to 13.15. These all-up costs are based on labour, material and freight in locations near capital cities, and need adjustment for other sites.

An approximation suggested by Buchanan and Sinclair (1964) is to increase the estimate by one per cent for every 100 miles (160 km) from a capital city. This increase assumes that accommodation and messing facilities are not required for most of the

TABLE 13.11

Allowances for horizontal length of conveyor (coal conveyors).

Belt width (mm)	Ground level ($000/m)	Elevated ($000/m)
650	2.2	4.4
800	2.3	4.9
1000	2.4	6.3
1200	2.5	6.9
1400	2.7	7.4
1600	2.7	7.5
1800	3.1	8.3
2000	3.3	9.0

TABLE 13.12

Allowances for support trestles (coal conveyors).

Trestle height (m)	$000 (each)
2	5.1
5	7.2
10	15.0
15	23.0

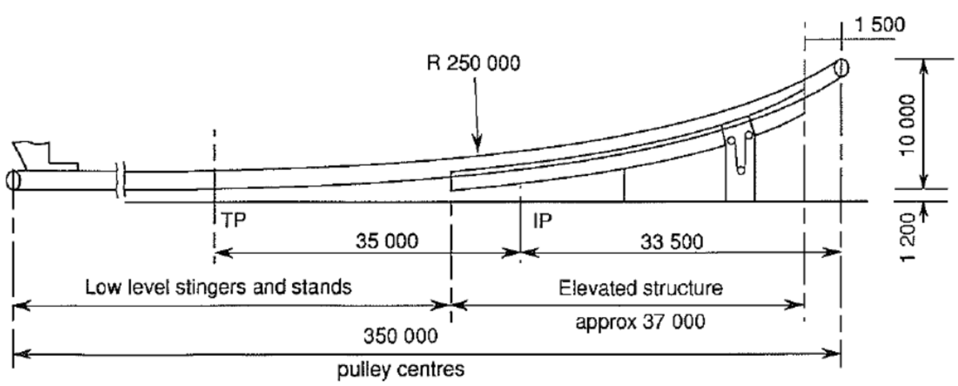


FIG 13.1 - Diagrammatic layout for worked example.

TABLE 13.13
Allowances for drive components (coal conveyors).

Motor size (kW)	$000 (each)	Motor size (kW)	$000 (each)
15	58	150	195
30	66	185	223
45	77	250	271
55	87	315	313
75	103	335	345
90	123	355	381
110	141	385	432
132	162		

TABLE 13.14
Allowances for head and tail assemblies (coal conveyors).

Belt width (mm)	$000 (each)
650	179
800	211
1000	232
1200	275
1400	306
1600	351
1800	412
2000	452
2400	493

TABLE 13.15
Allowances for take-up and transfer towers (coal conveyors).

Height (m)	$000
5	29
10	46
15	73

construction personnel; alternatively, if such facilities are required, the estimate should be increased overall by three per cent.

Basis of estimate

The order-of-magnitude estimate is prepared by adding indicative allowances for the five significant variables for each belt width:

1. horizontal length of conveyor (m), for either ground level or elevated conveyors
2. support trestles for elevated conveyor
3. drive components (motor, gearbox, brake, electricals, etc)
4. head and tail-end assemblies (including take-up system)
5. take-up and transfer towers.

The allowances shown are for coal conveyors and should be increased by up to 20 per cent for heavy-duty use with materials such as iron ore.

These allowances are deduced from the values below. For belt widths and other items not shown, allowances are deduced by interpolation.

Horizontal length of conveyor

See Table 13.11.

Support trestles for elevated conveyors

Support trestles (of A-frame format) are required at say 12 m intervals under elevated conveyors. Determine the approximate heights of the required trestles, and calculate the allowance to be added, as shown in Table 13.12.

Drive components

Allowances for drive components includes allowances for electrics (see Table 13.13).

Head and tail-end assemblies

These allowances are for tail-end pulleys, snub pulleys, head and tail frames, standard head chutes, belt and pulley scrapers with gravity-type take-up assemblies.

The amounts shown in Table 13.14 are for simple systems. Four-pulley take-up systems increase these allowances by 100 per cent.

Take-up and transfer towers

The allowances shown in Table 13.15 include provisions for engineering, design, construction indirects and commissioning, but do not include such items as chutes, hoppers and feeders associated with the use of the installed system, or electric supply to the drive.

Worked example

Using the design example above, the order-of-magnitude estimate is prepared by adding the allowances for:

- Horizontal length of ground level conveyor (800 mm belt)
  315 m @ $2300/m — 724 500
- Horizontal length of elevated conveyor (800 mm belt)
  35 m @ $4900/m — 171 500
- Support trestles for elevated conveyors
  Require three lengths; say 3.3, 6.7 and 10.0 m
  1 × 3.3 m (by pro-rating) — 8415
  1 × 6.7 m — 9648
  1 × 10 m — 15 000
- Drive components
  90 kW motor — 123 000
- Head and tail-end assemblies
  800 mm belt width — 211 000

- Take-up tower
  1 × (say) 5 m height — 29 000
- Order-of-magnitude estimate — = 1291.063
- Above allowances based on coal – add 20 per cent for iron ore — 258 213
- Total estimate — = 1 549 276

Allow $1 550 000

SLURRY TRANSFER AND PUMPING

Heterogeneous and homogeneous slurries are discussed in this section, including pumping systems needed to transport slurries over long distances.

Data required for estimate

Slurry pumping systems comprise pumps, either centrifugal or positive displacement, and in-plant piping or tailings piping. Slurry pipelines to transport solids over long distances are more sophisticated, requiring more detailed evaluation and are excluded from this chapter. To estimate the indicative cost of the system, the following data are required:

- elevation difference between start and end of pipeline – H (m), rise +ve, fall –ve
- median particle size as d_{50} (μm)
- pipeline length (m)
- solids density as specific gravity (sg)
- solids throughput rate (t/h)
- target solids concentration as per cent by weight (CW).

Characterisation of slurry

Pipeline flow of slurries can be classified into two types: heterogeneously flowing and homogeneously flowing. Figure 13.2 illustrates the typical hydraulics of the two types.

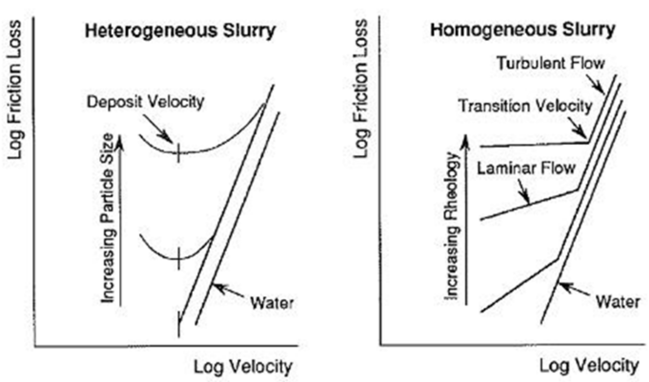


FIG 13.2 - Typical slurry behaviour.

A heterogeneous flow is one for which a certain velocity is required to maintain solid particles in suspension. Below this critical deposit velocity a stationary bed of solids forms. For the present purposes, a heterogeneous slurry is classified as one for which the deposit velocity is greater than 1 m/s. Typical examples are beach sand in water, coarse tailings and lump coal.

A homogeneous slurry is defined as one for which the deposit velocity is below 1 m/s or for which no deposition occurs at all. The flow behaviour depends on slurry rheology (viscosity). In laminar flow, there is an absence of turbulence to maintain solids suspension. It is desirable to operate in turbulent flow; that is, at a velocity above the transition velocity to avoid deposition.

Slurry behaviour type, heterogeneous or homogeneous, is largely determined by the particle size and solids sg. Other factors such as slurry rheology and pipe size also influence behaviour. For an initial assessment the dJ0 (median) particle size is used to characterise the slurry.

Figure 13.3 presents the demarcation between heterogeneous and homogeneous slurries as a function of particle size and solids sg. The hatched area allows for typical variations in rheology and pipe size.

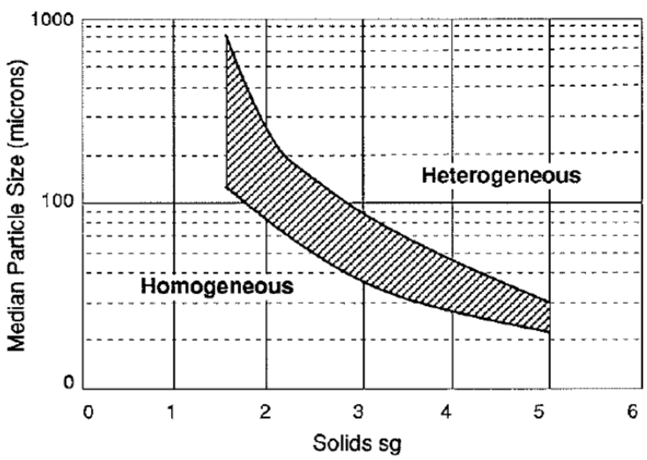


FIG 13.3 - Characterisation of slurries.

Operating in this velocity range also results in minimum pipe wear. If the deposit velocity is above this optimum range, the operating velocity must be higher and steel pipes will wear.

For preliminary purposes, the approximate pressure gradient is calculated to within a ±15 per cent accuracy. For homogeneous slurries refer to Table 13.16 and the example in Figure 13.5. For heterogeneous slurries refer to Figure 13.4 and the example in Figure 13.6.

Pumping pressure

Having obtained the pressure gradient, the pump pressure is calculated as follows:

$$P = J \times L/1000 + (S.sg) \times 9.81 \times H$$

where:

P	is the pumping pressure in kPa
J	is pressure gradient in Pa/m (or kPa/km)
L	is pipeline length in metres
S.sg	is slurry specific gravity in g/mL
H	is the static lift in metres from the pump to the pipeline exit

TABLE 13.16
Slurry systems – homogeneous correlation.

Slurry pressure drop factor F1							
Velocity (m/s)	Slurry rheology high	Slurry rheology normal	Slurry rheology low	Velocity (m/s)	Slurry rheology high	Slurry rheology normal	Slurry rheology low
1.2	15.8	13.0	10.4	3.0	79.2	67.2	56.0
1.3	18.1	15.0	12.0	3.2	88.8	75.5	63.2
1.4	20.6	17.1	13.7	3.4	99.0	84.3	70.7
1.5	23.2	19.3	15.6	3.5	104.0	88.7	74.5
1.6	26.0	21.7	17.5	3.6	109.6	93.5	78.7
1.7	28.9	24.1	19.6	3.8	120.7	103.2	87.1
1.8	32.0	26.7	21.8	4.0	132.3	113.3	95.9
1.9	35.2	29.5	24.0	4.2	144.5	123.8	105.1
2.0	38.5	32.3	26.4	4.4	157.0	134.8	114.6
2.1	42.0	35.3	28.9	4.5	163.1	140.2	119.4
2.2	45.6	38.3	31.5	4.6	170.0	146.2	124.6
2.3	49.3	41.5	34.2	4.8	183.5	158.1	135.1
2.4	53.2	44.8	37.0	5.0	197.5	170.3	145.9
2.5	57.2	48.2	39.9	5.2	211.9	183.0	157.1
2.6	61.4	51.9	42.9	5.4	226.8	196.1	168.7
2.7	65.7	55.5	46.1	5.5	234.0	202.5	174.3
2.8	70.1	59.3	49.3	5.6	242.2	209.7	180.7
2.9	74.6	63.2	52.6	5.8	258.0	223.7	193.1
3.0	79.2	67.2	56.0	6.0	274.3	238.0	205.9
Transition velocity factor F2 (Laminar/turbulent transition flow)					1.98	0.92	0.34
Typical rheology classification			Yield stress (Pa)		10 - 20	1 - 4	0.1 - 0.5
			Plastic viscosity (mPa.s)		30 - 50	15 - 25	3 - 10

For pressures less than about 4500 kPa, centrifugal pumps in series are suitable. Higher pressures will generally require positive displacement pumps; these are piston, plunger or diaphragm types. Pump power is calculated as follows:

$$\text{Power (kW)} = \text{P (kPa)} \times \text{Q (m}^3\text{/s)} / \text{efficiency}$$

where:

Q is volumetric flow rate

P is the pumping pressure in kPa

The efficiency of a centrifugal pump is given by pump curves and is usually in the range 0.6 to 0.7. The efficiency of a positive displacement pump is typically about 0.85.

Pipe wear and pipe selection

Prediction of pipe wear is complex and beyond the scope of this section. Unlined steel pipe is generally used for slurries classified as homogeneous in Figure 13.3, providing the velocity is less than about 2 m/s. In general, slurries classified as heterogeneous in Figure 13.3 will require wear-resistant pipe, especially if the velocity is high. If the particles are sharp-edged, hardened pipe may be required, but in most cases soft resilient linings such as rubber or polyurethane give suitable wear resistance. A popular pipe material for low-pressure applications is high-density polyethylene (HDPE). Typically, polyurethane and rubber are six to eight times more wear-resistant than steel and HDPE is three to four times. For initial assessment purposes, the best approach is to select the pipe material known to be used in similar industries.

Homogeneous slurries – calculation steps

Step 1 Calculate volumetric flow rate, Q (m^3/h) Q = t/h × (1/sg + 100/CW - 1).

Step 2 Calculate velocities for various pipe diameters that give velocities in the range of interest, say 1.5 to 3 m/s.

Solids sg	Velocity (m/s)	Particle d50 Size (Microns) 100	150	200	250	300	400	500	Pipe Dia. (mm)
Solids sg 1.40	4.0					0.043	0.56	2.1	
	3.5				0.013	0.11	1.07	3.4	
	3.0				0.048	0.30	2.05	5.5	Pipe Dia.
	2.5			0.012	0.17	0.79	3.9	8.9	(mm) 600
	2.0			0.072	0.61	2.1	7.5	14.4	250
	1.5		0.15	0.44	2.2	5.5	14.3	23.3	150 50
Solids sg 2.00	5.5			0.011	0.14	0.59	2.9	6.6	
	5.0			0.025	0.25	0.94	4.0	8.5	Pipe Dia.
	4.5			0.058	0.46	1.5	5.5	10.9	
	4.0			0.13	0.83	2.4	7.6	13.9	(mm)
	3.5		0.011	0.30	1.5	3.8	10.5	17.8	600
	3.0		0.042	0.70	2.75	6.1	14.5	22.8	
	2.5		0.15	1.6	5.0	9.7	20.0	29.1	250
	2.0		0.56	3.65	9.1	15.5	27.6	37.3	150
	1.5	0.042	2.05	8.35	16.6	24.7	38.0	47.7	50
Solids sg 2.70	6.0			0.16	0.89	2.4	7.3	13.2	Pipe Dia.
	5.5		0.012	0.27	1.30	3.3	9.1	15.6	(mm)
	5.0		0.027	0.47	1.95	4.5	11.3	18.4	
	4.5		0.062	0.80	2.90	6.1	14.0	21.8	600
	4.0		0.14	1.40	4.30	8.4	17.5	25.9	250
	3.5		0.32	2.35	6.35	11.4	21.7	30.6	150
	3.0		0.73	4.0	9.4	15.5	27.0	36.3	
	2.5	0.035	1.65	6.9	14.0	21.2	33.6	43.0	
	2.0	0.17	3.7	11.7	20.7	28.9	41.8	50.9	50
	1.5	0.85	8.5	20.0	30.7	39.4	52.0	60.2	
Solids sg 4.00	6.0		0.22	1.70	4.7	8.8	17.5	25.5	Pipe Dia.
	5.5		0.36	2.35	6.1	10.7	20.2	28.5	(mm)
	5.0		0.60	3.3	7.9	13.1	23.4	32.0	600
	4.5	0.018	1.0	4.6	10.1	16.1	27.1	35.8	250
	4.0	0.047	1.7	6.5	13.1	19.7	31.3	40.2	150
	3.5	0.12	2.8	9.2	16.8	24.1	36.2	45.0	
	3.0	0.32	4.7	12.9	21.7	29.6	41.8	50.4	
	2.5	0.83	7.8	18.2	28.0	36.2	48.8	56.5	50
	2.0	2.2	12.9	25.6	36.1	44.4	56.0	63.4	
	1.5	5.6	21.6	36.0	46.6	54.4	64.7	71.0	

FIG 13.4 - Heterogeneous slurry correlation pressure drop factor F3.

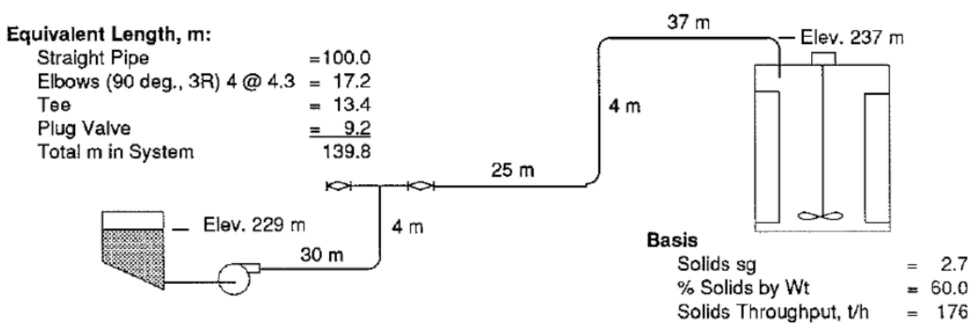


FIG 13.5 - System sketch.

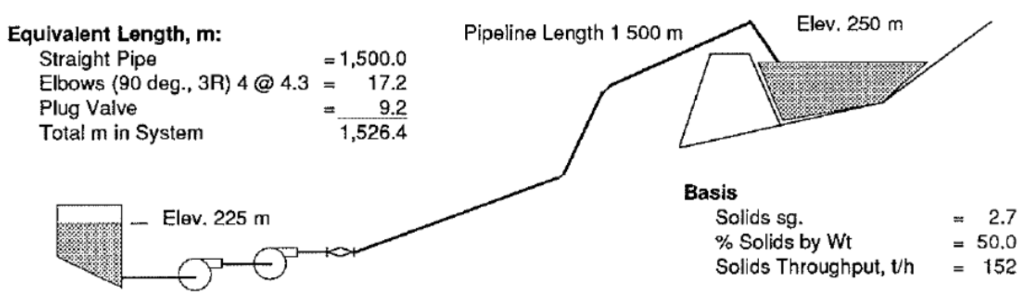


FIG 13.6 - System sketch.

Step 3 Decide likely rheology range: high, normal or low. As a guide, low rheology is approaching water and high rheology is a high-density thickener underflow slurry (in general, similar rheology to a 24-hour bench-settling test).

Step 4 Calculate transition velocity (VT) for the various diameters using factor F2 (Table 13.16).

Step 5 Select the pipe size giving an operating velocity just above the VT.

Step 6 Calculate the pressure drop J (Pa/m) using factor F1 (Table 13.16).

Heterogeneous slurries – calculation steps

Step 1 Calculate volumetric flow rate, Q (m^3/h) Q = t/h × (1/sg + 100/CW - 1).

Step 2 Calculate velocities for various pipe diameters that give velocities in the range of interest, say 1.5 to 6 m/s.

Step 3 Using Figure 13.4, select pipe size that gives operating velocity above limits given for relevant pipe size (interpolate as required).

Step 4 Calculate heterogeneous pressure gradient using F3 (Figure 13.4) and F1 (Table 13.16, low rheology).

$$J_{het} = J_{hom} + F3 \times (sg - 1) \times C/(sg - C \times (sg - 1)/100)$$

where:

I = F1 (low rheology) × SI.sg × $D^{-1.25}$/1.22

Indicative capital cost	**$**
Piping (Table 13.17)	
Materials supply	17 192
Installation 100 m @ $363/m	36 300
Pump	
Equipment supply	
Closest centrifugal pump 25 kW	82 000
Installation (F4 = 0.5) Cost = 0.5 × 82 000	41 000
Total indicative cost	**$176 492**

Indicative capital cost	**$**
Piping (Table 13.18)	
Materials supply	70 631
Installation 1600 m @ $116/m	185 600
Pump	
Equipment supply	
Closest pump two in series 100 kW ea	86 000
Installation (F4 = 1.0) Cost = 1.0 × 86 000	86 000
Total indicative cost	**$428 231**

Example 1 – homogeneous slurry

Basis for Example 1

d_{50} particle size (microns) = 31 — Slurry rheology – normal

Pipe diameter trial basis (mm) = 203

- Characterisation of slurry

  Plot d_{50} = 31 microns versus solids sg = 2.7 on Figure 13.3
  Answer is 'homogeneous slurry'

- Calculate volume flow rate, Q (m^3/h)

  Q = 176 × ((1/2.7) + (100/60) - 1)
  = 182.5

- Calculate laminar/turbulent transition velocity (m/s) VT

  Slurry sg (Sl.sg) = 100/((60/2.7) + (100 - 60))
  = 1.607

  Select factor F2 from Table 13.16 for 'normal rheology'

  F2 = 0.92

  VT = 0.92 × ((0.203) ^ (0.3) × ((1.607) ^ (-0.65))
  = 1.09

- Calculate actual velocity (m/s)

  Velocity = (182.6/3600) / (π × (0.203) ^ (2/4))
  = 1.57 (safe in turbulent flow)

- Calculate slurry pressure drop J (Pa/m)

  Interpolate factor F1 from Table 13.16 for 'normal rheology'

  F1 = 24.1 - (24.1 - 21.7) × (1.7 - 1.67)/(1.7 - 1.6) = 23.38

  J = 23.38 × 1.607 × (0.203 ^ (-1.25))
  = 275.7 Pa/m

- Calculate pumping pressure (kPa)

  P = 275.7 × 139.8/1000 + 1.607 × 9.81 × (237-229)
  = 38.5 + 126.1 = 164.6 kPa

- Calculate pump power (kW)
  (Assumed efficiency 65%)

  Power = 164.6 × (182.5/3600)/0.65
  = 12.8 kW

TABLE 13.17
Piping materials supply (for Example 1 – homogeneous slurry).

Item	Quantity	Unit	$/Unit	Cost ($)
Plain steel standard wall (API 5LB ERW)	100	m	70.2	7020
Pipe fitting allowance	100	m	20.3	2030
Slurry valve ANSI 150#	2	ea	4071	8142
Total piping materials supply				17 192

TABLE 13.18
Piping materials supply (for Example 2 – heterogeneous slurry).

Item	Quantity	Unit	$/Unit	Cost ($)
HDPE pipe (Class 9) (SDR11 PN 16)	1600	m	21.3	34 080
Pipe fitting allowance	1600	m	20.3	32 480
Slurry valve ANSI 150#	1	ea	4071	4071
Total piping materials supply				70 631

Indicative cost

The indicative capital cost for a slurry transfer and pump system is readily determined from the cost indicators, based on nominal pipe size and pump power, summarised in Table 13.19.

These indicative costs are based on December 2010 costs in Australian dollars.

THICKENERS AND CLARIFIERS

Costings for clarifiers and thickeners, including high-rate thickeners (HRTs), are considered in this section. The two most common methods for testing and sizing thickeners for scale-up are considered.

Conventional thickeners and clarifiers

Usually, the term 'thickener' is used to describe a sedimentation device whose primary objective is to produce a settled pulp of increased pulp density, with the separated solution as the overflow. The term 'clarifier' is applied to a device whose primary objective is to clarify the overflow solution, with the thickener underflow being of secondary importance.

Example 2 – homogeneous slurry

Basis for Example 2

d_{50} particle size (microns) = 167 — Slurry rheology – low

Pipe diameter trial basis (mm) = 203

- Characterisation of slurry — Plot d_{50} = 165 microns versus solids sg = 2.7 on Figure 13.3
  Answer is 'homogeneous slurry'
- Calculate volume flow rate, Q (m^3/h) — Q = 152 × ((1/2.7) + (100/50) - 1)
  = 208.3
- Calculate actual velocity (m/s) — Velocity = (208.3/3600) / (π × (0.203) ^ (2/4))
  = 1.79 (safe in turbulent flow)
  (see Figure 13.4, velocity is below deposition; choose smaller pipe diameter)
  Select 200 mm HPDE class 9 I.D = 163 mm
  Velocity = (208.3/3600) / (π × (0.163) ^ (2/4))
  = 2.77 (safe – see Figure 13.4)
- Calculate homogeneous slurry pressure drop J_{hom} (Pa/m) — Slurry sg (Sl.sg) = 100/(50/2.7) + (100 – 50) = 1.459
  Interpolate factor F1 from Table 13.16 for 'low rheology'
  F1 = 49.3 – (49.3 – 46.1) × (2.8 – 2.77)/(2.8 – 2.7)
  = 48.34
  J_{hom} = 48.34 × 1.459 × (0.163) ^ (-1.25)/1.22
  = 558.2 Pa/m
- Calculate heterogeneous slurry pressure drop J_{hom} (Pa/m) F3 (interpolate from Figure 13.4) = 0.73 + (1.65 – 0.73) × (3.0 – 2.77)/(3.0 – 2.5) + (4.0 – 0.73) × (200 – 167)/(200 – 150) = 3.31 J_{het} = 558.2 + 3.31 × (2.7 – 1) × 50/(2.7 – 50 × (2.7 – 1)/100) = 558.2 + 152.2 = 710.3 Pa/m
- Calculate pumping pressure (kPa) P — P = 710.3 × 1626.8/1000 + 1.459 × 9.81 × (250 - 225)
  = 1155.5 + 357.8 = 1513.3 kPa
- Calculate pump power (kW) (Assumed efficiency 65%) — Power = 1513.3 × (203.3/3600)/0.65
  = 134.7 kW

TABLE 13.19
Capital cost indicators for slurry transfer and pump systems.

		Cost indicators (A$/unit)								
	NPS mm	50	75	100	150	200	250	300	400	600
Description	Unit									
Piping										
Materials supply										
Plain steel std wall (API 5LB ERW)	m	9	18.6	26.5	46.6	70.2	99.5	121.9	N/A	246.8
Plain steel std wall (ATSM A106B Seamless)										
Steel with 6.35 mm rubber lining	m	9.3	19.2	27.3	48	72.3	105.5	129.3	N/A	310.2
	m	257	315	367	442	564	660	727	917	1395
Steel with 6.35 mm polyurethane	m	440	480	525	610	745	860	975	1280	1965
HDPE pipe (Class 9) (SDR11 PN 16)	m	3.2	4.4	6.5	13.6	21.3	33	52.6	84.7	209
Pipe fitting allowance	m	2.7	4.3	6.0	10.6	20.3	47.7	61.3	114.1	276
Slurry valve ANSI 150#	ea	N/A	2712	2933	3381	4071	6038	9898	14 952	32 620
External tape coating	m	8.9	14.1	16.6	24.4	32.8	40.8	49.1	65.3	98
Installation										
In-plant piping	m	131	174	199	271	363	499	613	1004	1702
Tailing										
Mechanical jointed	m	32	48	61	89	116	145	170	221	323
Welded construction	m	22	34	43	63	82	102	119	153	230
	Indicated	Cost indicators (A$1000/unit)								
	Power (kW)	25	50	100	200	300	500	750	1000	1400
Description	Unit									
Pumps										
Equipment supply										
Centrifugal slurry pump	kW	19	28.5	43	65	82	102	143	150	209
Positive displacement plunger pump	kW	48	94	187	375	562	936	1405	1873	2622
Positive displacement diaphragm pump	kW	90	179	358	715	1073	1788	2673	3575	5005
Pump station installation	Factor F4			Formulas						
Installed in existing building	0.5			Pump station installation						
Installed on mine site	1.0			Cost = F4 × pump supply cost						
Installed at remote sites	1.5									

N/A = not applicable.

In most cases, the amount of underflow from a clarifier is relatively small. A 'conventional' thickener or clarifier is usually understood to be a cylindrical tank with a central feedwell and a raking mechanism powered by a central or peripheral drive. The following analysis is confined to this type of configuration.

Thickener test work and sizing

The two laboratory methods that are most commonly used for scale-up of conventional thickeners are the Coe and Clevenger method and the Talmage and Fitch method, which derives from the theory of Kynch. Both methods have as their basis the observation that during sedimentation, a suspension passes through a series of distinct phases, as follows:

- free settling, in which no interference between particles or flocs occurs
- hindered ('zone') settling, in which interaction between particles or floes influences the settling rate

- compression settling, during which the interaction between particles or floes is the determining influence on settling rate.

In passing through these phases, the 'flux' of the suspension, defined as the solids mass settling rate, usually expressed in t/m^{2}h, varies mainly as a function of solids concentration. In considering the operation of a continuous thickener, these zones exist at different levels and the solids pass through each zone in turn as they settle towards the thickener base. The solids throughput is limited by the 'critical concentration', which is the point of minimum flux at the top of the compression zone. This minimum flux can be identified by the two static settling test methods referred to above, and is used to determine the area required for settling.

Coe and Clevenger method

The suspension to be tested is made up of a range of solids concentrations and this is used as the starting concentration for each settling test. The concentration that results in the lowest solids flux is identified by this means as the critical concentration, and the minimum flux thus determined is the required scale-up parameter.

Worked example

A slurry sample is received at a solids concentration of 150 g/L. Thickener area is to be determined to handle a solids rate of 80 t/h, and thicken it to a density of 30 per cent w/w without the use of flocculant.

Procedure

Five 2 L cylinders are filled with sample, agitated and allowed to settle. When the solids have settled below the 600 mL level in each cylinder, the supernatant is removed and the settled solids transferred to a 1 L cylinder. The five samples are then made up to 600, 700, 800, 900 and 1000 mL, respectively, by adding supernatant back into the cylinder. Hence, five samples are obtained with different starting concentrations, but with the same mass of solids in each. Each sample is agitated, and the settling interface versus time profile recorded. A typical series of settling curves is shown in Figure 13.7.

These curves are essentially straight lines and hence a settling rate is calculated for each solids concentration. In a thickener of cross-sectional area A, the solids settling rate in a zone of solids concentration C_c is expressed as:

$$U_c = C_c \times a$$

where:

U_c settling rate in m/h

C_c solids concentration in g/mL

a cross-sectional area in m^2

A mass balance at steady state yields the result:

$$Q \times C_u = Q \times C_c + U \times C \times A \quad (13.1)$$

where:

Q bulk volume flow in m^3/h

C_U solids concentration in thickener underflow in g/mL

Hence:

$$Q/A = U_c \times C_c/C(C_u - C_c)$$

if:

$$G = Q \times C_u/A \quad (13.2)$$

where:

G solids flux in t/m^{2}h

then:

$$G = U_C/(1/C_C - 1/C_U) \quad (13.3)$$

From settling rate versus concentration data shown in Table 13.20.

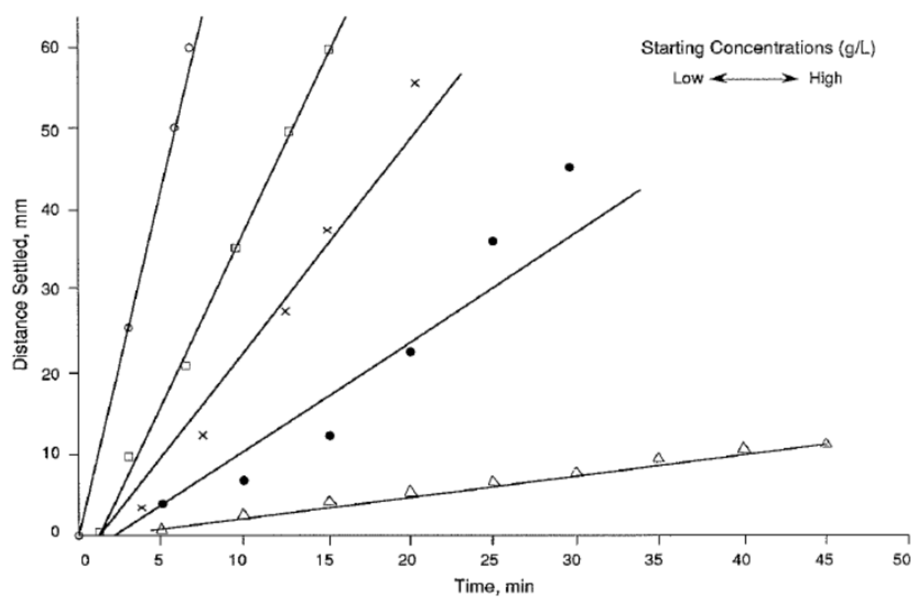


FIG 13.7 - Typical settling curves (Coe and Clevenger method).

TABLE 13.20
Solids flux determination (Coe and Clevenger method).

U_c (m/h)	$1/C_u$ (mL/g)
0.266	13.20
0.137	7.35
0.080	5.88
0.040	4.41
0.007	2.94

Plotting U_c versus $1/C_u$ gives the curve shown in Figure 13.8.

It is required to thicken the solids to 30 per cent w/w, which is equivalent to 2.58 mL/g (solids sg = 4.0). Drawing a straight line through a point at 2.58 on the horizontal axis tangential to the underside of the curve and applying Equation 13.3 above yields:

$$G = 0.1/(8.40 - 2.58) = 0.0172 \text{ t/m}^2\text{h}$$

For sizing purposes, a conventional safety factor of 1.2 is used. Hence, for a feed rate of 80 t/h, the calculated thickener area would be:

$$A = (80)(1.2)/0.0172 = 5581 \text{ m}^2$$

Therefore, an 85 m diameter thickener is recommended for this application.

Talmage and Fitch method

Using the Talmage and Fitch method, only a single settling test is required. The zone interface is plotted against time, as shown in Figure 13.9.

From the settling data, it is sometimes possible to identify the critical concentration point or 'compression point' as a discontinuity in the settling curve. If this is not evident, the alternative method is to bisect the angle made by the straight-line portions of the settling curve and where the bisecting line crosses the curve, then assume this to be the compression point. A tangent is then drawn to the curve at the compression point. For a nominated underflow density, the corresponding settled volume is calculated and a horizontal line drawn through the vertical axis at this point. The required settling time is then obtained at the intersection of this line and the tangent to the curve.

For a sample with solids mass w in grams, the solids flux is calculated from the expression:

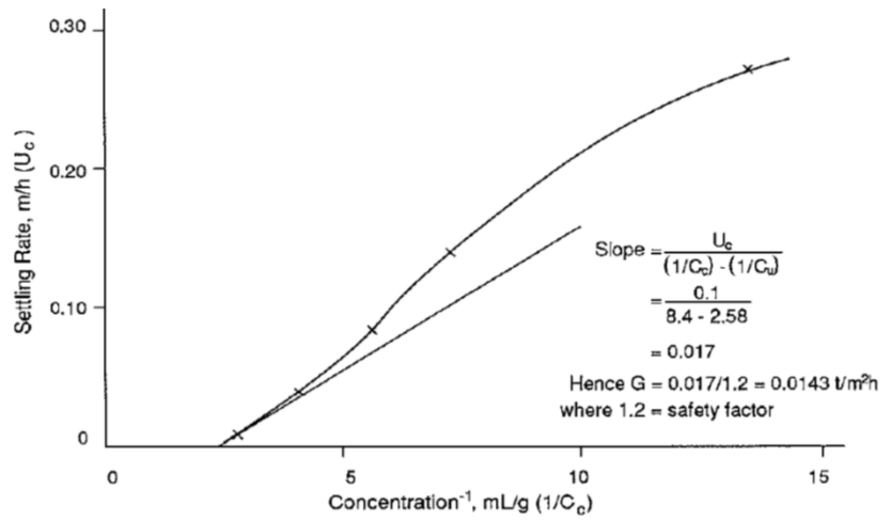


FIG 13.8 - Solids flux determination (Coe and Clevenger method).

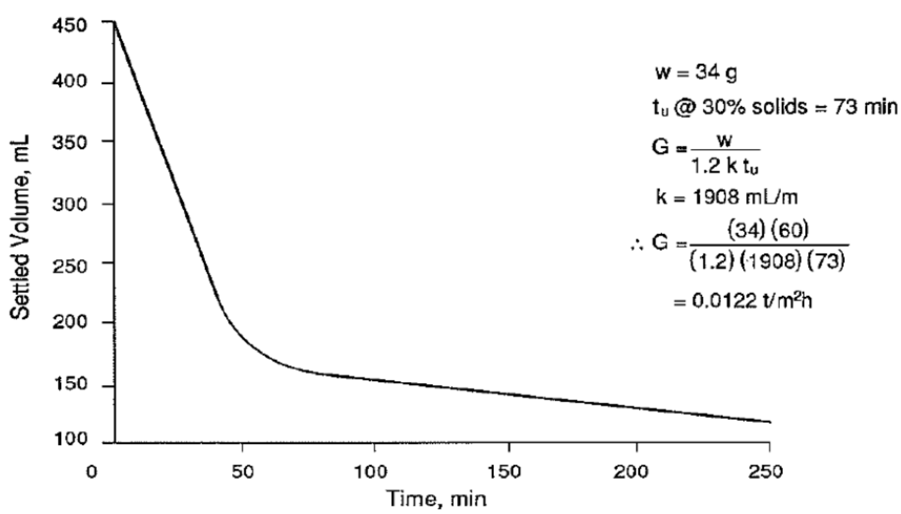


FIG 13.9 - Typical settling curve (Talmage and Finch method).

$$G = w/(k \times t_u) \quad (13.4)$$

where:

k cylinder factor in mL/m

t_u settling time in hours

For the previously worked example:

w 34 g

k 1908 mL/m

t 73 min

t_u 1.22 h

From Equation 11.4, G = 0.0146 t/m²h; hence, thickener area required:

$$A = (80)(1.2)/0.0146 = 6525 \text{ m}^2$$

Therefore, a 92 m diameter thickener is indicated by this method of analysis.

Although both methods are based on the same principle, which is identifying the zone of minimum flux and calculating the solids flux in that zone, it is common for the Coe and Clevenger method to produce a lower calculated area than Talmage and Fitch. Hence, the latter method is more conservative and is simpler. It has a further advantage when flocculant is used; a more realistic settling regime is obtained because no admixing of settled pulp and supernatant is necessary.

For these reasons, the Talmage and Fitch method (or variations) is the most commonly used technique for the determination of conventional thickener area requirements.

Clarification

The two factors used to assess the clarification of solids suspensions are the rise rate and detention time. These are expressed as follows in relation to the area of a clarifier:

$$U = Q/A \quad (13.5)$$

where:

U rise rate or velocity in m/h

Q volumetric overflow rate in m³/h

A area of clarifier in m²

and:

$$t = AH/Q \quad (13.6)$$

where:

t detention time in h

H depth of clarification zone in m

In many cases, the detention time is the most important factor for sizing of conventional clarifiers. The simplest procedure for determining detention time is as follows:

1. Fill a 1 L measuring cylinder with sample.
2. Add flocculant if required, invert the cylinder several times to achieve good mixing, and start timing when mixing is complete.
3. Withdraw 100 mL subsamples at a depth of about 50 mm below the surface at appropriate intervals (eg every five minutes), until supernatant clarity is acceptable.
4. Determine the suspended solids levels in the subsamples and plot these against time, as shown in a typical plot in Figure 13.10.
5. Determine retention time to achieve acceptable clarity.

In calculating the clarifier volume from static tests, as described above, it is normal practice to apply a safety factor of three to four to allow for dynamic effects, such as circulating currents and short-circuiting. It is also advisable to use a bulk settling test to check that clarifier rise rate is within acceptable limits. This is done by allowing the suspension to settle until the suspension approaches the 'zone settling' condition; that is, there is a marked difference between the clear supernatant and the settling solids.

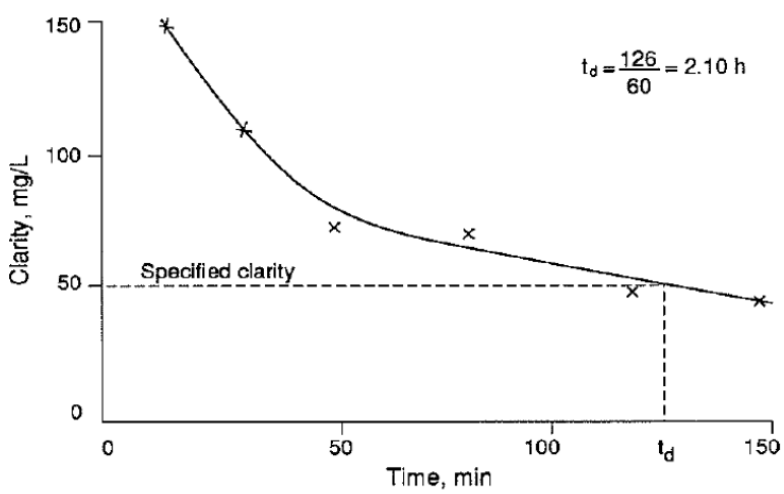


FIG 13.10 - Determination of retention time.

The supernatant is then decanted off, the remaining solids flocculated if necessary, and added to a 500 mL cylinder.

The cylinder is inverted several times and the suspension allowed to settle. The bulk settling interface level is then recorded against time, as shown in Figure 13.11. The bulk settling velocity U_b is then determined, as shown below. This represents the maximum permissible overflow rate in the clarifier.

The determined values of detention time t_d and bulk settling velocity U_b are then used with Equations 13.5 and 13.6 to determine the clarifier configuration:

For the example shown:

$$AH/Q = t = 4\ t_d$$

where:

4 safety factor and $AH = 4Q_{td}$

$$Q/A = U = 0.5U_b$$

where:

0.5 safety factor

A $Q/0.5\ U_h$

If design overflow rate is 50 m^3/h, A = 159 m^2 and H = 2.64 m.

Hence, the selected configuration is a 15 m diameter clarifier with a 3 m sidewall.

It can be appreciated that the above static settling tests approximate a dynamic operating system, and the results should be treated with caution. This is the reason for the large safety factors usually applied when scaling up. A preferred method of determining thickener or clarifier area is to perform dynamic bench-scale tests on a small thickener unit.

This approach is generally used for sizing HRTs, described in more detail below. However, it is also used with confidence to determine conventional clarifier areas by operating the test unit with a low bed level. Similarly, it is used to determine conventional thickener areas although underflow density is usually understated because of the limited compression zone in the test unit.

Rake mechanisms

The rake blades are configured to rake the settled solids towards the central discharge. Rake blades are supported on rake arms, which are usually of tube, channel or truss design. A triangular tube section, tapered towards the ends, minimises viscosity-induced torque. Usually two long arms and two shorter arms are used, although two arms only are used when the amount of solids being moved is small.

The rake arms are supported by a central shaft or cage, depending on the size of the machine. On thickeners up to about 50 m in diameter, the rake mechanism is usually supported from a full-span bridge, which rests on the tank wall. The bridge is of beam construction up to about 15 m, and truss construction above this size. Above 50 m, a centre column or pier is used to support the mechanism. This is of concrete or steel construction. A rotating cage supports the rake arms in this case and the solids are raked into a trench set around the base of the column. An access walkway is provided from the tank wall to the centre column for operator and maintenance access. On very large thickeners, it is common to use a hollow centre column known as a 'caisson', and to mount the underflow pumps inside the caisson. The underflow pipes are run up the centre column and along the access way.

Thickener drives

For units up to 40 m in diameter, the most common drive configuration is a single planetary gearbox with a fixed or variable speed electric motor. For larger units, with cage-type rake arm supports, a ring gear is usually used with two or more pinion drives, each of which has a planetary stage attached.

Hydraulic drives are also used and are becoming more popular because of a number of perceived advantages, including more accurate torque control and fail-safe operation. A hydraulic power pack drives a hydraulic motor that is close-coupled to a planetary gearbox as the final drive element. This arrangement is used up to

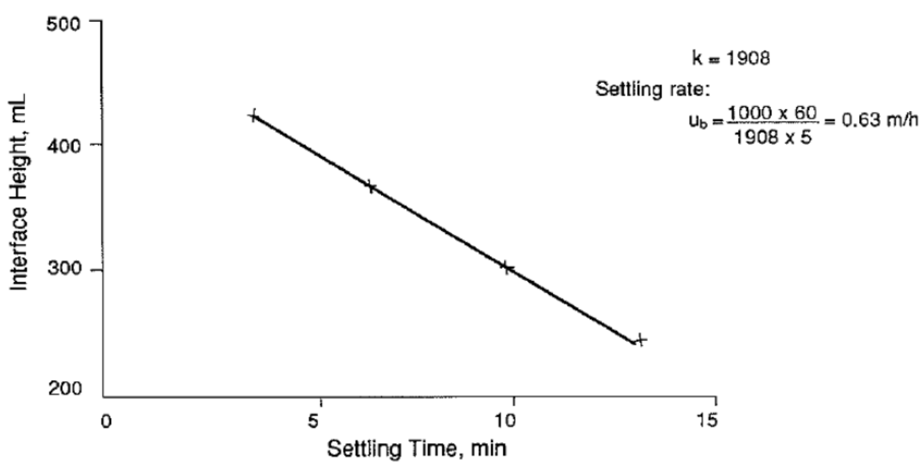


FIG 13.11 - Determination of bulk settling velocity.

a thickener size of about 40 m, above which a ring gear is required.

Most thickener units are fitted with a rake-lifting facility to overcome high torque conditions. For electro-mechanical drives, the rake-lifting mechanism is usually a screw jack driven by an independent electric motor. The lift is triggered by a high torque signal and limited by upper and lower limit switches. Where a hydraulic drive is used, the power pack is used to operate the lifting mechanism by means of hydraulic cylinders.

On conventional thickeners, torque selection is based on experience, usually expressed as a 'k-factor', used in the following expression:

$$T = kD^2 \tag{13.7}$$

where:

T torque (Nm)

D diameter (m)

The range of k-factors used is as follows:

Duty	**k-factor**
Heavy	300
Medium	225
Light	150

The torque rating of the drive must be known in order to establish the cost of the mechanism.

High-rate thickeners

Principles of operation, performance and testing of HRTs is presented in this section.

Test work and sizing

The method generally used to determine sizing and expected performance of HRTs is to carry out dynamic bench-scale test work in a scale model. A sample of 70 - 100 L of feed pulp is required for testing, with information on the required throughput, underflow density and overflow clarity.

Subsamples of 500 mL are usually taken to select a suitable flocculant. This is done by dosing a set of 500 mL samples with a given amount of a range of flocculants, and observing floe formation, initial settling rate and supernatant clarity. During these preliminary settling tests, the influence of sample dilution is also assessed, since floe formation is critically affected by pulp density in most cases. This is particularly true of fine solids; in some cases flocculation does not occur at all unless the pulp is diluted. On completion of these tests, the appropriate flocculant, dosage range and feed pulp density is established for the dynamic test work.

The procedure used for the dynamic test work is as follows:

1. The sample is kept agitated in a 70 L drum, and the feed drawn by a variable speed peristaltic pump.
2. Flocculant at 0.025 per cent concentration is injected at the delivery of the feed pump by a variable speed diaphragm dosing pump.
3. The flocculated feed enters the feed well of the test unit, which has a diameter of 90 mm and incorporates all the features of a full-scale HRT. A feed rate is selected that produces stable operation and acceptable overflow clarity.
4. At the selected feed rate the flocculant dose rate is varied during the test run and the underflow density measured at each set of conditions. The overflow clarity is observed and/ or sampled for clarity determination.
5. Two or more other feed rates are used to provide data at a range of feed rates and flocculant rates.

On completion of the tests, the data are tabulated or graphed, as shown in Table 13.21 and Figure 13.12.

In this manner, a suitable set of parameters is selected for scale-up. In the example given, the following design parameters are selected for scale-up:

Feed rate	0.90 t/m²h
Flocculant dosage	10 - 15 g/t
Underflow density	55 per cent w/w solids
Overflow clarity	Satisfactory

For a solids rate of 80 t/h, the required thickener area is 89 m², and a 12 m diameter thickener is selected.

Rake mechanisms and feedwells for high-rate thickeners

For HRTs, the rake configuration is similar to conventional thickeners, except that for a given size of thickener, the rake torque capability is usually significantly greater because more solids have to be moved and the bed depth is greater. The feedwell in a HRT is designed to achieve three objectives: to de-aerate the feed pulp, achieve effective flocculation and dissipate the energy of the feed before it exits the feedwell. Also, when dilution of the feed is required, a self-diluting system is incorporated by those manufacturers who possess the necessary technology. This feature eliminates the need for an external pumping system to recirculate thickener overflow.

Thickener control

HRTs are usually supplied with a bed-level control system and sometimes a solids inventory control system. These allow the thickener to be fully automated, which is desirable for high-rate units because of the rapid response times required.

Rake drive and lifting

As described for conventional thickeners, the rake drive is either electro-mechanical or hydraulic. Hydraulic drives are favoured for high-rate thickeners because of the advantages previously described. Most HRTs are fitted with rake-lifting mechanisms.

TABLE 13.21
High-rate thickener test results.

Suspension description: solution sg = 1.025			Milled ore primary orebody Solids sg = 2.70 pH neutral					Test unit 90 mm		
Test	Feed			Flocculant type = Magnafloc 333				Underflow		Overflow
Test	Flow	Solids rate	Solids	Density	Flow	Cone	Addition rate	Solids	Density	Clarity
No	(mL/min)	(t/mh)	(%)	(kg/L)	(mL/min)	(%)	gA	(%)	(kg/L)	
1	690	1.31	17.5	1.15	9.0	0.025	16.0	53.9	1.54	Clear
2	670	1.29	17.5	1.15	16.0	0.025	30.0	57.3	1.59	Clear
3	1000	1.93	17.5	1.15	15.0	0.025	19.0	52.5	1.52	Clear
4	1000	1.93	17.5	1.15	22.0	0.025	27.0	56.0	1.57	Clear
5	1000	1.93	17.5	1.15	10.5	0.025	13.0	51.0	1.50	Cloudy
6	370	0.71	17.5	1.15	4.0	0.025	13.0	57.9	1.60	Clear
7	370	0.71	17.5	1.15	2.5	0.025	8.0	56.5	1.58	Clear
8	370	0.71	17.5	1.15	1.3	0.025	4.0	44.0	1.41	Clear

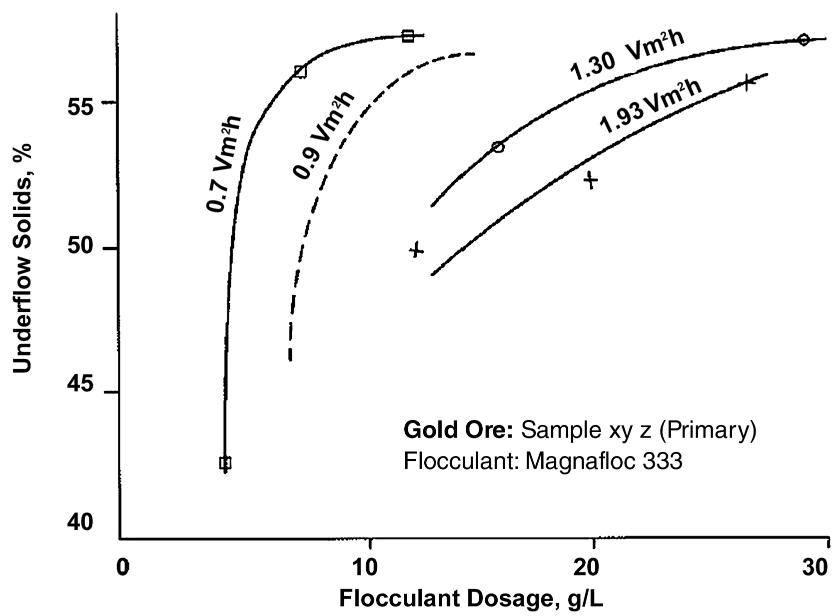


FIG 13.12 - High-rate thickener test results.

Torque

The torque rating on HRTs is usually selected on the basis of manufacturer experience, although a slump test and sliding friction test can be used to calculate torque empirically. Relating the torque ratings of HRTs to the k-factor, described earlier for conventional thickeners, gives up to 100 per cent increase in rating. It is not common, however, for k-factors to be used for high-rate torque selection.

Thickener costs

A number of factors are considered when selecting a thickener that have a bearing on cost rather than size. Two of the most important are:

1. the selection of either a high-rate or conventional thickener
2. thickener tank configuration.

High-rate versus conventional thickener

A HRT can only be used if the feed solids are flocculated. In some applications, the flocculating agent has a detrimental effect on downstream processes and the conventional thickener must be used in these cases. Other points of comparison include:

- high-rate thickeners
  - smaller diameter and smaller 'footprint'

    - lower capital cost
    - thickener control is possible due to short retention time
    - short retention time, beneficial where heat loss or secondary reactions are a problem
    - small inventory, which is beneficial if feed material has high value
    - a control system is generally required
- conventional thickeners
    - larger diameter
    - higher capital cost
    - large storage capacity
    - flocculant not always required
    - long retention times
    - slow response to changes in operating conditions
    - control systems are not required.

Thickener tanks

A wide range of tank types is used (Table 13.22).

TABLE 13.22
Comparison of thickener tanks.

Type	Comments
Above-ground or elevated tanks	Generally the most expensive, but access to underflow system is excellent
Conical-bottom tank compacted earth mound with underflow tunnel	Less expensive than above-ground if a cheap source of compactable fill is available
Flat-bottom tank at ground level with tunnel	Lower cost than above-ground tank; safety considerations
Flat-bottom tank at ground level with underflow lines buried and pumps at periphery of tank	Cheapest installed cost; risk of blocking underflow lines

Capital costs

Complete thickener capital costs for a free-standing tank are shown in Figure 13.13 for diameters from 2 to 35 m.

Figure 13.12 relates to an above-ground, mild steel tank and mechanism, with all wetted items epoxy painted, all non-wetted items painted with chlorinated rubber and rake lifting on 5 m and larger thickeners. The graph is used for both high-rate and conventional thickeners, with high-rate prices closer to the upper limit.

From the example used earlier in this section, the following comparison is drawn between high-rate and conventional thickeners (see Tables 13.23 and 13.24).

FILTERS

Different types of filters, and their capital and operating cost bases, are presented in this section.

Selection of filtering equipment

Filters are the most common form of dewatering device for the production of a solid or semi-solid product from a slurry. Filters are also used to recover liquor products, for solution clarification and for removal of trash from slurries. The type of filter most suited to the application usually depends on a combination of several of the following considerations:

- available floor space
- batch or continuous operation
- feed size distribution
- feed slurry density
- filter weight
- filtrate clarity
- other properties of the filter cake, such as cake adhesion to the filter cloth, compressibility, oxidation on exposure to air

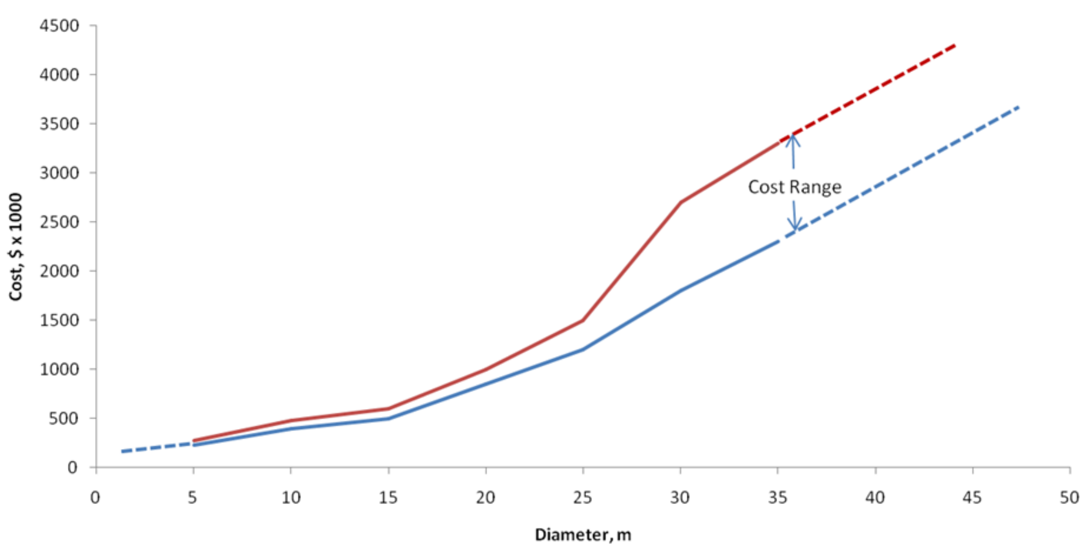


FIG 13.13 - Conventional and high-rate thickener capital cost.

TABLE 13.23

Comparison between high-rate and conventional thickeners.

	High rate	Conventional with floc	Conventional without floc
Feed rate (t/h)	80	80	80
Diameter (m)	12	40	60
Floc usage (g/t)	13	13	N/A
Installed cost of thickener	$900 000	$5 100 000	$8 000 000
Floc mixing system – installed cost	$200 000	$200 000	N/A
Annual cost of flocculant	$70 000	$70 000	N/A

N/A = not applicable.

TABLE 13.24

High-rate thickeners – typical separation rates and performance data.

Application	Feed (% solids)	Underflow (% solids)	Volumetric separation rate (m^3/m^2h)	Area loading rate (t/m^2d)
Alumina red mud washing	10 - 15	30 - 40	1.2 - 2.4	6.5
Blast furnace scrubber water	1 - 2	30 - 50	7 - 10	-
Borate tailings	5	35 - 45	2 - 5	5
Cement, wet process	10 - 15	40 - 50	5 - 7.5	13 - 33
Clean coal fines	1 - 7	25 - 40	6 - 10	3.5 - 1 6
Coal refuse	1 - 8	25 - 50	6 - 10	2 - 25
Copper concentrate	15 - 30	60 - 75	3.6 - 10	25 - 50
Copper, leach residue	5 - 15	45 - 55	2 - 7	6 - 33
Copper ore, ground	15 - 35	45 - 65	2 - 7	16 - 50
Copper tailings	10 - 35	50 - 65	1 - 10	16 - 50
Diamond tailings	10	45	7.5 - 10	20
Gold ore, cyanide leached	10 - 33	40 - 60	2 - 10	12 - 33
Kaolin slurry	1 - 6	20 - 30	6 - 7.5	-
Lead concentrate	10 - 40	60 - 80	7 - 10	50
Lime neutralised pond water (phosphates)				
Partially neutralised	2.7	30 - 40	5.9	4
Second stage neutralised	2.5	13 - 20	4	2.5
Magnesium hydroxide from brine	9	25 - 50	7.5	20
Magnesium hydroxide from sea water	2 - 6	10 - 25	2 - 10	1 - 10
Magnetite (heavy media)	12 - 18	60 - 70	7 - 10	30 - 50
Mine water clarification	-	-	2 - 10	-
Molybdenum tailings	20 - 30	50 - 70	5 - 10	30 - 50
Phosphate slimes	2 - 8	12 - 18	6 - 8	8 - 17
Potash (clay-salt slurry)	0.5 - 1.2	6 - 10	2	-
Sand tailings	2 - 10	30 - 40	8 - 12	-
Scrubber water	5 - 15	20 - 50	7 - 10	5 - 50
Silver ore, cyanide leached	10 - 33	40 - 60	-	-
Soda ash, primary feed	1 - 3	8 - 20	5 - 10	30 - 50

TABLE 13.24 CONT ...

Application	Feed (% solids)	Underflow (% solids)	Volumetric separation rate (m^3/m^2h)	Area loading rate (t/m^2d)
Taconite tailings	10 - 15	55 - 65	7 - 10	-
Tar sand tailings (cyclone overflow)	6 - 7	30 - 50	2 - 7	-
Uranium acid leached ore	10 - 30	45 - 65	5 - 8	16 - 50
Uranium precipitate (yellow cake)	1 - 10	15 - 40	1 - 7	1 - 10
Zinc concentrate	15 - 40	60 - 75	3 - 10	40 - 50
Zinc middlings	5 - 10	60 - 70	5 - 10	12 - 33
Zinc tailings	20 - 35	50 - 70	2 - 10	16 - 50
Zirconium (extraction liquor)	0.12	5 - 10	5	-

- moisture content required of the filter cake
- rheological properties of the feed (eg viscosity)
- slurry temperature
- need to wash the filter cake
- whether flocculants may be used
- whether some of the above are critical for further unit processing or contractual commitments
- whether the solid or the solution is a valuable product.

In selecting and costing the optimum filter/s, the complete solids-liquid separation circuit (including slurry storage, thickening, clarification, washing, product handling, storage and drying) is usually costed and optimised.

The selection of the lowest cost filter does not necessarily result in the overall lowest cost dewatering circuit. In some cases, the optimum circuit requires the use of two different filters. In other cases, a range of filters may be acceptable and selection is then made on other factors such as capital cost, operating cost and operating philosophy. A filter selection chart is shown as Figure 13.14.

Types of filters

There are four major types of filters commonly used in mineral processing and coal washing applications:

1. clarifying and screening filters
2. vacuum filters
3. pressure filters
4. belt presses.

With the exception of clarifying and screening filters, each filter type requires ancillary equipment that is selected and costed with the filter.

Ancillary equipment includes slurry thickeners and storage agitators, vacuum pumps and receivers for vacuum filters, high-pressure water pumps and/or compressed air for some pressure filters and flocculant mixing, storage and dosing equipment for belt presses.

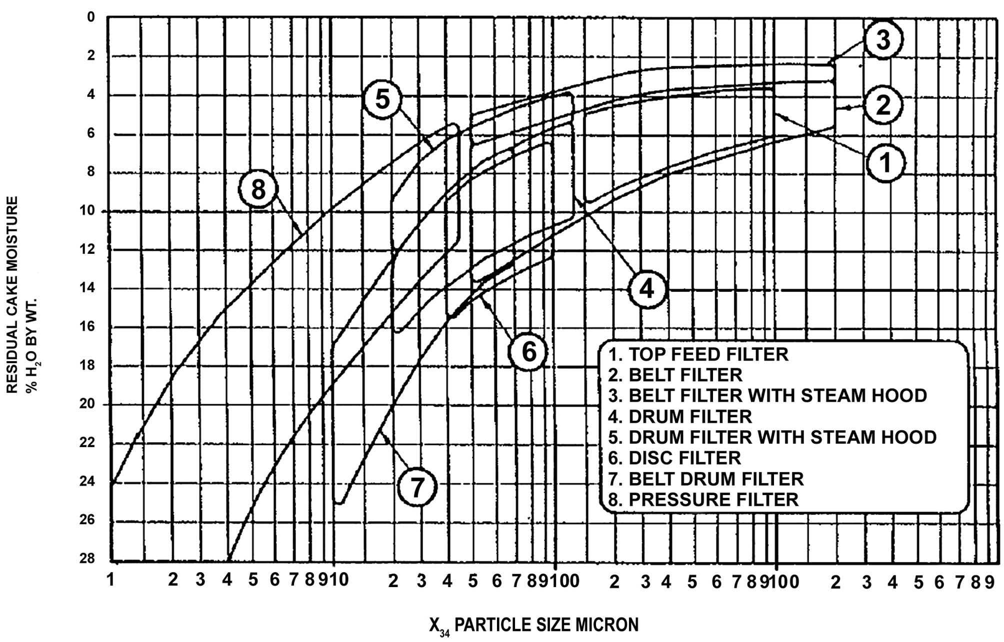


FIG 13.14 - Feed size ranges (µm) versus cake moisture content (per cent H_2O by weight at a solids sg of 3.5) for common filter types (source: Allis – Sala Basic S.23).

In addition, the selection (and cost) of feed pump(s) varies with the filter selection. Schematic filter arrangements for a vacuum drum filter and a horizontal pressure filter are shown in Figures 13.15 and 13.16, respectively.

The selection of the filter cloth is critical to most filtering applications. The importance of filtration properties, cake adhesion and mechanical strength of filter cloths varies depending on the type of filter selected (Kirk-Othmer, 1980; Bragg, 1983).

Calculation of required filter size is also critical for costing dewatering alternatives. The determination of the filter size requires filtration rates assessed by filter testing of a fresh, representative sample of the feed slurry. Vacuum filtration testing methods are simple, using standard laboratory apparatus that are well documented in the literature (Dorr-Oliver, 1980; Kirk-Othmer, 1980; Bragg, 1983). Pressure filtration testing facilities are now commercially available (Larox Pty Ltd, Allis Mineral Systems, Dorr-Oliver Pty Ltd, Ingersoll-Rand (Australia) Ltd and Clough Engineering Ltd); these suppliers can advise on optimising filtration cycles.

Commonly used filters are shown in Figures 13.15 to 13.29. These figures are accompanied by a brief description of their typical applications. There are many other filters available. Similar filters from different manufacturers may have considerably different operating parameters. Capital cost 'steps' between unit sizes also vary between manufacturers. For further

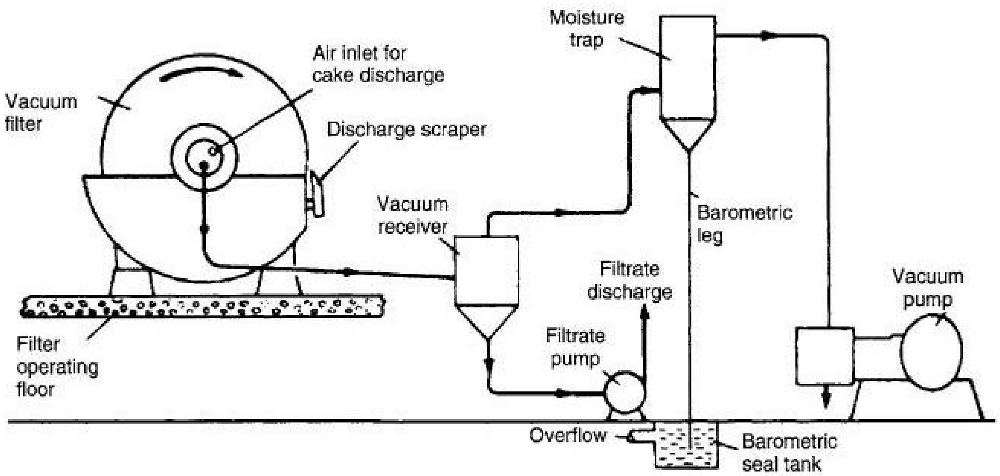


FIG 13.15 - Typical vacuum drum filter system (source: Wills, 1991).

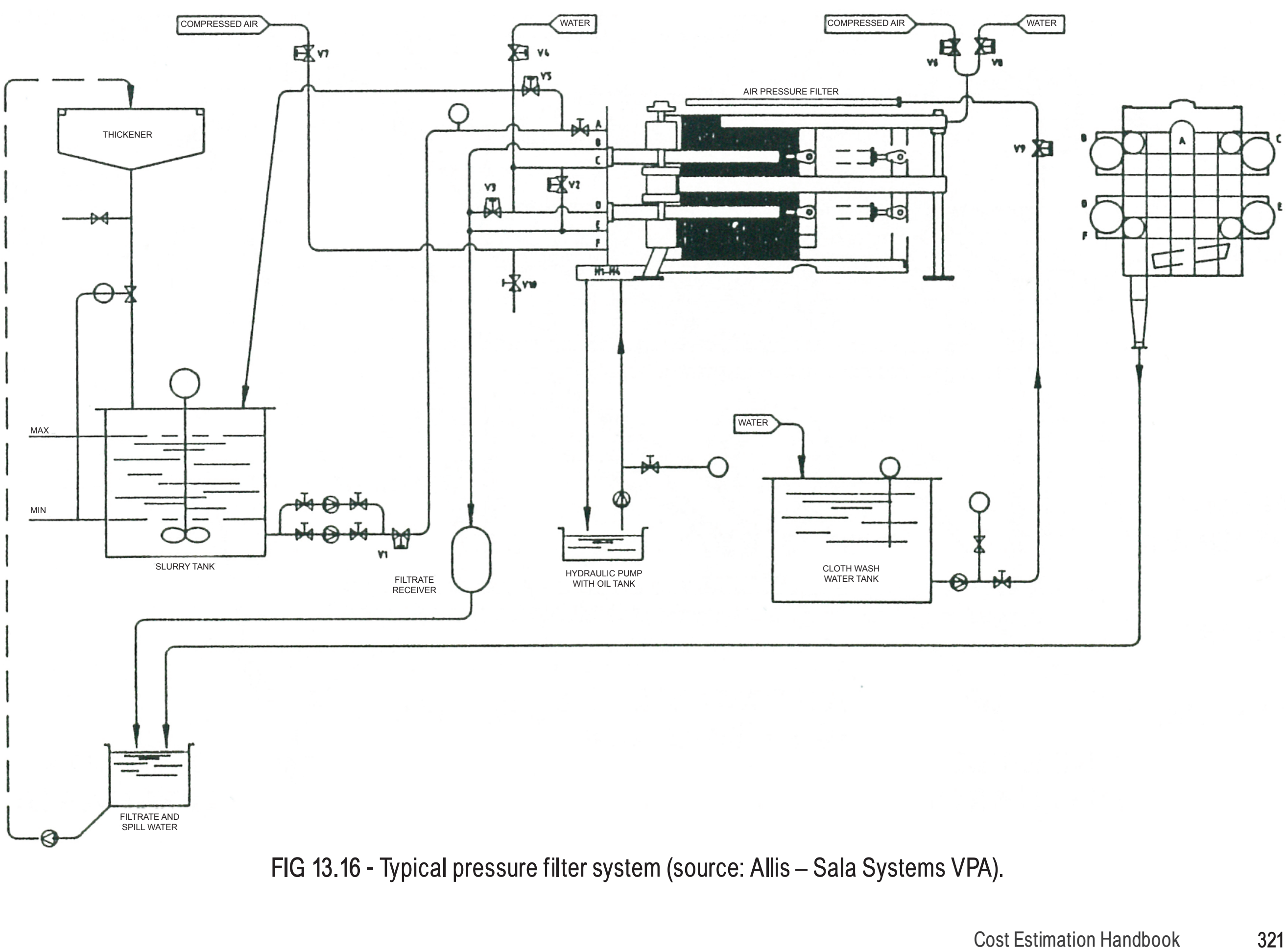


FIG 13.16 - Typical pressure filter system (source: Allis – Sala Systems VPA).

information on particular filters and their selection, readers are referred to the literature and manufacturers' catalogues and brochures in the reference list.

Clarifying and screening filters

Clarifying screens and filters include sand filters and linear belt screens.

Sand filters

Sand filters are normally used to 'polish' up to 200 ppm of solids from solution as a unit operation in a clarifying circuit. Solution is pumped through a bed of closely sized sand, which removes fine solids from the solution. The sand bed is maintained by back-flushing either at regular intervals or in automatic sand filters, such as that shown in Figure 13.17, when the back-pressure reaches a preset limit.

Linear belt screens

Linear belt screens, as shown for Figure 13.18, are typically used in the removal of trash, particularly wood fibre, from slurries. The slurry is distributed over a moving belt, typically a 0.5 mm square aperture through which the slurry gravitates. Wood pulp and oversize materials are washed before being discharged off the end of the belt.

Vacuum filters

Vacuum filters operate at atmospheric pressure on the inlet side and at a reduced pressure on the outlet side, limiting their effective filtering pressure to one atmosphere. They typically operate at around 0.6 bar (24" Hg). They are continuous and are readily accessible for inspection and maintenance. Maintenance costs vary, being least with drum filters and greatest with disc filters. It is essential that the vacuum be maintained as the installation is designed to avoid cracking of the cake. As a rule, compressible solids cannot be filtered on a vacuum filter. Feed, discharge and washing arrangements vary with the filter type.

Drum filters

The drum, covered with a suitable filter cloth, is partially submerged in an agitated bath of slurry. The drum is internally divided into longitudinal sectors. As the drum rotates and a sector is submerged, vacuum is applied to that sector and the filter cake forms. Vacuum continues to be applied to the sector as it

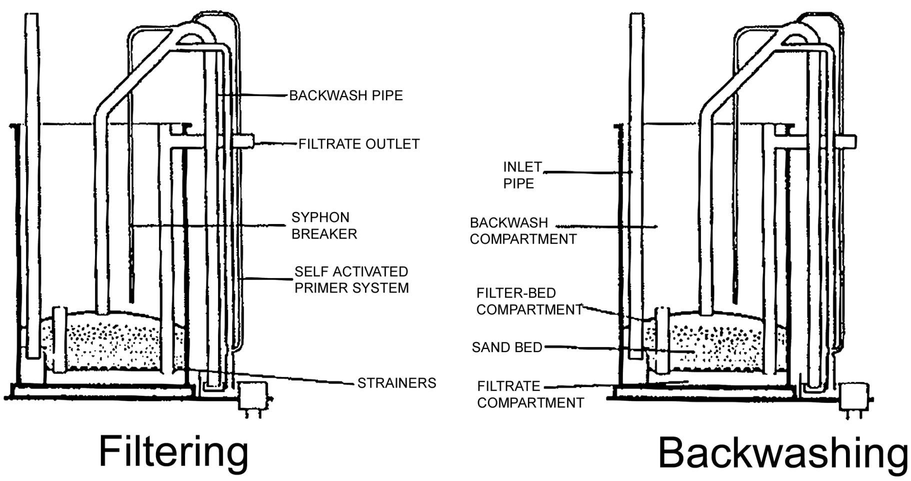


FIG 13.17 - Automatic sand filter (source: Delkor Pty Ltd).

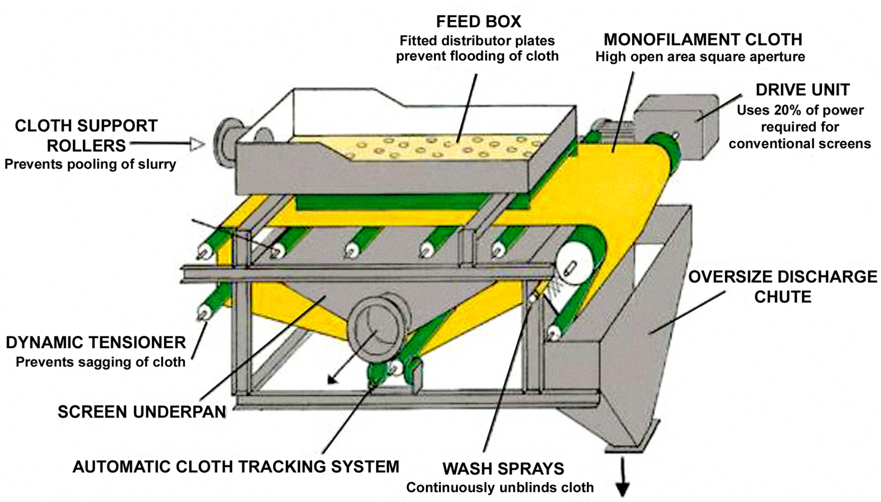


FIG 13.18 - Linear belt screen (source: Delkor Pty Ltd).

FIG 13.19 - Vacuum drum filter (source: Dorr-Oliver Pty Ltd).

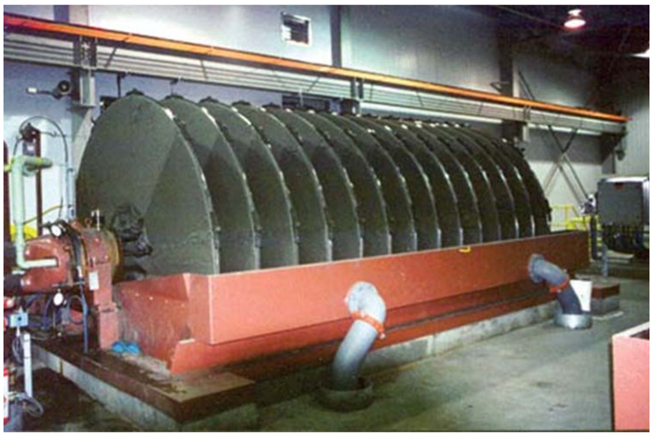


FIG 13.20 - Vacuum disc filter (source: Dorr-Oliver Pty Ltd).

emerges from the slurry bath and enters the drying and washing zones. As each sector approaches the discharge point, the vacuum is removed and the cake discharged. Discharge methods include air blowing, scraper discharge, belt discharge, string discharge and roll discharge.

Drum filters can also be top-fed. Washing is achieved by adding water to the partially dried cake. Cake drying can be assisted by passing the drum under a steam hood during the drying cycle. Where filtration of extremely fine solids is required, a precoat filter may be used. A 50 to 100 mm layer of diatomaceous earth or perlite precoat is built up on the drum and acts as the filter medium. Cake is discharged by scraping it, along with a small amount of the precoat from the drum surface. Pressure belt drum filters incorporate a pressure belt over the top of the drum to apply pressure and shear to the formed cake. Significantly lower cake moistures are obtained by this procedure.

Disc filters

Similar in principle to the drum filter, vertical discs pass through an agitated slurry tank. Each disc comprises several sectors, each containing a frame enclosed in a cloth bag. As the disc rotates, each sector undergoes vacuum or air blowing, depending on its position in the rotation cycle. Disc filters provide a high filtration area to floor space ratio. Washing is not practiced on disc filters and they are only suitable when relatively thick (>12 mm) cakes are formed.

Horizontal belt filters

A horizontal belt filter (HBF) is gravity fed, allowing coarser feed particles to gravitate to the belt. Vacuum is usually applied through one or more vacuum boxes over which the grooved elastomer belt travels. The filter cloth is carried on this belt. Wash water can be applied at one or more locations along the belt. Discharge is

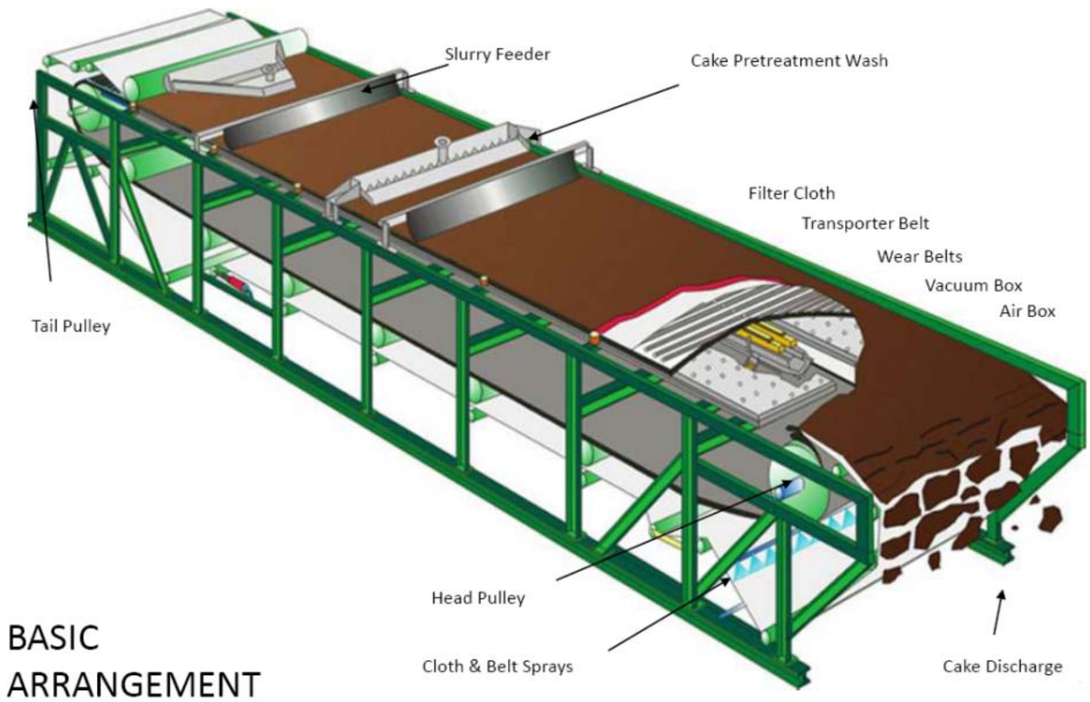


FIG 13.21 - Horizontal belt filter (source: Delkor Pty Ltd).

normally by belt discharge, although roller discharge can be incorporated for sticky cakes. Variations on the HBF include the rigid belt filter, travelling vacuum box filters and the tilting pan filter.

Pan filters

Pan fillers are specialised circular horizontal vacuum filters used where segregation of the mother liquor from various strength wash liquors is important, or where large volumes of liquid or wash liquor need to be separated from readily draining slurries. Cake is discharged by a scroll, leaving a heel of cake on the filter cloth. In some cases, this heel can be at a significantly higher moisture level than the discharged cake. As new feed spreads over the cloth, an air blow mixes the heel with the incoming feed, simultaneously cleaning the cloth.

FIG 13.22 - Pan filter (source: Dorr-Oliver Pty Ltd).

FIG 13.23 - Ceremec disc filter (source: Larox Pty Ltd).

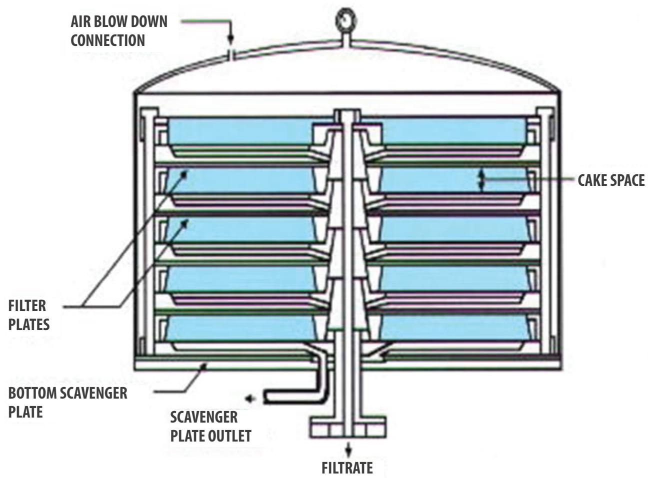


FIG 13.24 - Horizontal plate pressure filter (source: Diatomite Qld).

Pressure filters

Pressure filters are naturally batch units, although many now operate in semi-continuous mode. Several types are available, mostly based on the plate-and-frame principle, although some still use the pressure vessel principle. Plate-and-frame type filters consist of horizontal or vertical plates and are manually or automatically operated, with or without diaphragm compression and air blowing to further reduce cake moisture. Recent advances in pressure vessel type filters include the hyperbaric disc filter (Andritz) and the positive pressure Ceremec disc filter (Superflo Technologies Pty Ltd). Older types still find application in particular areas.

The availability of high-pressure diaphragms in tube presses and some recessed plate filters make them ideal for filtering compressible solids.

Horizontal plate pressure filters

Modern horizontal plate pressure filters (Figure 13.24) are generally automatically operated by a programmable logic controller (PLC) to give semi-continuous operation.

Typically, such filters are programmed to close the filter, form a precoat on the filter cloth, charge the filter, use membrane filtration to ensure an even cake, wash the cake with separation of the wash water, air blow the cake, open the filter, discharge the cake and wash the filter cloth. Opening of the filter is achieved by mechanical or hydraulic movement of the plates or by hydraulically expanding flexible seals between fixed plates.

Vertical plate pressure filters

Large automatic filters and small manually operated batch filters both find considerable application (Figure 13.25). The former are generally similar in operation to the horizontal pressure filters described above. They are available in larger sizes (although their cake area is half the filter cloth area) and have quicker discharge and closing times, but they take up greater floor space.

Small plate-and-frame filters (Figure 13.26) are generally manually operated batch operations. They have a relatively low capital cost and enable very high pressures to be used for the recovery of fine materials, often from dilute feed slurries.

Kelly filters

Retractable-shell pressure filters are typically used for clarification of hot, highly scaling liquors such as occur in the alumina industry. Suspended solids are filtered from hot caustic liquor at 100 to 105°C through precoated vertical leaves (Figure 13.27).

Red mud thickener overflow can also be clarified in Kelly filters. Filter cake is removed by emptying the filter of feed, retracting the leaves out of the filter and hosing off the cake with a high-pressure water jet. In other applications, water washing and steam drying of the cake are incorporated before discharge.

Tube presses

The tube press (Figure 13.28) is a PLC-controlled, semi-continuous bank of membrane filter presses operating at very high pressures of up to 140 bar. Originally developed for filtration of china clays, tube presses are ideal for dewatering fine materials, especially when low cake moistures are required. Under the high pressure applied, the resistance of the filter cloth is small compared with the resistance of the filter cake.

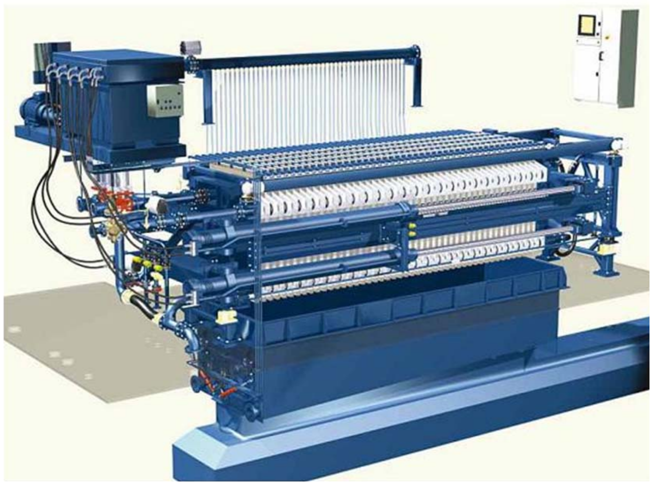


FIG 13.25 - Vertical plate pressure filter (source: Metso Minerals).

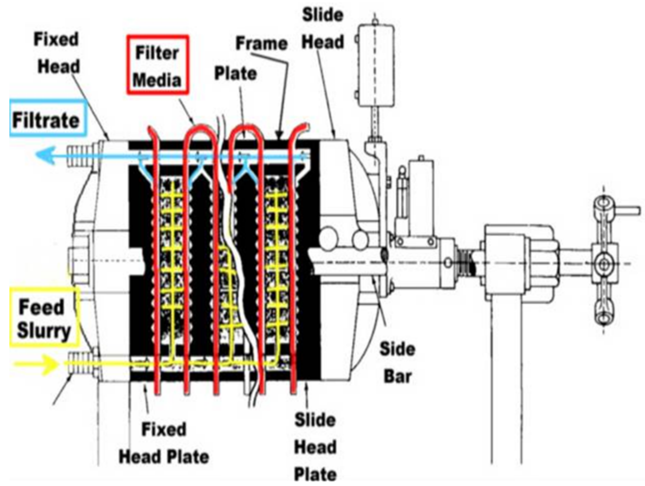


FIG 13.26 - Plate-and-frame filter (source: mine-engineer.com).

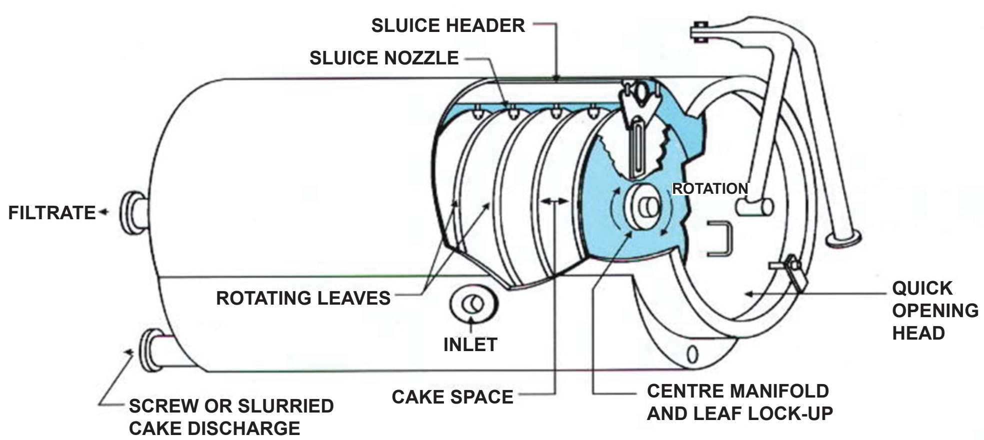


FIG 13.27 - Kelly Filter (source: Diatomite Queensland).

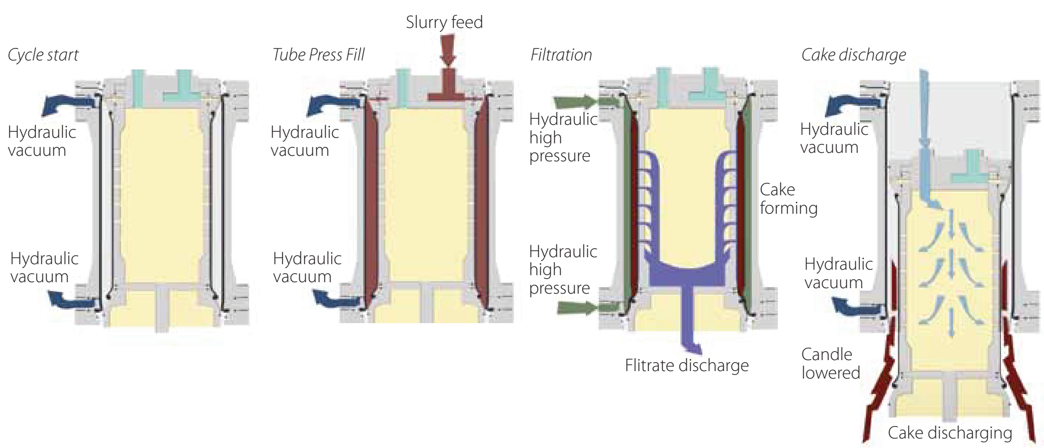


FIG 13.28 - Tube filter operating cycle (source: Metso Minerals).

This allows a tightly woven filter cloth of very fine pore size to be used. Cake washing and air blowing can be incorporated in the filtration cycle.

Belt filters

Ideal for readily flocculated feeds, especially of low solids content, band press filters (Figure 13.29) require an initial flocculating stage.

Flocculated feed passes through a decantation section before the drained pulp is fed between two semi-parallel filter belts. The belts pass around a series of successively smaller diameter rollers. With the belts under tension, pressure on the cake steadily increases. More importantly, as the belts pass around each roller, shear is created within the cake, further releasing residual moisture from the cake. This shear results from the two belts moving at the same speed, but with the outer belt moving on a longer radius than the inner belt.

Band-press filters are common in coal and waste treatment applications. Their application in minerals dewatering is likely to increase because of their low total dewatering installation cost. As with HRTs (discussed previously), the high cost of flocculants may be low in comparison with the capital cost of a more complex filtering installation.

Capital cost

In Figure 13.30 the capital costs for common filters types are presented on a cost per square metre of filtration area basis (band press filters on a cost per metre width

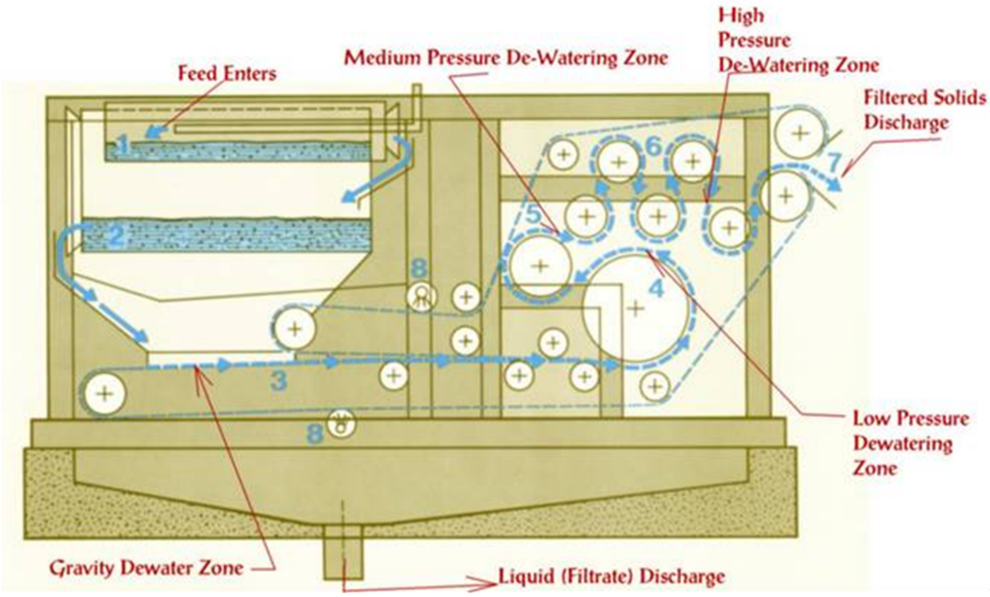


FIG 13.29 - Band press filter (source: Warman International Ltd).

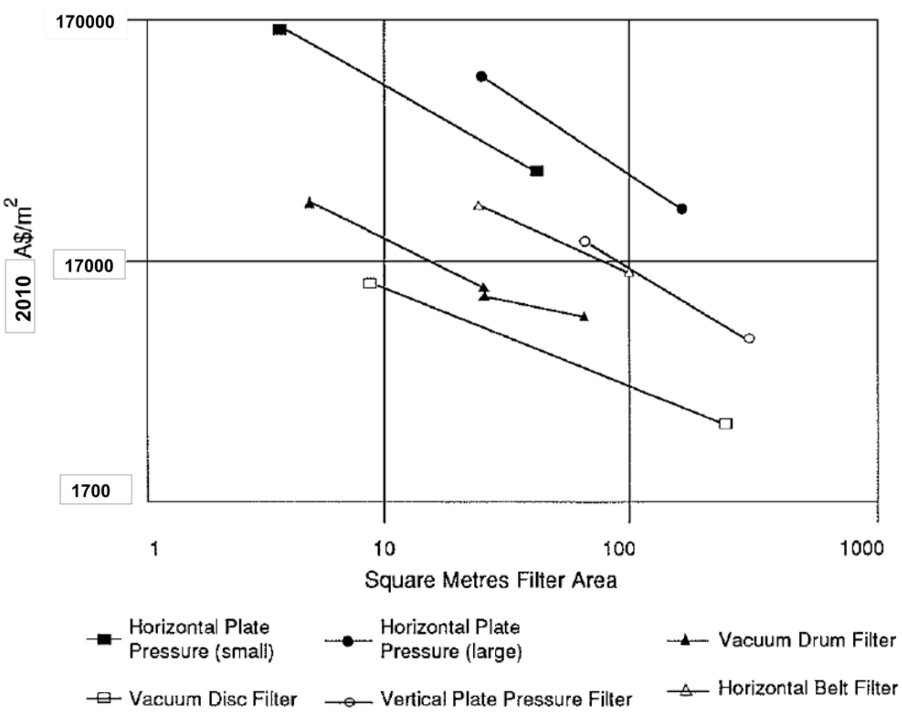


FIG 13.30 - Indicative capital costs of common filter types.

basis). It is beyond the scope of this handbook to list filtration rates, as these are specific to the filter feed, the selected filter and other operating parameters. Cost per tonne of feed filtered is estimated after simple but appropriate test work to establish such specific filtration rates.

Filtration area is the total surface area of useable filter media. Filtration rate takes into account the effective filtering area and effective time for which that area is used. Thus, the filtration rate uses filter cycle times, wash times, air blow times and belt washing, etc and adjusts for filter 'dead time'.

Filtration rate also adjusts for single- or double-sided filtration in which double-sided filter chambers may have twice the surface area but half the filtration rate per square metre.

With an increasing trend towards modular design, particularly in filters such as pressure filters, horizontal belt fillers and belt presses, step-changes in capital costs occur. As step sizes are brand-specific they are not indicated on the graphs, but are generally clearly defined in manufacturer's literature. Cost per square metre is usually cheapest at the largest sized unit before such a step-change.

The costs shown in Figure 13.30 are for basic filters of generally mild steel construction. Formulas for the calculation of these capital costs for filters are given in Table 13.25. They do not take into account peripherals, installation or filter buildings. Some rule-of-thumb factors for use with these figures are given in Table 13.26.

These costs are suitable only for first approximation cost estimation. Their accuracy is not sufficient to enable accurate selection of one filter type over another. Selection of filter type should be made only after the filtration rates have been established by test work for each filter type under consideration.

Many filters are fully imported into Australia. Filter prices are volatile and vary widely depending on their country of origin and on ruling exchange rates. In estimating filter costs, current budget pricing and applicable exchange rates should be obtained from filter suppliers.

Operating cost

General beneficiation operating costs are provided in Chapter 12 – Beneficiation – Concentration. Where a range of filters is acceptable for a specific duty, operating cost and operating philosophy affect equipment selection, and hence capital cost. Selection of a higher capital or operating cost filter can, in some cases, eliminate slurry thickening or other unit processes.

Operating costs are usually site-specific and vary depending on the feed sizing, feed moisture, use of flocculants, washing requirements, dryness of the filter cake and other local costs.

In general, filter operating costs increase with fineness of the feed sizing and the required degree of dryness of the cake. Typical operating cost ranges are:

Filter	**Cost ranges**
Vacuum filters	\$0.50 - \$l.50/t solids
Large (>44 m^2) pressure filters	\$1 - 2/t solids
Small (<38 m^2) pressure filters	\$2 - 3/t solids

The life and cost of the filter cloth is often a large proportion of filter operating cost, followed by diaphragms, power and labour costs. When long air blowing cycles or flocculants are required, these can add substantially to the operating cost. Selection of a

TABLE 13.25
Mathematical relationships for the expression $y = ax^b$, used in Figure 13.30.

Filter type	a	b	Applicable x range (m^2)
Horizontal plate pressure filter (small)	200 270	-0.64631	3.2 - 38
Horizontal plate pressure filter (large)	254 138	-0.60093	22 - 144
Drum filter (small)	40 301	-0.57601	3.8 - 24
Drum filter (large)	9432	-0.09645	24 - 55
Disc filter	21 740	-0.45355	8.25 - 240
Vertical plate pressure filter	68 397	-0.43612	65 - 305
Horizontal belt filter	43 841	-0.34275	22 - 100
Linear screens	27 858	-0.39890	1.2 - 8.2

Notes:

1. Refer to text for comments regarding pricing, especially the influence of exchange rates.
2. Ceremec disc filters are advised to be of similar price to horizontal plate pressure filters.
3. Linear screens not plotted.
4. Belt press filters are costed on a belt width basis; sand filters are costed on a diameter basis. Neither is plotted.
5. Capital costs of other filters are not provided.

TABLE 13.26
Typical base filter supply and rule-of-thumb add-on costs.

Filter type	Included in base cost	Peripherals	Notes
Vacuum drum	Drum, agitated vat, motors, cloth	Feed pump, vacuum pump, filtrate receiver and filtrate pump; add 60%	For 316 grade stainless steel or rubber coated wetted parts add 20%
Vacuum disc	Drum, agitated vat, motors, cloth	Feed pump, vacuum pump, filtrate receiver and filtrate pump; add 60%	–
Horizontal belt	Filter, rubber belt, cloth, motor	Feed pump, vacuum pump, filtrate receiver and filtrate pump	For 316 grade stainless steel or rubber coated wetted parts add 15%
Ceremec	Ceramic disc assembly, agitated vat, motors, PLC	Feed pump, vacuum pump, filtrate receiver and filtrate pump	316 grade stainless steel wetted parts, ST52-3N frame standard
Vertical plate pressure	Filter, cloth, polypropylene plates and frames, power unit, PLC. For 316 grade stainless steel, wetted parts add 9 - 14%	Feed pump, filter cloth wash pump, air compressor, valves; add 15 - 20%	Installed filter including peripherals (but not building) add 35 - 45%
Horizontal plate pressure	Filter, cloth, stainless steel plates and frames, power unit, PLC	Feed pump, filter cloth wash pump, air compressor, valves; add 15 - 20%	Installed filter including peripherals (but not building) add 35 - 45%
Tube presses	Filter, cloth, diaphragm, candle, PLC, hydraulic pump	Feed pump, air compressor	–
Band presses	Mixing tank, mixer, motor, two belts	Feed pump, flocculant mixing and storage	–

specific filter type should consider these operating cost aspects in conjunction with the capital cost estimation.

ACKNOWLEDGEMENTS

The help and suggestions of the many colleagues who have assisted in the preparation of this section are gratefully acknowledged. As Arthur F Taggart wrote in 1944:

No one who has tried to write a technical handbook, or a part thereof, realises how little he or any other one person knows about the subject that he considers his specialty.

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